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Evaluation of advanced industrial control projects: a framework for determining economic benefits

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Abstract

A framework for the evaluation of advanced control projects is illustrated in this paper by way of a case study on level control in mineral flotation. The approach taken is to investigate how improvements brought about by advanced control can be measured to a required level of statistical significance, *after* the controller has been commissioned. Measured improvements are translated into increases in cash flow that result from implementing an advanced controller, which is then used for project evaluation. As an introduction to the case study, a framework for establishing advanced controllers for industrial processes, which culminates in determining economic benefits, is discussed. Ways of obtaining controller benefits through the reduction of downtime and product variations, are also described. The risk of implementing an advanced control project is discussed together with typical scenarios under which such projects are often undertaken.

Keywords: Economic evaluation; Industrial control; Implementation; Mineral processing; Mineral flotation; Multivariable control; Performance evaluation; PI controllers; Process control; Project selection

1. Introduction

The framework described in this paper resulted from the efforts of the authors and co-workers to measure the economic benefits that can be obtained from implementing Mintek's FloatStar flotation level control system (Henning, Schubert & Atasoy, 1998). This was done to illustrate to the designers of the controller and the potential clients, that the controller is not only a technical success, but an economic success as well.

It is easy to demonstrate in simulations and on industrial flotation circuits that the FloatStar controller is better at controlling the levels of multiple flotation banks than conventional single-loop PI controllers (Schubert, Henning, Hulbert & Craig, 1995). It is, however, more difficult to show that the FloatStar is an economic success as the achieved benefits, although significant in monetary terms, are often difficult to measure.

The control literature is not of much help here as little attention is paid to the measurement of monetary benefits resulting from improved control. A notable exception is the Warren Centre Advanced Control Project (Marlin, Perkins, Barton & Brisk, 1987, 1991) in which quantitative estimates of improvements in plant performance brought about by seven control upgrade projects are investigated. A general benefits analysis method is proposed that is applied to all seven cases. Estimates of improved operation relative to a base case are used together with throughput, time period, service factors and increases in operating profit, to calculate the resulting estimated benefit. Brisk (1993) continues on this theme, emphasizing that it is possible to obtain credible control benefit estimates. More recently, Lant and Steffens (1998) presented a quantitative statistical tool for performing benefits analysis studies in the control of wastewater treatment processes.

The benefit analysis method proposed by Marlin et al. (1987, 1991) provides a useful framework for estimating the benefits of controllers that have yet to be implemented. The "before and after" type experiments (improvement estimates relative to a base case) that are performed by Marlin and co-workers do however have severe shortcomings in producing statistically significant data for use in analyzing the benefits that result from

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existing controllers. This issue is addressed as part of the case study presented in this paper.

The approach taken here is to investigate how improvements brought about by advanced control can be measured to a required level of statistical significance, *after* the controller has been commissioned. Such a situation might seem unlikely at first glance, as economic benefit data are usually required to motivate capital for the implementation of an advanced controller. This issue is discussed in this paper as part of a qualitative analysis of the risk in implementing advanced control projects.

Measured improvements are translated into increases in cash flow that result from implementing an advanced controller. What to do with the resulting cash flow estimates is a subject in its own right, and will only be treated briefly here. Interested readers are referred to Allen (1991) for a comprehensive description of the economic evaluation of projects. Benefits resulting from more subjective measures such as safety and legal considerations, customer goodwill, and increasing prestige, will not be considered.

A framework for establishing advanced controllers for industrial processes, which culminates in determining controller benefits, is discussed in Section 2. In Section 3, ways of obtaining controller benefits through the reduction of downtime and product variations, are described. The risk of implementing an advanced control project is discussed in Section 4. Typical scenarios, under which such projects are often undertaken, are discussed. Finally in Section 5, a framework for the evaluation of advanced control projects is illustrated by way of a case study on level control in mineral flotation.

2. A framework for determining economic benefits

There are various processes involved in establishing a control system for a physical plant. These processes make many demands on the control engineer and the engineering team, and often manifest themselves in a step-by-step design procedure. Skogestad and Postlethwaite (1996) divide this procedure into 14 steps, step 1 being "Study the system (plant) to be controlled and obtain initial information about control objectives", and step 14 "Test and validate the control system, and tune the controller on-line, if necessary".

A framework for establishing advanced controllers for industrial processes, which incorporates these steps, is briefly described in Section 2.1. This framework, which is referred to here as the General Control Problem (GCP), culminates in determining the economic benefits of the controller, which is the subject of the rest of the paper. The steps followed in the economic performance evaluation process, as given in Section 2.2, are therefore part of the solution of the GCP.

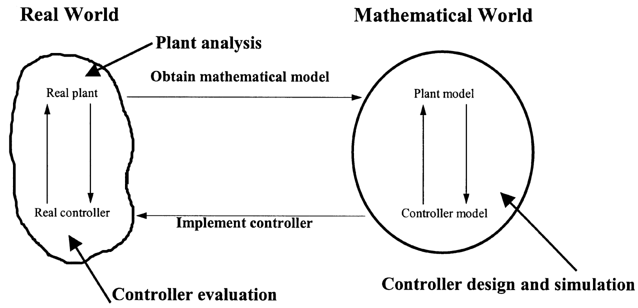

Fig. 1. The general control problem.

2.1. The general control problem framework

A pictorial representation of the GCP is given in Fig. 1 (Craig, 1997). This figure shows two worlds, a real world and a mathematical world. In the real world, plants are ill-defined and often difficult to describe, hence the rugged boundaries for this world. In contrast, the mathematical world is generally well defined, hence the circular boundary.

In the real world there is a real plant, the economic performance of which has to be improved through the implementation of an advanced controller. In order to do this, the project team has to do the following:

- *Analyze the plant for control purposes*. This includes the following:
  - perform a cost benefit analysis to get the initial project go-ahead (Marlin et al., 1987; Martin, Turpin & Cline, 1991)
  - obtain process knowledge;
  - classify process variables;
  - determine operational requirements that must be satisfied by the controller;
  - determine the control configuration and control law;
  - determine the role of the operator, before and after the implementation of the controller.
- *Obtain an adequate mathematical model for the plant*. This step is required in the GCP to go from the real world to the mathematical world. This model has to be as simple as possible while still allowing operational requirements to be met. The process used to obtain a model would depend on whether the plant exists or is still being designed, and whether the plant dynamics are well understood. Generally, both modeling and system identification will be used.
- *Controller design and control system analysis*. These tasks take place in the mathematical world. The plant model is used to design a controller according to the specifications determined in the plant analysis phase. It is critical to understand the limitations of the plant model (i.e. how different is it from the plant in the real

world?) and the impact that these limitations have on the controller design and analysis.

- *Controller implementation*. This step is required in the GCP to go from the mathematical world to the real world. Once the designer is satisfied that the controller works well in the mathematical world, and that it has a good chance of meeting the required specifications in the real world as well, the controller can be implemented.
- *Controller evaluation*. Once implemented in the real world, the performance of the controller needs to be evaluated, both functionally and economically. Does the closed-loop system in the real world meet the required functional specifications (as determined via a functional performance evaluation)? What is the economic benefit of the controller (as determined via an economic performance evaluation)?

The remainder of this work will focus on the last step in the solution of the GCP, i.e. that of controller evaluation. In particular, the case study described in Section 5 includes a functional performance evaluation (Section 5.3) and a detailed economic performance evaluation (Section 5.4).

2.2. Economic performance evaluation

The aim of an economic performance evaluation is to produce quantitative economic benefit measures (gut feelings are not good enough!) which are statistically significant. These measures are derived from plant data obtained from carefully designed experiments, and are used to show that the particular control system is a good investment. The control system is assumed to adhere to the functional specifications for which it was designed.

The following steps are typically followed in the economic performance evaluation process. They are applied to the case study in Section 5.4 where the relative economic performance of two control systems is determined.

- *Experiment design and data generation*: Design an experiment to generate unbiased production data that captures the economic performance of the process when the relevant control system is in use. Monitor the experiment and make sure it is carried out as planned.
- *Data analysis and hypothesis testing*: Analyze the generated data and determine the sample statistics for each control system. Test and accept or reject the hypothesis (e.g. "the new control system is better than the old one" — formally, a null hypothesis and an alternative hypothesis will be set up).
- *Monetary benefits*: Estimate the annual monetary benefit (increase in cash flow) resulting from the new control system.
- *Economic project evaluation*: Determine if the expected cash revenues due to the new control system is larger than the expected cash cost associated with the project.

Before describing an economic performance evaluation case, it is useful to consider how controllers can contribute to achieving economic benefits, as discussed in the next section.

3. Controller benefits

The primary objective of a control system applied in the process industry is to maximize profits by transforming raw materials into product while satisfying criteria such as safety, environmental regulations, product specifications and operational constraints (Seborg, Edgar & Mellichamp, 1989). In satisfying this objective, control system benefits can be obtained from minimizing downtime and reducing variations in the controlled variables.

3.1. Minimizing downtime

A distinction is made here between pseudo-plant-downtime and real downtime. Pseudo-downtime is defined as the time when the plant is able to produce but is in some unsteady state that controllers can influence, such as a startup phase. Real downtime is the time when the plant is unable to produce, e.g. when there is no raw material to process. The latter downtime falls within the industrial engineering domain and will not be discussed here (see e.g. Chase & Aquilano, 1992).

Pseudo-downtime can be reduced in essentially three ways, i.e. by preventing unscheduled plant shutdowns, by enabling fast recovery after process disruptions, and fast plant startups and shutdowns. These will now be discussed in turn.

The prevention of unscheduled plant shutdowns that are within the sphere of influence of the control engineer should form an integral part of any control strategy. This is particularly important in high-value-product industries. For example, in the petrochemical industry, one unscheduled shutdown could negate the benefits brought about by improved regulatory control over a one-year period. The use of control methodologies that allow for effective constraint handling, such as model predictive control (Clarke, 1994), goes some way to prevent such shutdowns from occurring. Fault detection techniques (Isermann, 1997) and associated contingency strategies can also play an important role in this regard. In general, one would expect unscheduled plant shutdowns to affect continuous processes more adversely than batch processes, as the latter are geared to startup and shutdown regularly.

In the operation of both continuous and batch processes, disruptions could occur which do not shut the process down completely, but take the process away from its steady operating condition. Fast recovery to steady operation after such process disruptions will also reduce pseudo-downtime.

The reduction of pseudo-downtime through fast plant startup and shutdown procedures can be significant, as discussed in Section 5.5. In this example, the process is complex with many interactions, resulting in the operator taking a long time to get the process to steady state manually. When such a process is regularly shut down for maintenance (e.g. once a week), a controller which deals with these interactions can significantly reduce the pseudo-downtime by getting a process to "steady-state" more quickly.

3.2. Reducing variations

One of the purposes of regulatory feedback control is to decrease product variability by negating the effect of disturbances and uncertainties on the variables being controlled (Seborg et al., 1989). A good feedback controller will reduce the standard deviation of the process parameters being controlled. This reduction, in itself, can result in significant economic benefits if the relative performance function, or relation between the controlled variables and money, is nonlinear or linear with constraints (Bawden & MacLeod, 1995; Schubert et al., 1995). Additional benefits can often be obtained from operating the process closer to a constraint, which is now possible due to the reduced fluctuations in the controlled variable.

Although an advanced controller is able to significantly decrease product variability, this is not enough to guarantee its long-term success. The strategy discussed in the next section will help the project team to assess the risk involved in the controller being a success. It also points to how the chances of getting the final go-ahead for the project can be increased.

4. The risk of advanced control projects

In this section a method is put forward to assess the risk, to the control practitioner, of implementing a control system that will be successful (functionally and economically) in the long term. Some risk reduction methods are also proposed. The modeling, assessment and management of risk, which is a measure of the probability and severity of adverse effects, is a subject in its own right (Haimes, 1998), and only a brief qualitative discussion will be given here.

4.1. Risk assessment

Haimes (1998) describes the risk assessment and management process in five steps: risk identification, risk quantification and measurement, risk evaluation, risk acceptance and avoidance, and risk management. In Cash, MacFarlan, McKenny & Applegate (1992) a portfolio approach to project risk assessment for information technology projects is proposed. A model is put forward which relates project risk to project size, the experience that the project team has with the technology, and the project structure (the degree of consensus as to what the project outcomes should be). These and other issues are not only important when implementing information technology systems, but should also be considered when implementing advanced industrial control systems.

Table 1
Management backing versus user acceptance

Management backing	User acceptance low	User acceptance high
Low	Very high risk	Medium risk
High	High risk	Low risk

A qualitative approach, based on the experience of the authors, is proposed in this section. It is aimed at providing a simple means of qualitatively capturing and analyzing the risk inherent in a control system implementation. The 2×2 grid shown in Table 1, which relates management backing to user acceptance, can be used to evaluate the degree of risk in implementing an advanced control system on a particular plant.

Depending on the amount of money involved in the purchase of an advanced control system, approval from the plant manager up to board level might be required. This implies that such a system has to be sold at numerous management levels before the go-ahead is received. Management is therefore used here as a collective term for the board chairman down to the plant manager.

The user is defined here as the eventual operator of the control system, but could under some circumstances also include the first level of technical supervision, such as the process engineer or plant metallurgist. In the context used here, the operators are more important as they will most likely have a longer direct association with the control system than the technical supervisor, who will most likely move into a management position with time. The user benefits of a control system discussed here are as perceived by the end-user and not by the control practitioner!

Both management backing and user acceptance have to be present for a control system implementation to be successful in the long term. As shown in Table 1, user acceptance is considered to be more important than management backing, because it is the user who has to operate the system on a daily basis. Management cannot watch over the operator 24 h a day to ensure that the system is used as intended. Operators must want to use the system because it makes their life easier by supporting them in achieving their goals. If there is little or no user acceptance from the operator, such a system will in time fail, even if strong management backing is present.

Each of the four quadrants in the grid will now be described in more detail.

Low management backing — low user acceptance. This is a very high risk situation. The likelihood of a control system being implemented under such circumstances is very low. However, when a controller is implemented on a free trial or rental basis, such circumstances could occur. It would be advisable for the control system practitioner to walk away from such a project, or if possible, to first work at improving the risk profile of the project.

High management backing — low user acceptance. This is a high risk situation which is characterized by management push rather than user pull. It occurs more often than it should, particularly in situations where management likes a new technology, motivated by gut feeling, and has it implemented without proper consultation with the end-user. A control system implemented under such a condition has little hope for long-term survival, and creates a dilemma for the control practitioner. Hard work is required, with management backing, to improve user acceptance. If this cannot be done, it is possibly best not to proceed with the project.

Low management backing — high user acceptance. This is a medium risk situation which is characterized by user pull. It is not very likely to occur in practice when an outright system sale is involved, as board approval for the purchase of such a system is unlikely. On a free-trial or rental basis, such a system has a good chance of survival, as the user acceptance will in all likelihood spill over into the management ranks.

High management backing — high user acceptance. This is a low risk situation which is characterized by user and management pull and is the ideal for any control system practitioner.

4.2. Risk reduction

The assessment of risk as discussed in Section 4.1 indicates which issues need to be addressed in order to reduce the risk of project failure. This section describes some ways of increasing management backing and user acceptance, in particular, benefit assessment of control systems and social aspects of transferring control technology, respectively. Other important risk reduction aspects such as sound project management, will not be discussed here.

On assessing the benefits of control systems. Plant managers are often skeptical of project benefit claims (Brisk, 1993) — and with good reason. If all promised benefits over the years had materialized, plant recoveries should now be in excess of 100%! Quantitative benefits, although difficult to obtain, are essential in achieving high management backing and hence reduce project risk. A cost-benefit analysis (Marlin et al., 1987, 1991) can be used to obtain the initial project go-ahead. Section 5 deals with how such quantitative benefits can be measured, *after* the controller has been commissioned, and converted to measures with which management is comfortable, such as net present value (NPV). The advantage of such quantitative measures are twofold:

- It provides management with figures that can be used to motivate capital for a control system
- It acts as a motivating target for the project team that implements the system.

Control systems can also provide benefits which do not lend themselves to evaluation in monetary terms, e.g. flexibility, and the impact of technologies on the company's long-term competitive position (Martin, 1993). Such non-monetary attributes complement financial considerations.

Some social aspects of technology transfer. Social aspects of technology transfer are often ignored by the control practitioner. The practitioner usually concentrates on providing a reliable control system, based on sound process knowledge and control technology. Although this is essential for project success, and can often go a long way in reducing project risk by increasing user acceptance, it alone is not enough. Control systems in the processing industries have to operate in a human environment. This fact is often ignored in the design of control systems, which results in the failure of automation projects. Common causes for failure that have been identified are; insufficient automatibility, inadequate user-system interfaces, and incompatibility between human needs and system requirements (Martin, 1993).

Therefore in order to increase user acceptance and hence reduce project risk, the end user must not only be taken into account when the system is designed, but should preferably be part of the design team/process. Such a strategy will

- produce a relevant user-system interface;
- provide a valuable training opportunity;
- help in developing a sense of ownership.

These aspects are crucial to long-term project success.

In countries with high levels of unemployment, there is often a natural resistance to automation, as automation can be perceived by organized labour to go hand-in-hand with job losses. High user acceptance is difficult to achieve under such conditions. An operator might argue that 'if this control system is a success, I or some of my colleagues might lose our jobs'. A control practitioner needs to be aware of the arguments for and against this position, and be able to relate these arguments to the control system being implemented.

5. Case study: level control of a mineral flotation circuit

The economic evaluation of a level controller for a mineral flotation circuit is described here using the framework discussed in Section 2. Flotation is often the first concentration process that an ore encounters between mining and its final purified state. After the ore is removed from the mine, it normally passes through a crushing plant where the rocks are reduced to the size of pebbles. After crushing, the material is treated in a milling circuit, where the intention is to reduce the particle size distribution until most of the valuable material is exposed and amenable to some form of upgrading. Flotation, which is described in Section 5.1, often follows as the first process that separates the ore into a higher-grade concentrate and a waste or tailings stream. After flotation, the concentrate is purified further by smelting or roasting. Sometimes the flotation tailings stream is treated by some other process (for example leaching) to recover the last of the high-value material.

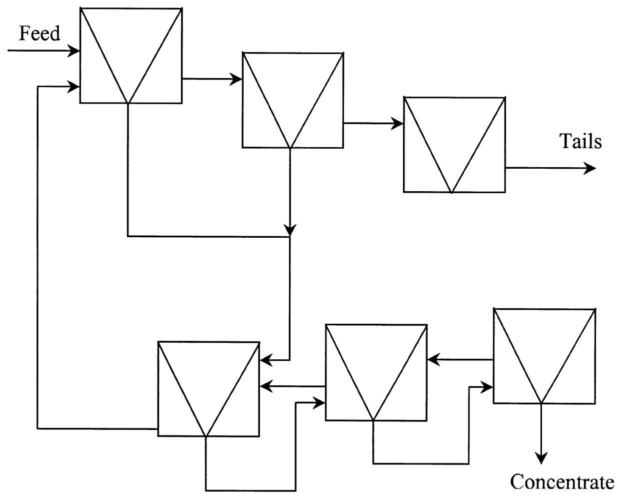

Fig. 2. A generic flotation circuit.

5.1. The flotation process

Flotation is a process widely used in the concentration of mineral-bearing ores. It is used extensively as a first step in the recovery of pyrite, platinum group metals, copper, nickel, zinc, lead, fluorspar, antimony and coal. The process is based on the difference in surface properties of the different minerals present in the ore. These properties can be altered or enhanced by the addition of a suite of reagents. The ore is introduced into a flotation cell as a slurry, which is agitated to keep the particles in suspension. Air bubbles with an appropriate size distribution are mixed with the slurry, and particles with hydrophobic surfaces will attach to the bubbles. These rise to the top of the cell, where a froth forms. The froth flows over the cell lip, forming a concentrate that is of a higher grade than the slurry fed to the cell. Since a single flotation unit will provide limited upgrading, a network of units is typically required to achieve the required concentration. Fig. 2 shows an example of a flotation circuit. Further examples may be found in Lynch, Johnson, Manlapig and Thorne (1981).

The process is more complex than this brief introduction might suggest. The state of the froth plays a crucial role in the ore concentration. Solids enter the froth by two principal mechanisms. The first is bubble attachment, which favours the desired minerals. The second mechanism is entrainment in the liquid surrounding the bubbles as they enter the froth. Entrainment is non-selective, and a gangue particle and a mineral particle have the same likelihood of being swept up into the froth. Once in the froth, attached particles will probably stay there unless the bubbles collapse, while entrained particles are likely to drain back to the pulp if the froth's residence time is high enough. The depth of the froth is a key variable. A shallow froth with a low residence time will allow a high flow rate to the concentrate, albeit with a high proportion of gangue. A deep froth with a long residence time will allow solids to drain back to the pulp, giving a higher concentrate grade, but a lower mass-pull to the concentrate launder. Thus setting the depth of the froth allows one to choose between recovery and grade. This trade-off between recovery and grade is graphically illustrated in Lynch et al. (1981).

5.2. Flotation level control

The froth depth is controlled by manipulating a valve which lets pulp flow out of the flotation cell. In the case of a single unit, it is a simple matter to control the level with a standard PI (Proportional and Integral) loop. In the more common case of an interconnected circuit of cells, level control is more complex due to interactions between adjacent levels. Moving a valve to correct the level in a unit introduces a disturbance in the next unit downstream. Traditionally, each flotation unit is controlled independently by a PI controller. An alternative is to design a controller which considers all the levels in the circuit simultaneously, such as the FloatStar level stabiliser developed at Mintek (Schubert et al., 1995).

Initial simulation work indicated that a multivariable approach to level control would lead to better regulation of flotation levels. A frequency domain analysis of the resulting sensitivity transfer function matrices, using singular values (Skogestad & Postlethwaite, 1996), was done to compare the disturbance rejection properties of the FloatStar with those of a set of SISO (single-input-single-output) PI controllers. A test circuit consisting of three flotation units in series was used for the comparison. The configuration of the feedback system is shown in Fig. 3, with the sensitivity transfer function matrix ($S_i(s)$) being the transfer function between output disturbances (d) and the error signal vector (e), i.e.

$$S_i(s) = [I + G(s)K_i(s)]^{-1}, \quad i = \text{PI or FS},$$

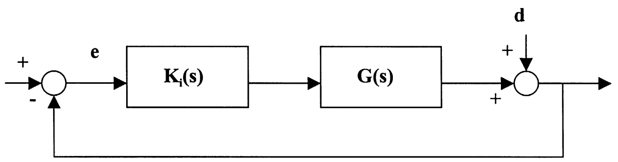

Fig. 3. Block diagram of controller and plant.

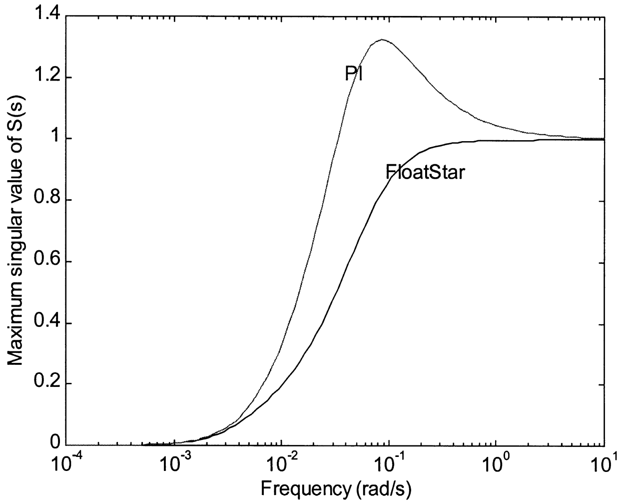

Fig. 4. Maximum singular values of $S(s)$ for the FloatStar and PI controllers.

where $G(s)$ is the plant transfer function matrix for the three flotation units in series, and $K_i(s)$ the transfer function matrix of the PI or FloatStar (FS) controllers.

Fig. 4 shows the maximum singular values ($\sigma_{\max}[S_i(j\omega)]$) of the sensitivity transfer function matrices of the two control methods. The higher the value of the function at any given frequency, the greater the effect of a disturbance at that frequency. It can be seen that at low frequencies both control systems are very effective at suppressing disturbances. At high frequencies (>1 rad/s) neither controller can do much about incoming disturbances. However, in the range 0.01–1 rad/s the FloatStar is clearly superior in this analysis. These frequencies correspond to disturbances with a period of between approximately 0.1 and 10 min. Many of the disturbances seen on the trends of flotation levels fall into this frequency range. It was therefore possible to predict that the FloatStar controller would provide an improvement in practice.

Schubert et al. (1995) take this theoretical analysis further and try to predict the metallurgical benefit that could be expected from a reduction in level fluctuations. The conclusion was that an improvement of 1% in recovery could be expected. The next step was to demonstrate that measurable benefits could be obtained on industrial flotation circuits.

5.3. Functional performance evaluation

There are two issues to consider when comparing the performance of FloatStar with that of a set of PI controllers. The first issue is whether FloatStar does, in fact, give better level control. The second question, more difficult to resolve, is whether improved level control leads to better metallurgical performance.

The improvement in level stabilisation resulting from multivariable control can easily be seen when looking at trend displays of level versus time. Examples can be found in Schubert et al. (1995) and Henning et al. (1998). No experimental design is required to detect the change, as illustrated in Fig. 5.

5.4. Economic performance evaluation

It is more difficult to show what the resulting metallurgical benefits are. The problem is to show, with statistical confidence, that an improvement has taken place against a background of relatively large plant variations. The expected benefit due to FloatStar is about 1% of recovery. The variation in recovery from shift to shift on a flotation plant is often 10% or more. Most of this variation will come from fluctuations in the characteristics of the feed ore, but some variations can arise from long-term changes in the operation and layout of the plant. For example, during the six months of a FloatStar trial on a pyrite flotation circuit, plant throughput was increased considerably, giving rise to a coarser particle-size distribution from the milling circuit. The larger particle size affects flotation performance. In addition, a change was made to the circuit configuration during the trial, and a different technique was used for sampling the feed in order to measure grade. In the absence of a suitable experimental design, these changes would invalidate the assessment of any process innovation.

Plants often try to evaluate a new concept by doing a single 'month-on, month-off' trial, i.e. running one system for a month and then running the competing system for another month, sometimes with a lengthy break in between the two parts of the trial. This will never be an acceptable way of obtaining statistically valid information, since operating conditions will invariably have changed during the two months under consideration, and these changes can overwhelm the effects that the experiment was designed to test for. Is the observed improvement (or deterioration) due to the system being tested, or is it caused by the new shift boss, the new reagent or the ore suddenly being mined in a different part of the reef? Only if the experiment has been carefully planned can one be confident of the cause and effect relationships.

Another common but erroneous approach is to compare performance of adjacent but independent circuits. For example, if circuit A is regulated by controller A and

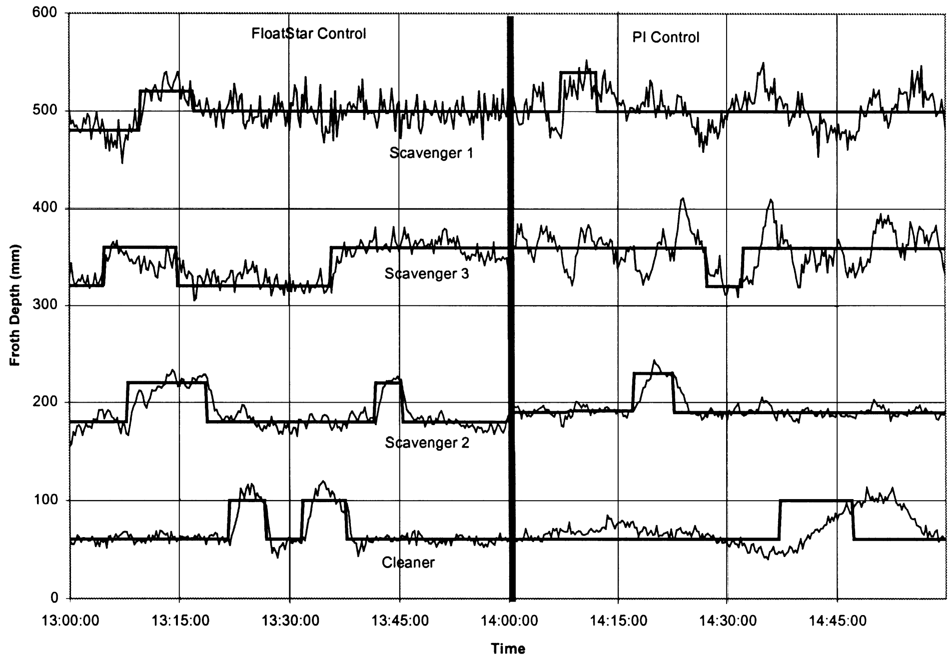

Fig. 5. FloatStar and PI level control comparison.

circuit B is regulated by controller B, surely one can simply monitor the performance of circuits A and B and assume that if circuit B works better, controller B must be better than controller A? This would be true only if the circuits were absolutely identical, which is never the case. Feed quality and throughput often vary, maintenance histories will differ, as can the amount of operator intervention. In a recent instance, two adjacent flotation circuits were declared to be identical by their production personnel. Closer examination revealed that one circuit received a far coarser feed from its milling circuit. In addition, it received all spillage material from the section, and this usually had a detrimental effect. The recovery of this circuit was consistently lower, but not by a predictable amount. It would be inadvisable to use these two circuits as a test-bed for a comparative study. A far more reliable approach is to use the experimental approach outlined below. Only one circuit is considered, and a *repeated* on–off regime is adopted, with switching taking place as often as is feasible.

Experiment design and data generation. FloatStar performance has currently been evaluated on five industrial flotation circuits, treating platinum, pyrite and nickel. In each case a trial schedule was drawn up in which level control was alternated between FloatStar and a set of PI loops. The length of each period depended on conditions on the particular plant, but varied between 1 and 3 days. The total length of the trials varied between 1 and 6 months. The time between switching depended on the residence times of the circuit and the frequency with which grades were analysed. It is important to decide on the rules of the trial in advance and to get agreement from all interested parties. The rules of thumb used in designing a trial are:

- Change only one parameter (FloatStar ON/PI ON cycle — also referred to here as an ON/OFF cycle) at a time as far as this is possible.
- Switch as frequently as possible between the operating conditions being tested (i.e. between different controllers). This will distribute the effect of unplanned variations equally between the systems being compared. Such variations will then only affect the results of the trial if they are highly correlated with the ON/OFF cycle (e.g. if the operator adds a different reagent during ON periods only), which is of course highly unlikely.
- Each ON and OFF period must be significantly longer than the longest time constants of the process, to allow for transition to the new steady-state and a sufficient period of data collection.

Ideally, the period of transition to the new steady state should be excluded, but this is not always possible.

One particular trial will be discussed in more detail to give an idea of the issues that arise. Background to the trial and the process can be found in Henning et al. (1998). In this trial FloatStar was run for a day, then PI control for a day. Switching was planned to coincide with the start of the morning shift. There are four shifts per day, and a composite sample is collected during each shift. The samples were analysed in the plant's laboratories and recoveries were calculated via the two-product formula (Lynch et al., 1981):

$$R = \frac{100c(f-t)}{f(c-t)},$$

where R is the recovery (%), c is the assay of metal in the concentrate, t is the assay of metal in the tails, and f is the assay of metal in the feed.

In practice, the switching did not always take place as planned, and it was necessary to draw up criteria for the rejection of data. If switching was done too long after the start of a shift, the data from that shift would not be used. Sometimes not all levels were switched over, and so control deemed "mixed-mode" was excluded from the data set. Data was also excluded for shifts during which the SAG (semi-autogenous grinding) mill providing the feed was down for too long. Shifts of exceptionally low throughput were also rejected. The detailed criteria were determined jointly by plant and Mintek personnel, and this was done *prior* to the data analysis.

Data analysis and hypothesis testing. Two different methods were used to assess whether there was a significant difference in the metallurgical results obtained with the two control systems. The t-test is the most commonly used method to evaluate the differences in means between two groups (Levin and Rubin, 1991). A null hypothesis (H_0) is set up, which assumes that no change has taken place unless proved otherwise. An alternative hypothesis (H_1) is also postulated, for example:

- The mean gold recovery when FloatStar is controlling (μ_{GRFS}) is higher than the mean gold recovery when the PI loops are controlling (μ_{GRPI}).
- The mean gold assay in the tailings stream is lower when FloatStar is controlling (μ_{GTFS}) than the mean tailings assay when the PI loops are controlling (μ_{GTPI}).

A symbolic statement of the problem, using the first alternative hypothesis given above, then becomes

H_0: $\mu_{GRFS} = \mu_{GRPI}$,

H_1: $\mu_{GRFS} > \mu_{GRPI}$,

$\alpha = 0.05$.

To reject H_0 with 95% confidence, the observed difference of the sample means would need to fall sufficiently high in the right tail of the distribution (top 5% of area in this case). Then the hypothesis that the FloatStar controller leads to better gold recoveries (H_1), cannot be rejected.

With hypotheses like these, a single-sided t-test is appropriate, as one is testing for an expected improvement. In cases where there is no a priori expectation of a difference, a two-sided t-test is used. An example of this would be testing for a difference in mean head grade between the two groups. Here the null hypothesis would still be that there is no difference, but the alternative hypothesis is that the mean head grade when FloatStar is controlling (μ_{HGFS}) is greater than *or* less than the mean head grade when the PI loops are controlling (μ_{HGPI}).

A symbolic statement of the problem is

H_0: $\mu_{HGFS} = \mu_{HGPI}$,

H_1: $\mu_{HGFS} \neq \mu_{HGPI}$,

$\alpha = 0.05$.

To reject H_0 with 95% confidence, the observed difference of sample means would need to fall sufficiently high in the tails of the distribution (2.5% of the bottom or 2.5% of the top area in this case). If not, it cannot be concluded that the mean head grade was different for when the FloatStar and PI controllers were operating.

The test assumes that the data are normally distributed within the groups, and that the sample variances are not significantly different to each other. The t-test itself is available in any of the standard statistical software packages, along with means of checking that the assumptions are met, e.g. Statistica (1997).

In this test, all the data gathered when FloatStar was controlling is collected into one group, and all the data gathered when the PI controllers were operating is collected into a second group. Another approach is to use a paired t-test. Here the data set is split into sequential on–off pairs and the difference in means for each on–off pair is calculated. The null hypothesis is that the mean of these computed differences is not significantly different from zero. It is argued (Napier-Munn, 1995) that the use of the paired t-test reduces the effects of any long-term trends in the data.

The results of these tests can also be confirmed by non-parametric tests, in which ranks rather than actual values are used.

Results — hypothesis tests. The results of this particular on–off trial were analysed by the plant, by Mintek, and also by the Julius Kruttschnitt Mineral Research Centre at the University of Queensland, Brisbane, Australia. There were slight differences in results due to the ways in which data were coupled into sequential on–off pairs. However, the result of all the various tests was that FloatStar gave a significant improvement.

The first analysis was a t-test on the entire valid data set. This consisted of 319 6-h shifts under PI control and 233 shifts under FloatStar control. The test showed (using a 95% confidence level) that there was:

- no significant difference in mean head grade (H_0: $\mu_{HGFS} = \mu_{HGPI}$ could not be rejected);
- no significant difference in mean concentrate grade;
- gold recovery was significantly higher with FloatStar on (H_0: $\mu_{GRFS} = \mu_{GRPI}$ was rejected).

The measured improvement in gold recovery due to FloatStar was 1.06%.

The main objective is to test whether the gold recovery has improved. However it is necessary to confirm that this improvement was not at the expense of concentrate grade (which should remain constant at a specified level). It is also necessary to confirm that the mean head grade is the same for both sets of data, because recoveries are normally higher when the feed grade is higher. It was also necessary to confirm that mean throughput was the same for the two data sets, as performance is normally better at lower throughput.

The second analysis was a paired t-test (Napier-Munn, 1995) using all the available on–off pairs with no lost shifts, i.e. all two-day sequences with a full day of PI control followed by a full day of FloatStar control. Thirty-one data pairs were found, reflecting 62 days of data. Once again there were significant differences in gold recovery and sulphur recovery.

The measured improvement in gold recovery due to FloatStar was 1.27%.

Further paired t-tests were performed on different data sets, obtained by including days on which there were interruptions due to SAG mill downtime or incorrect switching. All tests indicated that the null hypothesis (H_0: $\mu_{GRFS} = \mu_{GRPI}$) could be rejected.

Results — monetary benefits. Since there is a direct link between gold recovery and economic benefit, the plant management was able to calculate that FloatStar would pay for itself in 47 days in the most conservative analysis. (The payback rule (Ross, Westerfield & Jordan, 1993) states that an investment is acceptable if its calculated payback period is less than some pre-specified time period.) As the payback period is very short, it was not worth doing any further calculation. A simple benefit calculation, using hypothetical figures, is however shown below for illustration purposes.

Now that it has been established that the measured improvement in gold recovery due to FloatStar was about 1%, the monetary benefits resulting from better flotation level control can be estimated. This can be done using the benefit calculation method described in Marlin et al. (1987). The variables used in the benefit calculation are summarized in Table 2.

The annual benefit is calculated as

$$AB = IM \times SF. \quad (1)$$

The 1% improvement in recovery figure is obtained from measuring the improvement in the mean tailings grade of 0.01 g/t (TGFS-TGPI). The total additional gold produced for the period of interest can now be calculated simply by multiplying the improvement in mean tailings grade by the annual plant tailings throughput. This can be done as there is no significant difference between the two controllers as far as mean head and concentrate grade are concerned. If this were not the case, the non-linear grade-recovery trade-off (Lynch et al., 1981) would have had to be used to obtain the true increase in gold recovery. (This grade-recovery trade-off is the performance function for mineral flotation, as discussed in Section 3.2, and explains why a reduction in level variations result in improved recovery (Schubert et al., 1995)). Therefore,

$$Ms = (TGPI - TGFS) \times Mp \times \frac{Mpf}{100} \times CF$$

$$= 0.01 \times 9.4e6 \times 0.97 \times \frac{1}{31.1} = 2932 \text{ ounces.} \quad (2)$$

The improvement (IM) is the increase in gold recovery per annum times the price of gold, i.e.

$$IM = Ms \times \mathrm{Pr} = 2932 \times 300 = \$879\,550. \quad (3)$$

The service factor is a function of the unit service factor, i.e. the fraction of time that the process unit is in mode of operation applicable to the control strategy, and a control service factor, i.e. the fraction of time that control improvement is achieved (Marlin et al., 1987). The control service factor is a product of availability of field instruments, control room equipment, use of control strategy, ability of control scheme to achieve benefits (when in service). The unit service factor is a product of, e.g. maintenance schedules and the availability of raw materials. Service factors for continuous processes are usually higher (e.g. > 0.95) than for batch processes. In this case a service factor of 0.95 is assumed. Substituting this value and the improvement into Eq. (1), the annual

Table 2
Variables used in the benefit calculation

Description	Abbreviation	Units	Value
Annual benefit	*AB*	$	Variable
Improvement	*IM*	$	Variable
Price of gold	*Pr*	$/ounce	300
Annual plant throughput	*Mp*	Ton	2 million
Material by mass reporting to tails	*Mfp*	%	97
Mean tails grade with FloatStar	*TGFS*	g/t	0.47
Mean tails grade with PI control	*TGPI*	g/t	0.48
Annual additional gold produced	*Ms*	ounce	Variable
Conversion factor	*CF*	ounce/g	1/31.1
Service factor	*SF*	None	0.95

benefit derived from the new controller becomes

$$AB = IM \times SF = 879\,550 \times 0.95 = \$835\,572. \quad (4)$$

Results — economic project evaluation. In order to evaluate whether the installation of the new controller and the related instrumentation are economically viable, the expected cash cost and cash revenues associated with the project must be estimated. These values can then be used together with capital budgeting tools, such as the net present value (NPV), to evaluate the project (Allen, 1991; Ross et al., 1993). NPV is estimated using the following steps:

- Estimate the future cash flows that the investment is expected to produce.
- Apply the discounted cash flow procedure to estimate the present value of those cash flows.
- Estimate the NPV as the difference between the present value of the future cash flows and the cost of the investment.

For economic valuation purposes it is assumed that the new controller is operating as promised and that it results in additional cash revenues of \$835 572 pa. (as indicated earlier). The data used for project evaluation are given in Table 3, and the projected cash flows in Table 4.

The NPV estimate is large positive, i.e. the present value of the future cash flows is significantly more than the cost of the investment. The project should thus be accepted based on the cost of the investment and the estimated cash flow that will result from increased gold recovery brought about by the FloatStar controller.

Table 3
Data used for project evaluation

Variable	Value
Cash revenues	\$835 572 pa.
Cash cost (e.g. maintenance)	\$24 000 pa.
Expected life span	5 years
Equipment salvage value (after 5 years)	\$10 000
Project cost (new controller plus commissioning)	\$100 000
Discount rate for such projects	20%

The results obtained in Table 4 are obviously dependent on the values chosen in Table 3. The chosen discount rate could, e.g. have a significant effect on the final result. In this case the choice of discount rate is almost irrelevant as the NPV is so overwhelmingly positive.

5.5. Additional benefits

The benefits described so far result only from reductions in level fluctuation. The plant also reported additional benefits. Since more gold was recovered in the flotation circuit, less gold entered the leach circuit that treated the flotation tailings. This in turn meant that less cyanide was needed in the leach circuit. There was also a very positive response from the operators due the fast stabilisation FloatStar gave after plant shut-downs and major disruptions. An attempt was later made at a platinum plant to quantify the benefits of reducing the time taken to reach steady-state. Schubert, Henning, Gebbie and Valenta (1999) show that the time taken to reach steady-state conditions can be reduced from about 150 min (with PI control) to 50 min (with FloatStar control). Quicker settling times mean that far less valuable material is lost to final tails during start-up or after plant disturbances. Monetary benefits derived from quicker settling times (reduction in pseudo-downtime as discussed in Section 3.1) can also be obtained using the methods described in Section 5.4. Such an analysis will however not be done here.

6. Conclusion

This paper described a framework for measuring economic benefits that result from applying advanced control to an industrial process. The use of this framework was illustrated by way of a case study on level control in mineral flotation using the Mintek FloatStar level control system. In this case it was easy to demonstrate that the FloatStar controller is better at controlling the levels of multiple flotation banks than conventional single-loop PI controllers. It was more difficult to show that the

Table 4
Project cash flows

Time (years)	0	1	2	3	4	5
Initial cost (\$)	− 100 000					
Inflows(\$)		835 572	835 572	835 572	835 572	835 572
Outflows (\$)		24000	24000	24000	24000	24000
Net inflow (\$)		811 572	811 572	811 572	811 572	
Salvage (\$)						10000
Net cash flow (\$)	− 100 000	811 572	811 572	811 572	811 572	821 572
Present value of future cash flows (\$) =	2 431 117					
NPV (\$) =	2 331 117					

FloatStar is an economic success as the achieved benefits, although significant in monetary terms, are often difficult to measure. A planned experiment such as that outlined in the paper is required to demonstrate the economic improvement. When such an experiment is used, it is possible to test for improvements that are small in relation to the typical fluctuations of a plant. These relatively small improvements are often very significant in economic terms, and the systematic assessment of competing alternatives can lead to major savings.

The framework for measuring economic benefits was developed as the control literature was only partially helpful in providing means of determining monetary benefits resulting from improved control. In industry, "before and after" comparisons are often used, as are comparisons between adjacent and similar circuits. Both of these methods have severe shortcomings in producing statistically significant data for use in analyzing benefits. The main problem is the impossibility of ensuring that all factors are identical except for that being tested.

As an introduction to the case study, a framework for the establishment of advanced controllers for industrial processes, which culminates in determining controller benefits, was given. Ways of obtaining controller benefits through the reduction of downtime and product variations, were also described. The risk of implementing an advanced control project was discussed together with typical scenarios under which such projects are often undertaken.

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