DEVELOPMENT OF A TECHNIQUE FOR PERFORMANCE EVALUATION OF INDUSTRIAL CONTROLLERS

M. Rossi [1], **C. Scali** [1*], **M. Amadei** [2]
[1] *Department of Chemical Engineering - CPCLab - University of Pisa (Italy)*
[2] *Research and technology -- Polimeri Europa - Milano (Italy)*

Abstract: The paper describes a method to account for different issues of performance monitoring of industrial control systems, under SISO control: detection of poorly tuned loops, process identification, controllers retuning and evaluation of performance improvements. The procedure can be completely automated and applied on-line or off-line; it starts from the analysis of plants data and ends up with a suggestion to the operator about the new controller settings. Characteristics and effectiveness of the technique are illustrated by simulations results and application to industrial plant data.

Keywords: Performance Monitoring, Identification, Controller Tuning, Industrial Control Systems

1. INTRODUCTION

Performance monitoring in industrial plants is an aspect of increasing importance nowadays, as witnessed by large efforts in advanced academic research and in plants applications. Several aspects still must be resolved in a systematic way: both theoretical (metrics to be used to evaluate control performance, extension to MIMO systems) and practical (automated application on industrial plants, minimisation of perturbation on plant operation, interaction with operators). An updated overview of these topics can be found in Thornhill and Seborg (2002).

A control loop can perform poorly for several reasons, as valve stiction, sensors failures, incorrect tuning of controllers. In the perspective of developing a global tool accounting for different causes, the issue of performance evaluation of loops controlled by PI(D) regulators is addressed in this paper.
This aspect has large importance, owing to their large diffusion in industrial plants; these controllers are not tuned at their best (Ender, 1993), because of the general tendency to avoid oscillations (synonym of instability) and therefore to apply conservative tuning, which results in slow responses. This is also a consequence of the fact that standards about the procedure are not strictly established on the plant and tuning is left to operators' skill; very often, the company policy, while asking for best use of available technologies, assigns few resources to this important task.

* e-mail: scali@ing.unipi.it

In addition, even a perfectly tuned controller may become inadequate when the process undergoes large variations, due to changes in operating condition or to the inlet of external perturbations. Process changes can be detected by performing periodic identification (step and relay tests can be used for this purpose), followed by a new design / tuning of controllers (Åström and Hägglund., 1995).
Even though these procedures can be completely automated (leading to adaptive control / autotuning), there are some drawbacks, because:

- explicit perturbations must be introduced in the plant (even though the amplitude can be somehow controlled),
- may become time consuming, owing to slow process dynamics or reiteration of experiments (Yu' 2000, Marchetti and Scali, 2000).

Without the need of introducing any additional perturbation, all the required information about the behaviour of the controlled process, can be extracted from plant data, which are continuously recorded and archived in real-time data-bases, then available for further analysis.
Several different performance index have been developed for specific purposes; among the most known and widely used: Harris index (Harris, 1989), which allows to compare actual controller with optimal (minimum variance) controller; Idle index (Hägglund, 1999), which allows to detect sluggish responses or the Oscillation Detection technique (Hägglund, 1995), which allows to detect too oscillating responses. These indexes (or their modifications) can be applied for on-line or off-line analysis; the developments of related software tools, as well as their implementation (DCS or external

computers) is still under investigation (Hägglund, 2002).

It is highly desirable that the performance analysis procedure, with indication of poorly acting loops, is associated with a new design/tuning of controllers and evaluation of achievable improvements. Therefore the development of software tools is called for, able to assist the operator in taking key decisions, with possible automated applications on the plant, after a suitable time of training on plant data. With this short introduction to the problem, as first this paper will briefly review basic aspects of two performance indexes to detect poor behaviour of the controller. Then, the different steps of the procedure, which goes through the steps of performance monitoring, process identification, controller tuning and improvement evaluation, will be illustrated. Finally, the effectiveness of the technique is illustrated by simulations results and application to real plant data, drawing some conclusions and indications for future work.

2. INDEXES TO MONITOR PLANT PERFORMANCE

In the reference scheme of a SISO control loop (Figure 1), *P* and *Pd* indicate the effect of manipulated (*u*) and disturbance (*d*) on the controlled (*y*) variable, effects which can be different in the general case.

The application of performance indexes is able to evaluate from the analysis of plant data when a controller gives too weak action, (slow closed loop response), or too strong action (oscillating response). The computation of these indexes should be as simple as possible, requiring few information, (for instance based only on values of controlled and manipulated variables), and giving rise to a clear classification of controller behaviour.
Slow responses can be detected by means of the Idle Index (Hägglund, 1999). The computation of this index is based on two characteristic times: t_{pos}, time interval when the product of the two gradients of controlled (CV) and manipulated (MV) is positive, and t_{neg}, (negative product of gradients).
Then it is possible to get a normalised index in the range [-1÷1], which for slow disturbance suppression approaches 1, while for faster responses assumes negative values:

$$Ii = \frac{t_{pos} - t_{neg}}{t_{pos} + t_{neg}} \qquad (1)$$

Too oscillating responses can be put into evidence by the oscillation detection technique (OD), presented by (Hägglund, 1995). The analysis can be split into two parts: detection of significant perturbations (*anomalies*) and detection of persistent oscillations.

Table 1: Classification of the perturbance response

OD	Ii ∈ [-1; -0.7]	Ii ∈ [-0.7; -0.4]	Ii ∈ [-0.4; 0.4]	Ii ∈ [0.4; 0.7]	Ii ∈ [0.7; 1]
1÷5	Good	Good	Good	Poor	Bad
6÷10	Accept.	Accept.	Poor	Poor	Bad
>10	Bad	Bad	Bad	Bad	Bad

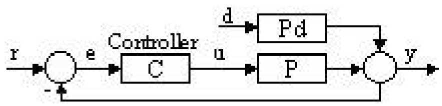

Fig.1: The reference scheme

Each single oscillation is characterised by its IAE:

$$IAE = \int_{t_{i-1}}^{t_i} |e(t)|\,dt \qquad (2)$$

where t_{i-1} e t_i are successive times where e(t)=0. To be significant, the value must be above the IAE of a half-period of a sinusoidal oscillation having a defined amplitude and frequency (amplitude equal to 1% of control range and frequency equal to the estimated critical frequency of the process). To detect the presence of a persistent oscillation, it is necessary to detect a significant number of oscillations n_{lim} over a supervision time T_{sup}. In this case the reference value is n_{lim} 10 in the time window T_{sup}.

By adopting the Ii and OD indexes it is possible to achieve a quantitative evaluation of the closed loop response to a perturbation, as reported in Table 1.

3. THE PROPOSED TECHNIQUE

As anticipated in the introduction, the proposed technique and the associated software tool developed for its implementation, has the objectives of performing different tasks of:

1) performance monitoring, with detection of "*anomalies*" (poorly performing loops),
2) identification of process and disturbance dynamics (Fig. 1)
3) controller tuning according to a desired performance criterion,
4) evaluation of expected improvement with the adoption of the new controller.

This architecture is depicted in Figure 2.

In the sequel these aspects will be fully illustrated, putting into evidence also implementation issues (DCS or External Computer, on-line or off-line) and interaction with the operator. About this point, while the final goal is a complete automated system (able to work by default from plant data to final tuning), interaction with the operator can always be introduced to force (improve) computer operations on the basis of specific experience on the plant.

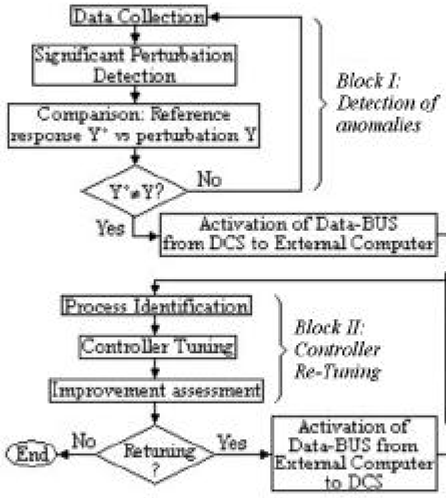

Fig. 2. The architecture of the proposed tool

3.1. *Performance Monitoring*

Plant data are acquired by the DCS system at given sampling times according to the usual procedure; they consist in values of controlled (CV) and manipulated variables (MV), set points, controller parameters, information about manual/automatic operation. In addition to routine elaboration (Figure 2, Block I):

a) data are analysed according to the OD technique and the presence of a "significant perturbation" in the controlled variable (Y) is detected,
b) this perturbation is compared with a reference response (Y°); if noticeable variation are found, the remaining part of the procedure is activated.

To be noted that:

- the presence of an "*anomaly*" does not mean necessarily that it is possible to get improvements: therefore it is compared with the last good response (Y°), which must be recorded in the system;
- both computations to detect the "*anomaly*", and data about Y°, do not require heavy additional load (computation and memory); they can be performed in the DCS, in agreement with Hägglund, (2002).

3.2. *Process Identification*

Once an "*anomalous response*" (which can be improved) has been detected by the DCS, the procedure of process identification is activated in the external computer, which receives all the necessary data (First step of block II, Figure 2). All elaboration of data for this and following stages can be accomplished by using standard Matlab and Simulink routines.
The output response of a controlled loop to a disturbance can be expressed as (Figure 1):

$$Y(d) = \frac{Pd}{1+PC} d \qquad (3)$$

The controller C is known, the disturbance d is assumed as unitary step; for P and Pd, models represented by Second Order Plus Time Delay (SOPTD) transfer functions have been assumed:

$$\tilde{P}(s) = \frac{K_p}{\tau_p^2 s^2 + 2\tau_p \xi_p s + 1} \cdot e^{-\theta_p s} \qquad (4)$$

$$\tilde{P}d(s) = \frac{K_d}{\tau_d^2 s^2 + 2\tau_d \xi_d s + 1} \cdot e^{-\theta_d s} \qquad (5)$$

Both $\tilde{P}$ and $\tilde{P}d$ are identified, as the output response for a given controller may become suboptimal when P or Pd change. The values of parameters are determined by minimising the sum of mean square error (SMSE) between plant (Yi) and model (i) response (N is the number of data):

$$SMSE = \frac{1}{N} \cdot \sqrt{\sum_{i=1}^{N} \left[Y_i - \tilde{Y}_i\right]^2} \qquad (6)$$

Minimisation is carried out by a modified Simplex method (Nelder and Mead, 1965), based on the Matlab function "fminsearch.m". The simplex method may fail to converge or may converge to a suboptimal solution, owing to the presence of local minima, especially when a large number of parameters are present.

About the number of parameters, it must be noted that K_d and θ_d are not real unknowns, as they influence only the amplitude of output response (and not the shape), and the time when the perturbation shows up: unknowns can be reduced to 6.

About the problem of convergence, the initial guess on the values of parameters is very important and then initial knowledge on the process plays an important role. Starting from the values of previous models (memorised until new models are computed), can be a good choice in the case of not too large variations in process parameters. More in general, an estimate of initial values for the parameters K_p and θ_p, (and from them τ_p, τ_d, ξ_p, ξ_d), can be given on the basis of CV and MV values; (details in Rossi 2002).

The number of points and characteristic times of the response come from the DCS; among them: time of inlet of a perturbation (t_{in}), time of detection of a perturbation (t_{det}), time to reach a new steady state (t_{ss}). This is the system default: the operator can decide to start the retuning procedure before t_{ss}, acting with a new controller to suppress the disturbance and this will bring advantages for long lasting perturbations (details in Figure 4).

3.3. Retuning and Evaluation of Performance Improvement

Controller retuning is performed assuming an ITAE objective; for PI controllers:

$$\min_{C}\left(ITAE\right)=\min_{Kc,ti}\left(\int\left|e(t)\right|\cdot t\cdot dt\right) \qquad (7)$$

The ITAE choice can be desirable in industrial process control owing to its characteristics of reducing tails in output response; (other objective functions can be adopted in the tool, if the case). To evaluate the effectiveness of a retuning, in terms of achievable performance, the auxiliary index RITAE has been defined as:

$$RITAE=\frac{ITAE_{new}-ITAE_{\min}}{ITAE_{old}-ITAE_{\min}} \qquad (8)$$

where: *new/old* stand for controller after/before retuning, *min* stands for ideal controller (minimum error). Values of RITAE: are strictly positive; in the range 0÷1, indicate improvements of performance; larger advantages are expected for values closer to 0. An illustrative example is reported in Figure 3 and in Table 2.

Figure 3 and table 2 also summarise information communicated to the operator at the end of the global procedure: old and new time responses and controller settings, minimum error response, values of the performance index (Ii, OD, RITAE). One additional information regards values of the actual model compared with old one and allows to distinguish performance deterioration due to change in the process or in the perturbation dynamics. At this point the operator has all the necessary information; in off-line operation: to evaluate (and grade) controller behaviour; in on-line operation to decide if it is worth to change controller parameters, something that can be accomplished manually or automatically, after operator consent. As final objective, once the procedure has been fully tested on plant data, changes in controllers parameters will be accomplished directly, with the supervising operator having the option of interrupting the automatic operation in every moment.

4. SIMULATION RESULTS

The technique has been tested by simulation on processes (P) and perturbations (Pd), represented by different transfer functions of First, Second, Higher order, Plus Time Delay. By varying values of parameters in a large range, a wide class of dynamics of possible interest in industrial applications have been analysed.
The key step of the procedure is the identification. The goodness of identification has been evaluated by

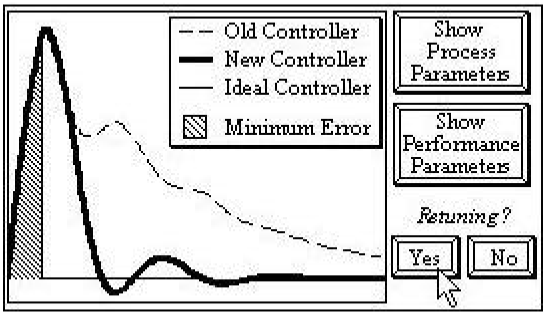

Fig. 3: Examples of output responses, with old, new and ideal controller (minimum error)

Table 2: Performance Parameters

C	Kc	τi	Ii	OD	RITAE
Old	0.58	17.5	+0.48	1	0.15
New	0.43	4.37	-0.11	3	

comparing closed loop responses obtainable by PI controllers based on the models ($\tilde{P},\tilde{P}d$) and on a perfect knowledge of the process (P, P_d). The visual comparison of responses and an analytical index (DITAE, defined as (percentage) Difference of ITAE), show that identification is able to capture the essential dynamics for control purposes (details in Rossi, 2002).

Some general results can be pointed out:

- The technique based on the modified simplex method showed very good convergence properties.
- Best identification (time, accuracy) is obtained when process dynamics are the same of model (SOPTD), but a good fitting is obtained for a wide class of processes.
- Also, delay dominant dynamics ($\theta/\tau \gg 1$), as well as underdamped responses, are *easier* to identify, in terms of duration and accuracy.
- Only in the case of very high order (overdamped) disturbance dynamics, the technique may fail and this is communicated to the operator, who can change some default settings.

An example of typical simulation results is reported in Figure 4. Perturbation 1 and 2 can not be improved (Y=Y°); perturbation 3 (a decrease in the feed flow rate), causes a more oscillating response which is detected by a comparison between Y and Y° (the ratio in this case). By default the system waits until new steady state conditions are reached and then identifies changes, retunes the controller and, for subsequent perturbations, is able to improve performance according to response 4.
The operator can decide to act before the new steady state has been reached (*Start elaboration*); this way, after the very short elaboration time and controller retuning (*End elaboration*), it is possible to act on

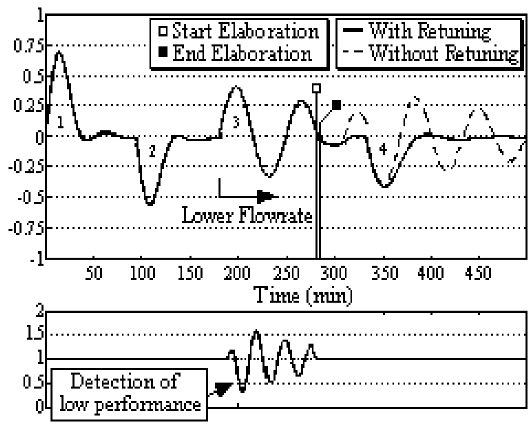

Figure 4: Example of typical simulation results

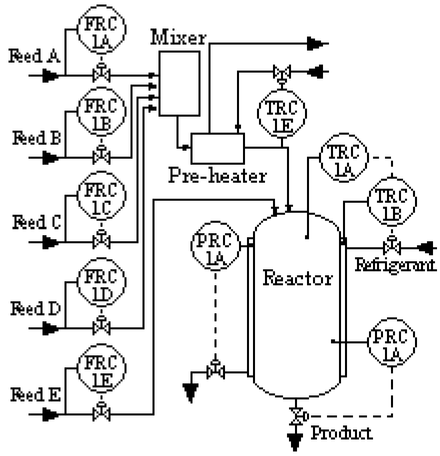

Fig. 5: Plant scheme

the perturbation, suppressing last oscillations. It is evident that the elaboration of data can start only when the response dynamics is sufficiently developed, in order to allow a correct identification; in this stage a crucial role is played by operator experience; anyway the system can work by default, without any assistance.

The robustness of the technique to the noise on plant data has been investigated. The effect of noise has been taken into account by adding random errors to clean data; internal parameters to model the effect of noise have been selected in order to make it as similar as possible to the noise present on industrial data (see section 5).
The ratio between noise and signal amplitude (N/S) has been changed in order to analyse the deterioration of results and evaluate a maximum allowable amplitude.
The presence of noise on plant data influences the stages of detection of a significant *anomaly*, through the comparison between Y and Y°, and the stage of identification.
Errors due to noise are reflected in achievable closed loop performance by the regulator based on the model. A systematic evaluation of performance deterioration has been accomplished for different process dynamics by means of the DITAE index. Assuming a maximum DITAE error equal to 10%, for acceptable performance, a ratio N/S=30% is allowed; then the technique can be considered sufficiently robust for applications on industrial data.

5. APPLICATION TO PLANT DATA

The method has been tested for an off-line application on industrial data, kindly made available by Polimeri Europa s.p.a., from a polybutadiene plant for the production of SBR.

The plant section under analysis consists of a mixer of reagents, a pre-heater and the reactor cooled by an external jacket. Several sources of perturbation can affect the polymerisation reactor: changes in reagents/product properties, fouling of heat exchange surfaces, different stages of the batch process and changes in environmental conditions.

Pre-heating is necessary to make easier to maintain the reaction temperature at the optimal value. This variable is the most critical to control both for safety and high quality control: only 3 °C of deviation from set point values are allowed: larger values will cause the activation of alarms and eventually shut down of the plant. Several sources of perturbation can affect polymerisation: changes in reagents/product properties and environmental conditions, fouling of heat exchange surfaces, different stages of the batch process.

For this reason, pre-heater outlet and reactor temperatures are controlled by one PID and two cascaded PI controllers, respectively, while all other loops are controlled by simple PI controllers (Fig.5):

The evaluation of performance has been carried out following the logical scheme reported in Figure 2; the technique has been extended to cascade control loops, without substantial modifications (Rossi, 2002). When the perturbation enters in the inner loop, the method is applied as such; when in the outer loop, the process dynamics is changed to include also the inner loop (process plus controller).

The following data were available (Excel format):

- Controlled variable: Y(t)
- Manipulated variable: u(t)
- Set-point value: SP
- Controller parameters k_c, τ_i, τ_d
- Controller status (Manual/Automatic)
- Valve opening (%)

Data refer to 5 flow, 2 pressure, 3 temperature (one cascade) control loops. From a preliminary analysis, the reaction can been indicated as the plant section more perturbed by external disturbances. In 24 hours of operation 3 significant *anomalies* have been

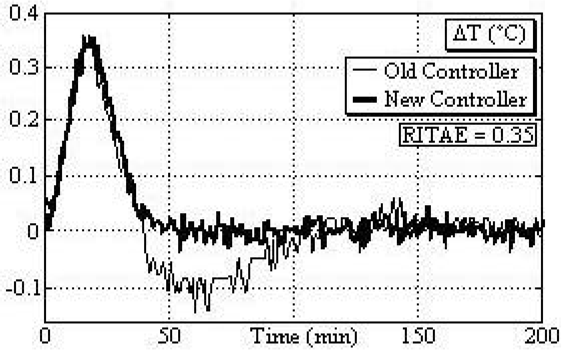

Fig. 6: Suppression of disturbance #1

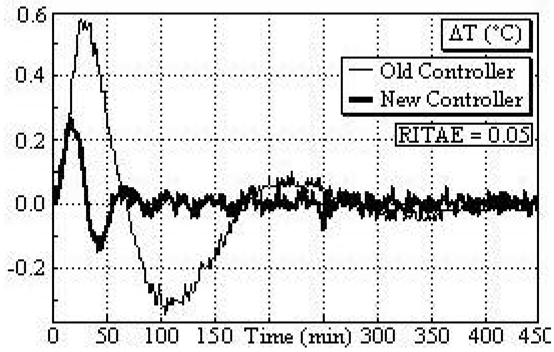

Fig. 7: Suppression of disturbance #2

detected by the analysis of the recorded plant data: the amplitude of deviations from set-point values is rather small, but it should be recalled that the a very tight control is required and that small deviations in reactor temperature propagate as larger perturbations in jacket temperature and flow rates.

For one case the technique was not able to identify the detected *anomaly* and this was explained in terms of two different perturbations acting simultaneously. For the other two cases, the application of the technique allows to improve the response: the perturbation is suppressed in a much lower time with the new controller and the RITAE index assumes values closer to zero (Figure 6 and 7).

6. CONCLUSIONS

The proposed method presents a complete approach to performance monitoring of SISO control loops, including different issues of detection of *anomalies*, process identification, controller retuning and improvements evaluation.

Only the first stage (detection) is accomplished in the DCS, with relatively low additional computational load, while remaining stages are developed in a parallel computer.

The capability of the proposed technique has been validated by simulation and by first off-line application to industrial data.

The technique shows to be very flexible, as it can be applied both off-line and on-line, can work complete automated procedure by using default settings or accept interaction with the operator in some key points.

Further work will be devoted to different objectives: introduction of constraints on control actions and of different objective functions in the retuning block; extension of the software package to detect valve stiction and sensor failures; on line applications to address implementation issues and to better define different levels operator interactions.

7. REFERENCES

Åström, K.J. and Hägglund T. (1995): "PID Controllers: Theory Design and Tuning"; *ISA, Research Triangle Park, N.C.(USA)*.

Ender, D.B. (1993): "Process control performance: Not as good as you think.", *Control Engineering*, **40:10**, pp. 180-190.

Hägglund T. (2002): "A friction compensator for pneumatic control valves", *Journal of Process Control*, Vol. 12, pp. 897-904.

Hägglund T. (1999): "Automatic detection of sluggish control loops", *Control Engineering Practice*, Vol. 7, pp. 1505-1511.

Hägglund T. (1995): "A control loop performance monitor", *Control Engineering Practice*, Vol. 3, pp. 1543-1551.

Hägglund, T. (2002). Industrial Applications of Automatic Performance Monitoring Tools*; Proc. 15th IFAC World Congress; T-Mo-A11-1*, Barcelona, Spain.

Harris T.J. (1989): "Assessment of Control loop performance", *The Canadian Journal of Chemical Engineering*, Vol. 67, pp. 856-861.

Marchetti, G. and Scali C. (2000). Use of Modified Relay Techniques for the Design of Model Based Controllers for Chemical Processes. *Ind. Eng. Chem. Res.*, **39**, 3325-3334.

Nelder J.A. and Mead R. (1965): "A simplex method for function minimization", *The Computer Journal*, Vol. 7, Issue 4, pp. 308-313.

Rossi M. (2002): Monitoring and Performance Evaluation of Controlled Loops; Ms Thesis; Chem. Eng. Dept. (University of Pisa).

Thornhill N.F., Seborg D.E. (2002): "New Directions in Control System Performance Monitoring", *T-Mo-A11, Session of 15th IFAC World Congress*, Barcelona.

Yu, C.C. (1999): "Autotuning of PID Controllers", Springer-Verlag, London (UK).