Chapter 14

Process Control

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14.1 Introduction

In general terms, control is concerned with the manipulation of the inputs to a system (a machine, process or plant) so that the outputs meet certain specifications. Control is thus a broad concept which encompasses long-term process operating strategy based on process evaluation, as discussed in Chapter 13, as well as manual control and the various forms of real-time automatic control such as logic or sequencing control, single-variable or multivariable continuous control and supervisory control. These various forms of real-time control are collectively known as process control and form the subject of this chapter.

In order to give a satisfactory treatment of the subject, both the science and the technology of process control must be dealt with. The science of process control includes the theories of dynamic modelling, feedback, stability, disturbance rejection, interaction and controller design. These principles are therefore initially reviewed. The technology of process control includes the vast technology of process measurement and monitoring as well as aspects such as the interfaces between the process, the operator and the control system and centralized and distributed control system architectures. These aspects are discussed further in subsequent sections.

An important characteristic of ore treatment plants is the difficulty in obtaining on-line measurements of key process variables and the large maintenance effort required to keep the process sensors operating reliably. This is primarily due to the aggressive properties of process streams such as pulps, the harsh environment encountered in extraction plants and the difficulty of designing reliable and representative sampling systems. These are perhaps the most important distinguishing characteristics of process control in ore treatment. Process measurement technology for gold extraction is discussed later in this chapter.

It is by no means easy to answer the questions that arise when considering the introduction of process control in ore treatment plants. Some of these questions are:

a) What are the key process variables that should be controlled?
b) Is there an economic justification for control?
c) What can be controlled?
d) What control philosophy should be used?
e) Will the reliability of the proposed control system be high enough?

Questions (a) and (b) are closely interrelated and concern the determination of the need for process control. In the past, control system design was usually done only after the process design had been completed. This still happens in many cases. However because of advances in on-line sensors and the availability of flexible and sophisticated computer-based control systems, the interaction between process flowsheet design, equipment selection and process control system design can now be exploited. Much can be gained by linking process design and control system design for new plants. It remains difficult, however, to quantify the economic benefits of a proposed process control system. Some financial models have been proposed for this purpose (Purvis and Erickson, 1982; Chang and Bruno, 1983).

Question (c) reduces to the following three sub-questions:

What can be measured?
Are suitable actuators available?
What configuration of control loops should connect the sensors and actuators, and will it be possible to obtain satisfactory dynamic performance?

This last sub-question will be discussed under the principles of automatic process control. Question (d) relates to operational aspects such as whether to use manual or fully-automatic control and whether to use analogue or computer control. Question (e) concerns detailed system design, equipment selection and maintenance issues. This chapter aims to provide essential background material on process control theory and technology for ore treatment applications in the gold industry.

14.2 Historical Background

Although standard commercial instrumentation and process control equipment had been widely and successfully applied in the chemical and petrochemical industries since the 1930's, as Fletcher (1972) pointed out, instrumentation first found its way into South African gold extraction plants only in the 1950's. Even through the 1960's, instrumentation and process control were not regarded as being as vital in the gold industry as they were in the advanced chemical industries. Fletcher stated that by the early 1970's two schools of thought had emerged. One encouraged instrumention and automatic process control in spite of the poor performance of the equipment then available and of the general shortage of qualified instrument maintenance personnel and engineers who were experienced in the special requirements of ore treatment. The other believed that efficient process operation could be achieved more readily via manual control by better-skilled operators. At the time of writing (1986), the latter opinion is rapidly disappearing and the benefits of process control and optimized operation are becoming more widely accepted. Fairly advanced automatic control systems using mimic panels for the operator interface have been extensively used during the last 10 years and computer-aided monitoring and control and the use of sophisticated video displays for the operator interface are now fairly common in modern ore treatment plants.

14.3 Principles of Continuous Process Control

14.3.1 Control objectives

It is necessary to approach the subject of process control system design from a wide perspective. The control of each individual unit in an ore treatment plant is complex and control of the overall operation of an entire plant is an extremely complex undertaking. Furthermore, it is well known that it is very difficult to reconcile the large-scale real problems which are encountered in process control practice with the simple well-defined problems treated in standard control theory textbooks (Dorf, 1986; Stephanopoulos, 1984; Smith and Corripio, 1985).

The first and usually also the most difficult step in control system design is to understand and formulate clear control objectives. A good understanding of long-term process operating strategy is essential for this. The complexity of the problem can be appreciated by considering the following factors underlying operating strategy in gold extraction plants.

The most important factor is that ore that is brought to the surface must be treated in the gold plant as quickly as possible in order to minimize the ore inventory held on surface. The reason for this is that the mining operation represents approximately 90% of both capital investment and operating cost, and any untreated ore inventory is thus an extremely costly investment. Also, especially in the case of older plants with a large number of small individual process units and little or no instrumentation, both plant design and plant operating procedures must be very conservative so that the fluctuating throughputs dictated by mining can be accommodated with only a small effect on gold recovery. Another important factor is that such vital information as the amount of gold in the incoming ore feed to the plant and its physical properties is not available to process operators and metallurgists until the following day. The plant must therefore be run without knowledge of the material currently passing through it. This also dictates conservative operation.

Operating strategy based on the above factors has as its prime requirement the need to maximize mill throughput which is therefore a typical key control objective. Other possible control objectives are improved recovery of the valuable minerals, improved head grade to cyanidation or reduced operating costs — for example, by savings in reagents or more productive use of plant personnel.

It is normal practice to focus on certain specific high-level control objectives. These would vary from plant to plant and would also change from time to time within one plant. When the demand for gold is high, for example, the objective might be to maximize throughput. When demand is low, improved recovery and/or lower operating costs could possibly become the key objectives. It may be very important to design sufficient flexibility into a process control system to allow for such changes in objective.

The control objectives for each individual process within an ore treatment plant must also be clearly stated in relation to the overall plant objectives. For example, in a grinding circuit, possible objectives would be to maintain the finest possible product size at constant throughput, to maximize

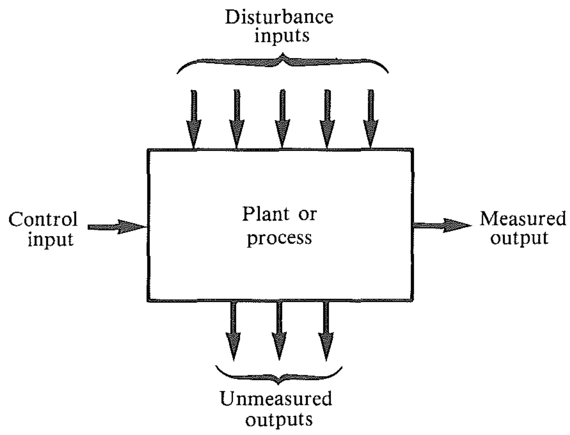


Figure 14.1. Classification of process variables.

throughput and keep product size within a limited range or to maximize downstream plant performance (Lynch, 1977). Such different objectives can require significantly different control schemes.

The high-level objectives discussed above all imply some form of performance optimization. Before any such optimization can be attempted, it is essential to be able to steer the process towards the desired steady-state operating point and to keep it there. The primary control objective in any overall process control scheme is therefore that certain key physical variables (e.g. flows, concentrations, densities, levels, temperatures, pressures and speeds) are kept as close as possible to their target values, called set-points, for as much of the time as possible. It is necessary for the control system to manipulate suitable process inputs to force this to happen in spite of disturbances. The presence of disturbances and uncertainty as to the exact behaviour of the plant are in fact the fundamental reasons for using control.

As mentioned above, the practice of conservative plant operation is natural in older plants with complex flowsheets and little or no instrumentation. Developments in gold process plant design during the 1970's and 1980's, such as the trend towards simpler flowsheets using high-capacity run-of-mine mills, a small number of high-capacity thickeners and continuous downstream recovery processes, together with improvements and greater flexibility in instrumentation and control system technology, have resulted in a growing ability to control plant operation and have therefore also created new options in process operating strategy. A clear specification of control objectives is the key to exploiting these developments for the optimization of plant economic performance.

14.3.2 Classification of process variables

In dealing with process control concepts, it is useful to classify the physical process variables into two categories:

Outputs: These are the key process variables to be kept as close as possible to their set-points, that is, controlled. The outputs can be further sub-classified as measured outputs and unmeasured outputs.
Inputs: These are the variables that, when changed, cause one or more outputs to change. The inputs can be further sub-classified as control inputs and disturbance inputs.

Sometimes the control objectives relate to properties that cannot be measured, that is, unmeasured outputs. In such cases, other variables which can be measured easily and reliably must be used to infer the unmeasured output.

The control inputs are also called manipulated variables. These are the variables that are changed by the controller to drive the outputs towards their set-points. All other process variables that affect the outputs in any way are called disturbance inputs or process load variables. Variations in these disturbance inputs cause unwanted changes in the process outputs. These definitions are summarized in Figure 14.1.

14.3.3 Control system structuring

As indicated in Section 14.3.1, the real problems that must be tackled in control system design are large and complex. On the other hand, control theory deals only with certain small, well-defined problems. Unfortunately there is, as yet, no systematic approach for bridging the gap between these two extremes. The design of a plant-wide control system must be approached on the basis of successive refinement by trial and error; it is therefore highly dependent on the experience and ingenuity of the designer. Two main approaches are used to guide the structuring of an overall process control system. These are known as the bottom-up approach and the top-down approach.

The bottom-up approach is used most often in practice. It begins with the choice of individual output variables to be measured and controlled and the choice of control inputs. Simple, standard control configurations are then used as building blocks. These are introduced in various combinations until a control system that meets the control objectives is obtained. A detailed exposition of this approach is given by Stephanopoulos (1984). The individual local controllers rely on the principles of single-variable feedback and feedforward.

The control configuration shown in Figure 14.2 is referred to as feedback control. The principle of this approach is to reduce the effect of a disturbance input by first measuring its effect on a process output and then calculating the necessary correcting input. It is necessary for the disturbances to enter the local system in a well-defined way for this to work. It is also necessary to be able to adjust a control input that enters the system in the neighbourhood of the disturbance input.

Control based on the feedback principle has the major advantage that it incorporates an inherent self-correcting action and that it is not necessary to have detailed information about the characteristics of the process, provided that a high gain can be used in the loop. For the same reason it can

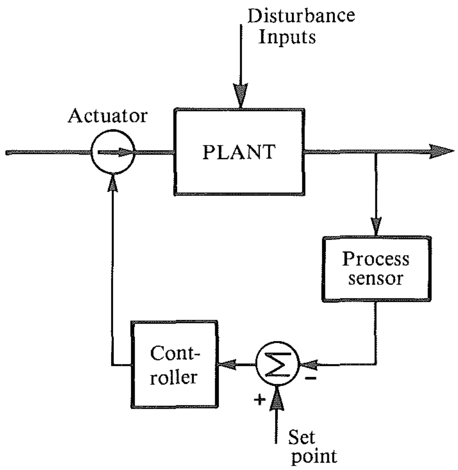


Figure 14.2. The principle of feedback control for reduction of the effect of a local disturbance input.

deal with disturbance inputs even though these may not be measurable.

The basic feedback control configuration is by far the most important and widely used building block in process control systems designed according to the bottom-up approach. Many common industrial processes behave as combinations of capacity and pure delay and can tolerate a high loop gain (Dorf, 1986; Stephanopoulos, 1984). A simple standard method of calculating the correcting control input can therefore be used in most cases. This has led to the standard three-term controller which is discussed further in Section 14.3.4.

Situations commonly occur in the bottom-up approach where standard multiple control loop configurations can give improved control. These include the cascade control configuration, multiple actuator (i.e. control input) control systems and the selective control configuration. These are treated thoroughly by Shinskey (1979). An example illustrating the use of two cascade control configurations is the ball mill grinding circuit control scheme shown in Figure 14.3. This scheme is based on maintaining constant cyclone feed water:solids ratio.

If the disturbance inputs can be directly measured, the alternative control configuration shown in Figure 14.4 can be used. This is referred to as feedforward control. A control input that attempts to counteract the effect of the disturbance is generated from the measured disturbance input. This presupposes that a reasonably accurate model of the characteristics of the process is available for calculating the necessary control input. For this reason, feedforward control is not as robust as feedback control and is therefore usually used in conjunction with a feedback control loop in order to improve its robustness to unmodelled effects.

The top-down approach to control system structuring begins with a

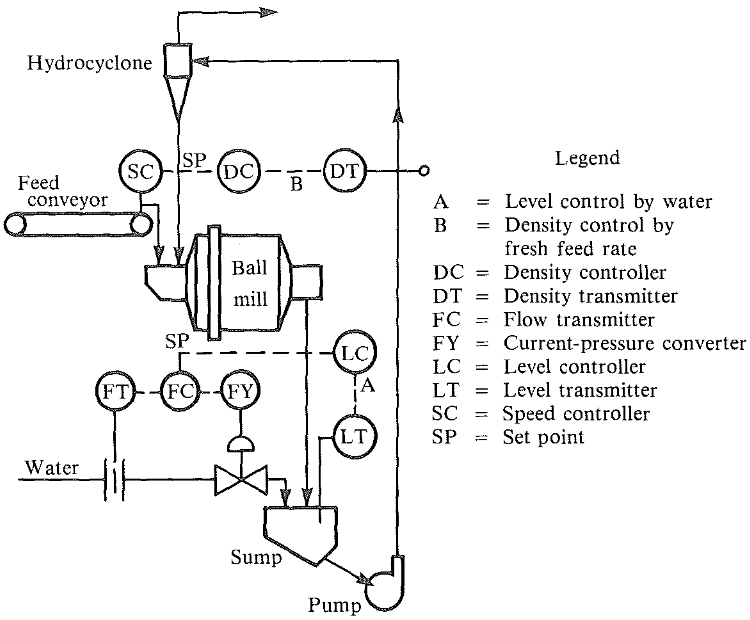


Figure 14.3. Ball mill grinding circuit control scheme based on maintaining constant cyclone feed density.

careful analysis of the control objectives. This leads to the selection of the control principles to be implemented. In a grinding circuit, for example, typical control principles could be cyclone inlet solids flow control, cyclone underflow density control or mill power maximum-seeking control. Based on each selected control principle, the variables that should be controlled are identified. These variables may not be measurable and alternative measured outputs that are closely related to the variables given by the control principle must be chosen. Suitable control inputs must also be chosen.

The next step is to group related outputs and the control inputs that affect them into small sub-systems such that there are only weak couplings between the sub-systems. Unfortunately there are no general rules to guide this grouping process and it must be done by trial and error, possibly guided by a detailed analysis or simulation of process behaviour. A suitable controller must then be designed for each sub-system to maintain its outputs at their set-points. Often the same standard single-variable feedback and feedforward control configurations discussed above are adequate for this purpose, but in other cases, due to strong interactions between variables within a sub-system, it is necessary to resort to the use of a more sophisticated multivariable controller. Good examples of the top-down approach applied to grinding circuit control which do lead to the need for multi-variable controllers are available in the reports by Hulbert and Braae (1981) and Hulbert (1983). A number of techniques such as the Inverse Nyquist Array method,

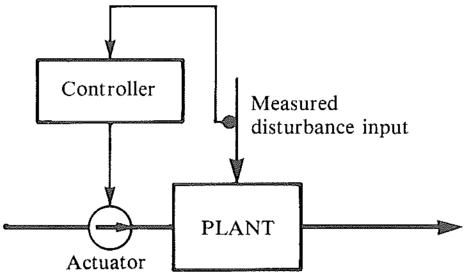


Figure 14.4. The principle of feedforward control.

the Dyadic Expansion method and the Characteristic Locus method are available for carrying out the detailed design of multi-variable controllers. Owens (1978) provides a good treatment of these methods. All these design techniques lead to an essentially identical final controller implementation.

Another excellent example of control system design according to the top-down approach is provided by Pauw and co-workers (1985). In this work, a clear identification of the control objective leads to a control system structure based on a combination of single-variable feedback and feedforward controllers, and logic control.

14.3.4 The standard three-term industrial controller

As stated in Section 14.3.3, a standard generalized form of automatic controller meets the majority of industrial requirements. The application of such a controller is illustrated in Figure 14.5. These controllers are generally designed as modules suitable for panel mounting and provide displays of the measured process output variable and of the controller output to the actuator as well as a convenient way of supplying the set-point. They are available from a large number of suppliers. The same standard generalized control module is also commonly implemented as a program in process control computer systems. Here, flexible colour video displays of measured process output and controller output as well as other controller parameters are usually provided.

Internally in the block shown as G_c these controllers carry out signal processing according to the following equation:

$$m(t) = K_c\, e(t) + \frac{1}{T_i}\int e(t)\, dt + T_d \frac{de(t)}{dt} \tag{14.1}$$

where

$m(t)$ = controller output,
$e(t)$ = error between set-point and measured process output,
K_c = proportional gain (reciprocal $100/K_c$ is known as proportional band),
T_i = integral constant (alternatively known as reset time),
T_d = derivative constant (alternatively known as derivative time).

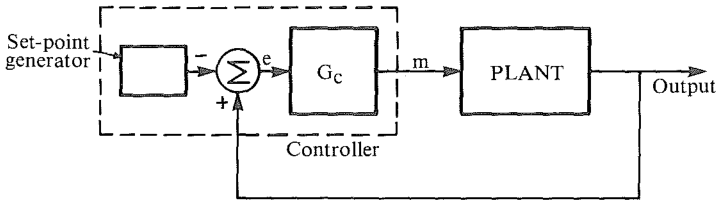


Figure 14.5. The standard industrial controller.

Because of the form of Equation (14.1), this type of control is known as proportional, integral, derivative or PID control, or alternatively as three-term control. The constants K_c, T_i and T_d can be adjusted easily. The operation of determining suitable values for these constants is known as controller tuning. Often only the first two terms are used in practice, the controller then being called a PI controller.

Full discussions of the standard three-term controller, its dynamic performance when used to control various common types of process and the significance of each of the terms are available in standard texts (Stephanopoulos, 1984; Shinskey, 1979).

Methods for tuning this kind of controller based on experimental tests are of particular importance because they can be effectively applied on site. Ziegler and Nichols (1942) originally introduced the idea of optimal settings for the adjustable parameters of conventional PI and PID controllers. These settings could easily be calculated using given formulae once an approximate model of process behaviour had been found by means of simple on-line tests. The Ziegler and Nichols approach has subsequently been refined and improved and has been widely used in practice. Modern versions (Fertik, 1975) provide detailed practical guidelines on how to tune a standard controller. In effect, the Ziegler and Nichols approach provides a direct way of incorporating a simple model of the dynamic behaviour of the process into a PID controller.

14.4 Process Measurements

14.4.1 Introduction

Industrial instrumentation is a vast subject, and emphasis in this section will therefore be placed only on those techniques that are most important in gold ore processing. Adequate treatment of general industrial instrumentation is available in standard texts (Considine, 1985; Doebelin, 1983; Bentley, 1983).

Before discussing specific measurement techniques, it is worth re-emphasizing the special practical problems that are encountered in ore processing plants due to the aggressive nature of the process streams and the harsh plant environment. In the gold industry, it is necessary to measure properties such as the flow rate of solids on conveyor belts, the flow rate and density of dirty liquids and pulps in pipes and launders, the particle size and size distribution of mill products in pulps, the flow rate and concentration

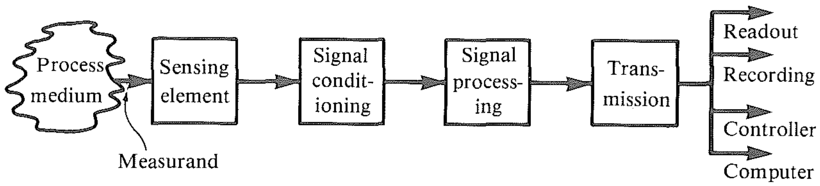


Figure 14.6. General structure of a measurement system.

of reagents in pulps and other chemical and physical properties of components that occur only in very small concentrations. Many of these properties can only be measured indirectly by employing appropriate primary and secondary sensing elements in series. Moreover, many standard industrial instruments developed for use in the chemical, petrochemical and similar industries are suitable only for relatively clean environments and cannot withstand abrasive pulps.

It is also vital to appreciate that successful and reliable operation of an industrial instrument depends upon all of the components associated with the measurement system. These include the process stream sampling devices, the sensing element, the signal transmission cable, the data presentation device, the auxilliary electricity, air and water supplies and the trained personnel and equipment needed to maintain and calibrate the measurement system. When selecting an instrument it is therefore very important to consider the total measurement system life-cycle costs, especially maintenance and spares-holding costs, in addition to initial capital costs.

The growing importance of process control and the increasing use of process control computers (which are less forgiving of measurement errors than a human operator) place even greater demands on the reliability and repeatability of instrumentation. In certain cases, very accurate data will also become an increasingly important requirement. Successful industrial measurement requires a thorough understanding of the basic principles of measurement and careful attention to all the components of the measurement system.

14.4.2 Terminology of instrumentation systems

Figure 14.6 shows the functional elements found in typical industrial measurement systems. The physical property to be measured is known as the measurand. The primary sensing element interacts with the process medium and produces an output that varies as some known function of the measurand. This output variable is often a physical property that is not in a form suitable for further processing or transmission. It is the function of the signal conditioning element or transducer to convert the output of the primary sensing element into a more suitable form such as a voltage signal, while preserving the information contained in the original output of the primary sensing element. The resulting signal may then require subsequent signal processing, for example amplification, filtering and analogue to digital conversion, in order to make it suitable for transmission.

The equipment to carry out the above three functions is usually housed in one or more enclosures located close to the point of measurement in the plant; it is commonly known as the instrument, field transmitter or sensor.

The remainder of the measurement system consists of the transmission medium (usually copper wires) to carry the signal to a convenient point such as a central control room and possible further signal processing in the control room. Thereafter the signal is available for uses such as driving read-out equipment, recording, providing the feedback signal for a controller or input to a process control computer.

14.4.3 The primary sensing element

The primary sensing element is the most critical part of a measurement system. The chosen physical or chemical principle of operation is the dominant factor in determining the performance characteristics of the whole measurement system. The materials of construction which can be used for the primary sensing element determine the ability of the instrument to withstand the process medium and the environment. In practice, the primary sensing element is always unintentionally sensitive to physical properties other than the desired measurand. These are called interfering inputs and they can, of course, vary as a function of time and thereby cause the output of the primary sensing element to vary erroneously. The user of the output of the measurement system is usually unaware of the error. Temperature is often the most troublesome interfering input.

All the other functional elements of a measurement system are also affected by interfering inputs. Adherence to good instrument engineering practice can, however, usually reduce the effect of these other interfering inputs to negligible proportions.

14.4.4 Accuracy and calibration

In process instrumentation, the accuracy of a measurement is the closeness of the displayed or output value to the true value of the measurand. It is quantified in terms of the measurement error or inaccuracy, that is, the difference between the measured value and the true value. The true value of a variable is defined as the measured value given by a standard instrument of ultimate accuracy.

In spite of this seemingly clear definition, the question of accuracy in process instrumentation is ambiguous and confused and the user should be very careful when interpreting accuracy specifications. Part of the problem is that the so-called true value is generally unknown except in certain specialized laboratories. The problem is also compounded by the influence of the environment and all the measurement system components (primary sensing element, circuitry and read-out equipment) on total measurement system accuracy. The accuracy of an instrument depends on the environment in which it must operate.

A proper accuracy specification first states the allowable ranges for all the significant interfering inputs and then gives an estimate for the maximum total error or inaccuracy. This is usually quoted as a percentage figure based on the full-scale reading of the instrument. Another method sometimes used gives the error as a percentage of the particular reading with a qualifying statement which applies to the low end of the scale. Note that an accuracy specification does not give the actual error for a particular measurement, but rather gives worst-case bounds for its error.

Repeatability is the ability of the measurement system to give the same output for the same input, repeatedly applied to it. The non-repeatability error is only one of the components contributing towards the total error included in the accuracy specification. Good repeatability is sometimes all that is necessary when the measurement is to be used for process control purposes. Note that repeatability is closely related to the statistical concept of precision discussed in Chapter 13.

The static characteristics of a measurement system element can be found experimentally by holding all the inputs except one at known constant values. The one input under study is then varied through a range of known constant values and the corresponding element outputs are noted. The input/output relationship obtained in this way comprises a static calibration valid under the stated constant conditions of all the other inputs. In order to describe the overall static behaviour of the instrument, this calibration procedure is carried out for the measurand or desired input and for all significant interfering inputs as well. The calibration procedure may include the operation of adjusting the instrument in order to reduce certain error components.

The static calibration procedure described above clearly refers to an ideal situation which can only be approached, but never reached, in practice. It is particularly difficult to ensure that all interfering inputs are held constant and to achieve the required known constant input values. Usually it is only the instrument manufacturer who performs such comprehensive tests. The user of the instrument should always be aware of errors introduced by differences between the ideal situation and the real plant environment. It is also important to note that calibration is a static test and that in practice the measurand varies dynamically. However, ore treatment plant applications generally involve the measurement of quantities that vary only quite slowly and the dynamic response of instrumentation systems is therefore seldom a serious limitation.

14.4.5 Basic physical measurements

14.4.5.1 *Flow rate of solids*

This important measurement has been dealt with under mass measurement in Chapter 13 (Section 13.7).

14.4.5.2 *Flow rate of liquids*

Measurement of the flow rate of a liquid in a pipe is straightforward in many industries; simple primary sensing elements such as the orifice plate are widely used (Considine, 1985; ISO 5167, 1980). In ore processing, however, the most important flowing media to be measured are pulps and water containing significant amounts of solids in suspension. These materials cause major problems with many of the standard industrial liquid flow measuring devices

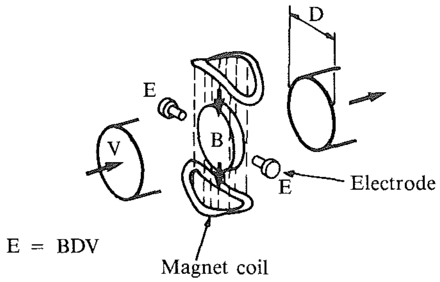


Figure 14.7. Principle of the electro-magnetic flow meter.

due to erosion and blocking. For this reason, the electro-magnetic flow meter, which causes essentially no obstruction to flow, is the most successful and widely used flow measuring device. This has a typical accuracy of $\pm 1\%$ of indicated flow, and measurement is independent of changes in viscosity, pressure, density and temperature.

The principle of the electro-magnetic flow meter is Faraday's Law of induced e.m.f. as shown in Figure 14.7. A magnetic field is set up across a pipe and a fluid flowing at average velocity, V, cuts the magnetic flux of flux density, B, at right angles. The induced e.m.f. is normal to both V and B and is detected by a pair of electrodes on the inner surface of the pipe as shown. The magnitude of the e.m.f. is given by the equation

$$E = BDV$$

where D is the diameter of the pipe.

The electro-magnetic flow meter is therefore actually a velocity measuring device. Assuming a full pipe and constant pipe diameter, the flow velocity can however easily be converted to a volume flow rate. It is, of course, necessary for the liquid to be conductive. However, with high-impedance electronic amplifiers, very low conductivities are acceptable and this is seldom a practical limitation.

A.C. electro-magnetic flow meters use a 50 Hz alternating magnetic field derived from the standard a.c. line current. This eliminates polarization effects (formation of an insulating layer of gas) at the electrodes and also permits the use of high-gain a.c. amplifiers which have far less drift than comparable d.c. amplifiers. A.C. electro-magnetic flow meters have the advantage of a relatively low cost and are very widely used. A disadvantage with a.c. systems, however, is that the strong alternating magnetic field induces spurious e.m.fs. into the sensitive measuring circuitry and the amount of interference is variable and changes with time. The so-called d.c. electro-magnetic flow meter has been introduced to overcome this problem for applications requiring high accuracy. These actually use a magnetic field that is switched on and off at a low frequency (usually around 10 Hz). When the field is switched off, any residual instrument output is considered to be an error and is stored and then subsequently subtracted from the total output obtained when the field is next switched on again. This provides a continuous, automatic zeroing effect that largely overcomes the problem of spurious induced e.m.fs. An additional advantage of d.c. systems is that they consume far less power than equivalent a.c. systems. Problems have however occurred when using d.c. systems with certain mining pulps. In these cases a severe spurious noise signal is induced across the electrodes. Expert advice should be sought in doubtful cases.

Certain basic guidelines apply to all liquid flow meter installations. The flow meter should ideally be installed in a vertical pipe section with an upwards flow direction to ensure that the pipe remains full. There should be no bends or other restrictions immediately upstream of the flow meter. Finally, for electro-magnetic flow meters it is important to have a good earth connection between the instrument casing and the measured liquid.

14.4.5.3 *Liquid level in a vessel*

It is often necessary to measure the height or level of the surface of the liquid in some vessel such as a tank or sump. A number of standard industrial level measurement techniques operate successfully in extraction plants. These include air bubblers in which the back-pressure of air in a dip tube is proportional to the head of liquid (assuming constant liquid density), various arrangements using floats or bobs suspended from wires or cables which are attached to a displacement sensing unit mounted above the vessel, electrical capacitance or conductivity probes and measurement of the differential pressure between the bottom of the vessel and the gas above the liquid surface.

Bubbler tube and differential pressure methods are simple and reasonably reliable, but have the disadvantage of being sensitive to the liquid density and subject to erosion, while floats and capacitance probes tend to be subject to fouling and need regular maintenance.

In addition to the above methods, level measurement instruments using ultrasound have recently become popular in extraction plants. These operate on the principle of electronically measuring the time taken for the echo of a pulse of radiated ultrasonic energy to return from the liquid surface to a transceiver. No moving parts are involved and it is not necessary to have any probes or sensing elements in contact with the liquid (a non-invasive technique).

One source of error in ultrasonic level measurement is the dependence of the speed of sound on the temperature of the medium through which it travels (air temperature). Also, the surface being measured must reflect sufficient sound energy. Froth or bubbles on the liquid surface can cause problems due to sound absorption. It is also important to ensure that the ultrasonic transceiver is installed on a vibration-free mounting.

Finally, it can be mentioned that this technique has also been successfully used to measure the height of the liquid/solid interface in thickeners where the solids form a mud layer below the water surface. Here the sound pulses are radiated downwards through the water to the mud line and are then reflected back to the transceiver.

14.4.5.4 *The water content of a pulp*

The water:solid ratio (w/s) of a suspension of ground solids in a flowing water stream is a very important variable for process control in a gold extraction plant. This is measured by sensing the average density of the pulp stream as it flows through a pipe. The nuclear density meter is the standard device for performing this measurement.

The operation of the nuclear density meter relies on the fact that the gamma rays produced by a radioactive source arrive at a detector with an intensity that is inversely proportional to the density and thickness of the material through which the radiation passes. The radioactive source, usually caesium-137, is located on one side of a pipe and a radiation detector is placed on the opposite side. With suitable amplification and correction for radiation attenuation by the walls of the pipe and assuming a constant internal pipe diameter, an electrical signal proportional to the instantaneous average density of the flowing pulp can be produced from the radiation detector output. This measurement technique is simple, non-invasive and has been found to be reliable in practice. The ionization chamber is the most satisfactory kind of radiation detector in these applications.

Important sources of error in nuclear density measurement include the presence of entrained air bubbles in the flowing pulp, the build-up of material on the inside walls of the pipe and the natural exponential decay of the radioactive source. The density of air is very much lower than that of pulp, and air bubbles therefore can cause a significant measurement error. This problem is particularly severe when the nuclear density meter is installed close to the delivery of a pump that is sucking in air due to air entrainment in its feed or due to poor seals or is experiencing cavitation. These points should be carefully checked. Problems due to material build-up can be reduced by installing the density meter in a vertical pipe section with an upward flow direction. Finally, compensation for errors due to radioactive source decay can be included in the density meter electronics. This is particularly important if small density changes are to be measured. With dilute pulps (w/s ratio > 10) the sensitivity to the above errors becomes worse due to the non-linear relationship between density and w/s ratio; the precision of density measurement therefore becomes poor. The calibration of nuclear density meters is discussed in Chapter 13 (Section 13.7.3).

14.4.5.5 *On-line particle size measurement*

A key parameter in assessing the performance of a milling circuit is the particle size distribution of a cyclone overflow. This is measured by a screen analysis of a representative sample for record purposes. For dynamic control, two automated measurement methods are available, namely, the Armco Autometrics Particle Size Monitor (PSM) and the Leeds and Northrup Microtrac analyser.

The Armco Autometrics PSM

The principle of the Armco Autometrics PSM is the measurement of the attenuation of ultrasonic waves as they pass through a sample of the pulp. From this attenuation, the pulp water:solids ratio and the percentage of

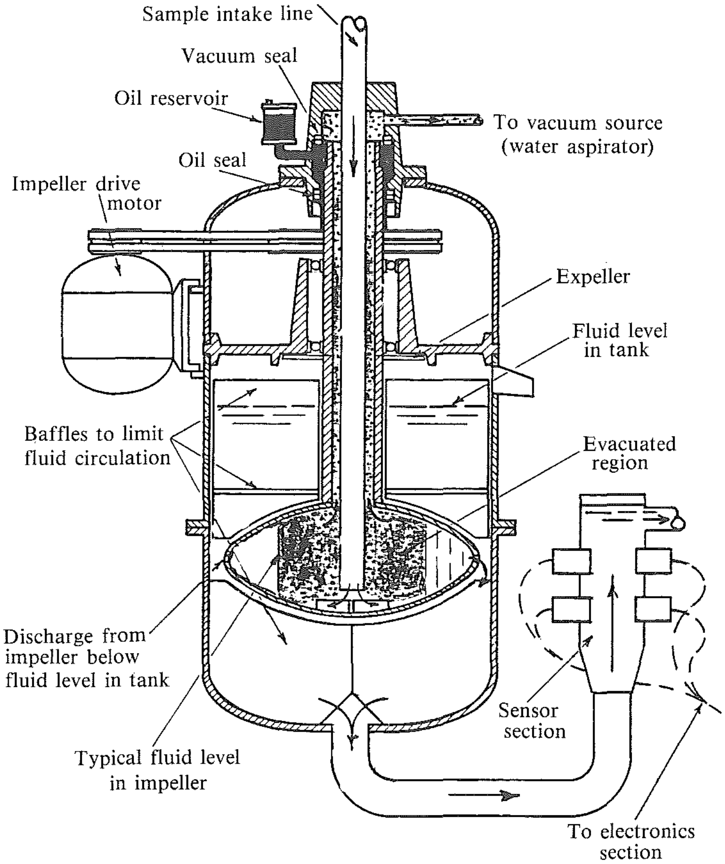


Figure 14.8. Armco Autometrics PSM - air eliminator.

material passing a particular size fraction are calculated. The instrument comprises three sections: the air eliminator, the sensor section and the electronics section.

The air eliminator draws a sample from the process pulp stream and removes entrained air bubbles. The de-aerated pulp then passes between the sensors where particle size and w/s ratio are sensed by ultrasonic means and the pulp returns to the main stream below the sampling point.

The air eliminator is shown in Figure 14.8. This shows the impeller section where air is removed and the sample cell where the two pairs of ultrasonic transducers are situated. One pair is set at an ultrasonic frequency which is not sensitive to particle size and the second set at a frequency that is affected by the particle size range selected for the particular instrument.

The instrument requires between 55 and 75 litres per minute of pulp extracted from a position within the process stream where there is sufficient turbulence to ensure a representative sample. Figure 14.9 shows a diagram

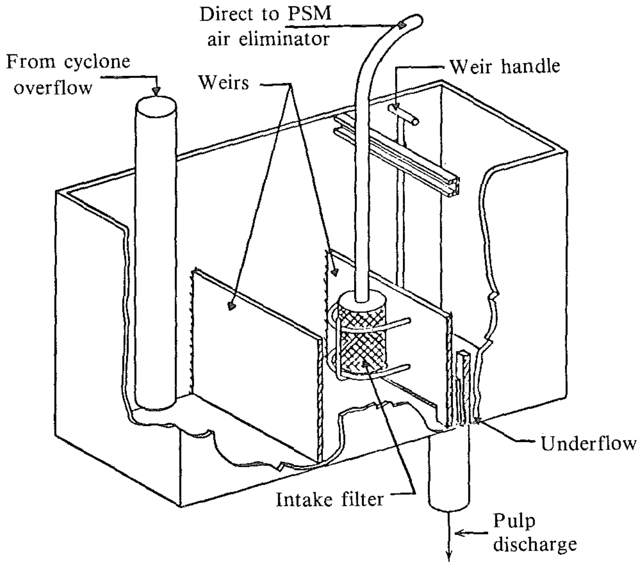


Figure 14.9. Armco Autometrics PSM – sampling system.

of a typical sampling box. It is possible to use the same instrument for more than one stream by multiplexing the sample extraction step. However, the time taken for the air eliminator and electronics to process and stabilize a measurement is lengthy. To sample one stream, switch to a second and then return to the first takes about 18 minutes.

The instrument is initially pre-calibrated by the manufacturer and then calibrated on site by taking samples from the discharge pipe of the PSM and relating the readings of the instrument at the time of taking the samples to a screen analysis and pulp w/s ratio measured later in the laboratory. The manufacturers claim an accuracy of $\pm 1,5\%$ to 2% of actual cumulative per cent passing a given mesh size and $\pm 2\%$ on the w/s ratio.

There are 27 operating Armco Autometrics PSMs in South Africa in the gold and platinum mines. They are particularly important in mill control strategies.

The Leeds and Northrup Microtrac Analyser

The Leeds and Northrup Microtrac measures the scatter from a laser beam projected through a liquid stream (Figure 14.10). Scatter angle is a function of particle size. Small particles scatter light at larger angles than large particles. The light scattered by the particles is passed through an optical filter, and its intensity and angle is detected and transmitted to a microprocessor which calculates the size and the quantity distribution of particles in the liquid. In operation, a sample of pulp is introduced into the instrument where it is diluted with water and then recirculated through the measurement cell where it is illuminated by the laser beam.

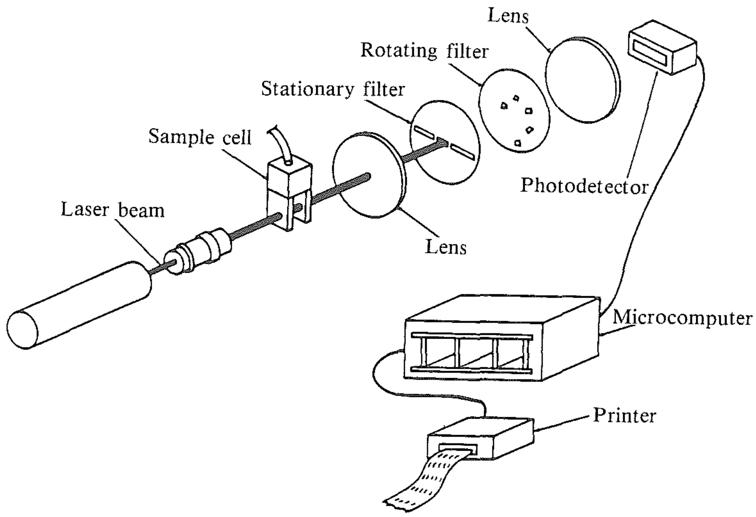


Figure 14.10. L & N Microtrac optical measurement system.

The calibration of this instrument is fixed by the optical filter design and the microprocessor software. There is automatic background compensation provided by a set zero control. In operation, continuous sub-samples of about 150 litres per minute are taken from the cyclone overflow using fixed launder or pipe samplers. These samples are screened to remove wood fibre and the pulp is passed through a 40 mm pipe from which an Isolok valve takes as many 3 ml samples as required. The Microtrac is able to give a complete particle size distribution every two minutes.

The Microtrac has been used in South Africa mainly as an off-line instrument but it is planned to use it in mill control strategies. By suitable design of the sampling system, it is possible to use a single Microtrac to measure the particle size distribution from several mills. The penalty would be a longer delay time between measurements on an individual mill. Whether this can be tolerated depends on the particular control strategy being employed.

14.4.5.6 *Basic chemical measurements*

A number of fundamental chemical measurements are required for the control of an ore treatment plant. Examples include the measurement of pH, oxidation-reduction potential, the degree of alkalinity of a pulp, the oxygen content and the cyanide concentration. The usual laboratory methods employed for determining these quantities involve removing a sample of the pulp, filtering it to separate the solids and then applying normal analytical chemical methods for the particular determination. For example, cyanide concentration in solution is determined by adding potassium iodide to a known volume of clear solution and then titrating with silver nitrate. These methods can be obtained from Lenahan and Murray-Smith (1986).

Standard industrial instruments are available for some measurements such as pH and oxidation-reduction potential (Considine, 1985). There have also been recent developments in automating certain of the traditional procedures to obtain rapid on-line measurements. Some of these are dealt with below:

The On-Line Gold Analyser (OLGA)

The first on-line gold analyser, known as the Tell-tale, was developed by Corner House Laboratories during the period 1965 – 1969. This design is still successfully used in a number of plants.

The AAC OLGA Mark III Gold Analyser incorporates many features of the original Tell-tale and is designed specifically to determine gold concentration in the barren solutions resulting from the zinc precipitation and filtration step in the production of gold by the Crowe-Merrill process (Brandt and co-workers, 1980).

The analyser system is situated in a small room centrally sited within the filter section of the plant. A maximum of 9 process streams in addition to the composite are continuously sampled from the filter plant at a rate of approximately 1 litre per minute. Each process stream flows into a constant-head sampling vessel that provides an analyser feed at a rate of 50 ml per minute. The analyser feed-lines from the sampling vessel are furnished with control valves that are actuated in a programmed sequence by a microcomputer.

The basis of the automatic analysis technique is preconcentration of the trace levels of gold in the barren solution by continuous solvent extraction, followed by flame atomic absorption spectrophotometric (AAS) determination of the gold concentration in the organic extract. The solvent extractant employed is a solution of 1% Aliquat 336 (a proprietary extractant) in di-isobutylketone (DIBK). The continuously extracted gold, preconcentrated 40 times, is collected and automatically analysed by the AAS spectrophotometer.

The instrument can measure gold in the concentration range 0,005 to 0,500 mg/l. In operation, the gold concentration in the composite solution is measured every three minutes. If the concentration exceeds a preset level, an alarm is sounded and the instrument then measures the effluent from each of the filters until the individual filter that is malfunctioning is identified. Figure 14.11 shows the OLGA schematically.

Automatic on-line analysis using AAS with electrothermal atomization

As an alternative to continuous solvent extraction, the use of AAS coupled with electrothermal atomization has been developed for measuring gold in solution. The basis for this method is the graphite tube atomizer, on to which between 5 and 50 microlitres of solution are injected automatically. The furnace goes through a heating cycle, controlled by a microprocessor, ending up with a temperature in excess of 2 000°C to atomize the gold for determination with the AAS.

For automatic on-line analysis, solutions from the process are delivered continuously to the room where the AAS is sited. Sub-samples are extracted

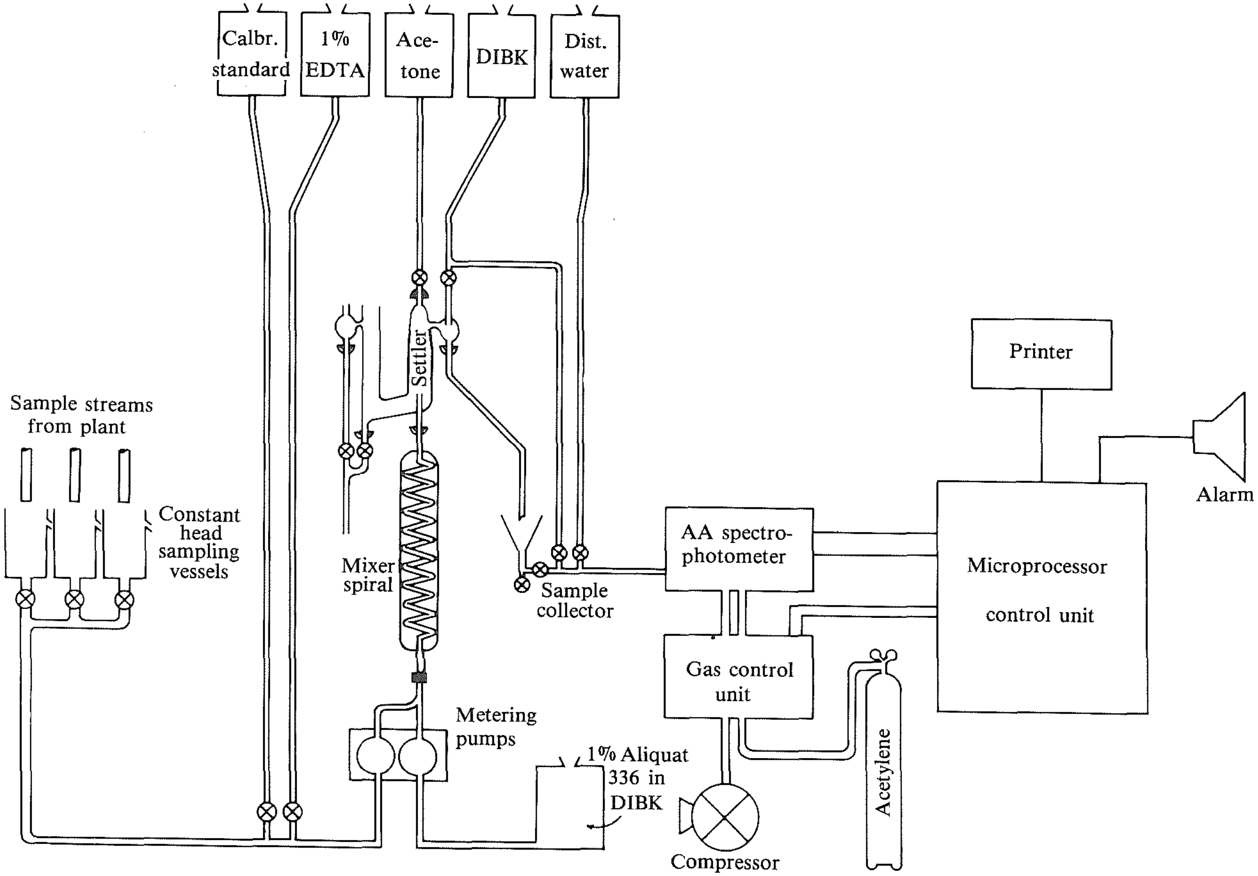


Figure 14.11. Schematic diagram of AAC Mark III on-line gold analyser.

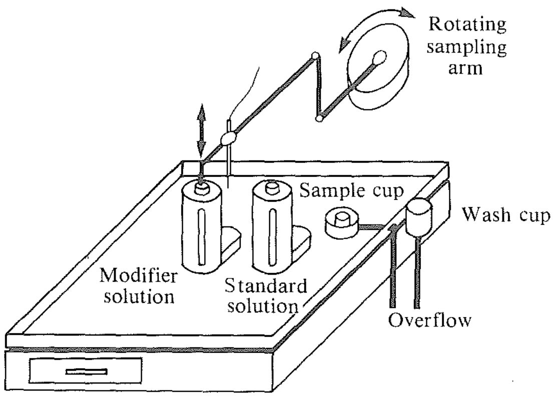


Figure 14.12. Sampling dispenser.

from the main sample lines and fed to a position where the sample dispenser can extract an appropriate aliquot. The sample dispenser feeds samples of the solution to be analysed together with modifiers required for the AAS analysis to the graphite tube atomizer. The dispenser is also programmed to take standard solutions for calibration of the equipment. Figure 14.12 is a diagram of a typical sample dispensing arrangement as used by Robért and Ormrod (1985).

Mintek have developed this technique for measuring gold in solution in CIP plants. Solutions are extracted from the absorption tanks using a filter tube immersed in the tank. Solution permeates through the filter tube under the hydrostatic head between the surface of the slurry and the tube which is immersed to a depth of about 1 metre (Figure 14.13). On a timed sequence, the probe is pressurized by nitrogen and a sample is transported to the analysis room for analysis on the AAS as described above.

Alternatively, sampling probes have been used in which the solution is drawn into the filter probe by vacuum and then transported to the AAS room by compressed air. This technique has been incorporated in an electrothermal atomization AAS system, developed by the J.C.I. Minerals Processing Research Laboratory, for the on-line monitoring of soluble gold loss from a conventional filter plant. A microcomputer controls the filtration operation and the sequencing of streams being monitored. It also performs the regressions, prints out results and triggers alarms when selected concentration levels are exceeded. This AAS system has been further developed by J.C.I. for the CIP process to cover the range 0,002 ppm to 70 ppm gold in solution. This is achieved by exploiting the features offered by the furnace system used. Eight furnace programmes and three calibration curves are used to obtain the dynamic range, all of which are selected by the microcomputer. The most suitable curve is automatically selected, depending upon the absorbence obtained for the solution being measured. This system monitors each absorption tank and, by using the higher calibration range, information can also be obtained on parts of the elution cycle.

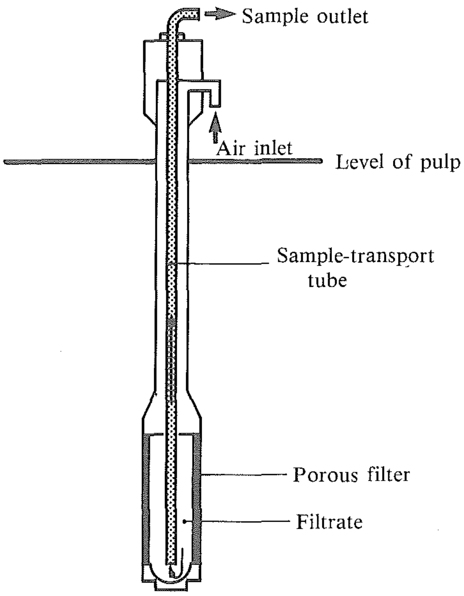


Figure 14.13. Sampling probe.

Automatic spectrophotometric analysis

There are on the market instruments which allow for continuous colorimetric determinations. With these instruments, a peristaltic pump takes up a measured volume of a solution for analysis which can then be mixed with known volumes of reagents taken with the same peristaltic metering pump but with different tubes. In this way coloured solutions are produced and these are passed through an optical measurement cell through which light passes. The absorption of the light gives a measure of the concentration of the particular element in the original solution. This technique has found wide application in many industries and has lately been applied to the metallurgical industry.

One particular application is the Gold Fields cyanide monitor. In this instrument a sample of solution is extracted from a leach tank using a filter tube sample extractor. This consists of a permeable filter tube about 500 mm long immersed in the pulp. A vacuum is applied inside the filter tube to draw in solution. After about 5 minutes, the vacuum is removed and compressed air fed into the tube to transport the sample to the analyser room and to blow the cake off the filter tube. The solution samples are piped to the outside of the analysis room and sub-samples led through the wall into overflow sampling pots.

The auto-analyser is programmed to take an aliquot from the sub-sampling pots, and then to take measured volumes of copper sulphate and EDTA to mix with the cyanide solution. Cyanide reacts with copper sulphate to produce a colourless solution. However, copper sulphate is always added in excess and therefore cyanide concentration is inversely related to the copper sulphate concentration which is determined as the solution flows through the optical cell.

On-line analysis of cyanide in solution
Automatic cyanide monitors have been developed for use in the control of cyanide addition to leach circuits. These determine the free cyanide content of leach pulps by automatic titrimetric analysis. A solution sample is extracted from the leach tank by means of a vacuum filter tube. In one instrument (the AAC Cycad monitor), the sample is then blown by compressed air to a cell fitted with a combination gold-thallium electrode and titrated with 0,1 N silver nitrate until an endpoint is reached at 700 – 750 mV. An output in the range 4 – 20 mA is then produced. The filtration/titration cycle takes approximately 3 minutes. A rinse cycle is performed periodically. This uses 5% sulphamic acid to remove calcium carbonate and sulphate scaling.

Another instrument which gives an adequate measurement for control purposes transfers the sample by means of a metering pump. This sample is mixed with a metered equivalent of standard silver nitrate and the mixture passes through a cell fitted with a gold-thallium electrode so that a continuously varying e.m.f. is produced corresponding to the sensitive section of the titration curve. Suitable signal processing then gives an output in the range 4 – 20 mA.

Lime and pH measurement
A logical extension to the above measurement technique is the sequential determination of both free cyanide and protective alkalinity on the same measured volume of pulp filtrate. Work on this extension is in progress. After the alkalinity due to cyanide has been neutralized by the addition of one equivalent of silver nitrate, the remaining, so-called protective alkalinity is determined by titration with 0,1 to 0,5 N sulphuric acid, using pH electrodes to achieve an endpoint at pH 8,3.

14.4.5.7 *Carbon and resin concentration meter*
This device being developed at Mintek is based on the phenomenon that ultrasonic waves are attenuated by particulate matter as described for the Armco Autometrics particle size monitor. However, this device uses a robustly constructed probe which can be lowered into the pulp. By transmitting and receiving ultrasound at two frequencies, the concentration of carbon or resin can be measured against the constant concentration of the pulp. Normally the attenuation of the lower wavelength is constant and due to the solid particles of the pulp, whereas attenuation of the high frequencies is due to the resin or the carbon and varies with their concentration. The probes use piezoelectric ceramic transducers encapsulated in epoxy resin. Figure 14.14 shows the plan of a typical probe.

14.4.5.8 *Carbon activity meter*
Different batches of activated carbon, whether new or reactivated, can have different gold-loading capacities. The activity can be determined simply by

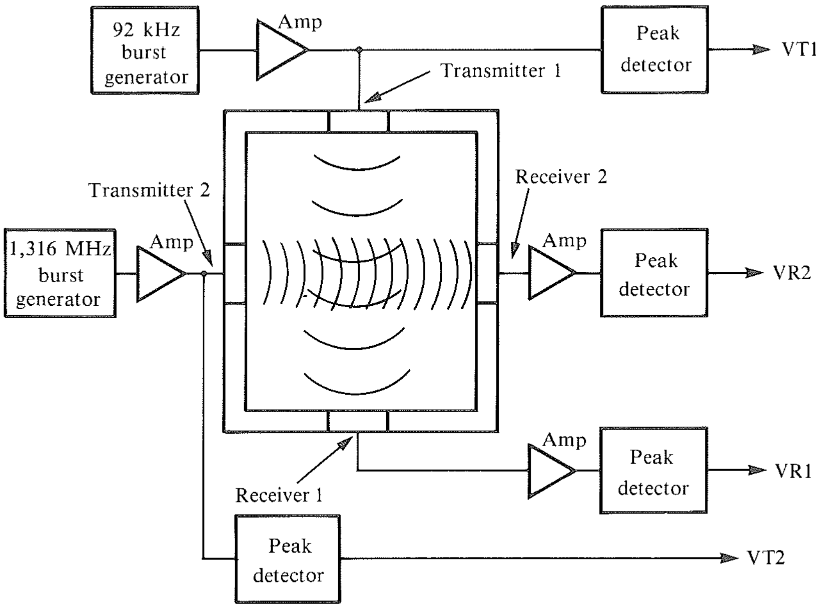


Figure 14.14. The Mintek carbon and resin concentration meter.

the contacting of a known amount of carbon with a known concentration of gold in the presence of sodium hydroxide and sodium cyanide (simulated plant conditions) and the periodic removal of an aliquot portion of the solution for the determination of its gold content by atomic absorption spectrometry. The rate of the gold absorption and the final gold capacity can be obtained from a plot of the gold concentration remaining in solution versus time. This approach yields excellent results and is straightforward, but does require laboratory facilities such as an atomic absorption spectrophotometer and, more important still, a person to perform the analysis over a period of 3 to 4 hours. However, this approach is not favoured on most plants because it means that the sample must be removed from the reactivation kiln and transported to a laboratory for assay, all of which can lead to long delays before the results of the analysis are obtained.

A small and relatively simple instrument has been developed at Mintek. This instrument, which is microprocessor-based and extremely easily operated, is designed for use on the plant in close proximity to the reactivation kilns and is based on the electro-chemical technique known as anodic-stripping voltammetry.

A 500 ml portion of a stock solution containing sodium hydroxide (0,01 mol/dm^3) sodium cyanide (0,02 mol/dm^3), gold (100 μg/g) and a small amount of either lead or thallium nitrate (1 to 4 μg/g) is added to a beaker. The electrode probe of the instrument, which consists essentially of a platinum electrode and a gold reference electrode, is positioned, and the stirrer is set in motion.

When the readings on the analyser have stabilized, 1 g of carbon is added and the instrument is allowed to run automatically. The gold concentration is measured every 3 minutes and plotted on a chart recorder. Operation of the instrument requires no special expertise; the operator needs to carry out only a few mechanical operations, e.g. measuring out a volume of stock gold solution with a measuring cylinder, weighing 1 g of carbon sample and pressing a button on the instrument to initiate the measuring process.

Besides its main functions of measuring the activity of reactivated carbon, the instrument can also be used to measure the activity of carbon samples removed from various absorption stages in the plant, which is of assistance in studies of possible poisoning effects during absorption.

14.5 Actuators and Final Control Elements

14.5.1 Introduction

Successful process control requires that all key process variables can be changed conveniently. The term controllability is often used in this context. To obtain a process that is controllable, it is first necessary to have a sufficient number of actuators or final control elements. Also, actuators should be selected so that the process variables can be changed over a sufficient range with good resolution. The actuator characteristic should be such that the gain does not change very much over the whole operating range. A common problem in flow systems, for example, is that the control valve is oversized, giving a very non-linear relationship between valve opening and flow. The actuators used in ore treatment plants fall into three main categories: control valves, material feeders and variable speed drives.

14.5.2 Control valves

The main types encountered in ore treatment plants are globe valves, ball valves, diaphragm valves and pinch valves as shown in Figure 14.15. See Considine (1985) for details of other types of control valves. Many valves installed in plants are operated manually, but for process control we are concerned with those versions which can be actuated by pneumatic, hydraulic or electrical means.

Globe and ball valves are suitable only for use on clean or slightly dirty fluids. Small ball or needle valves are often used for reagent flow control duties. Diaphragm valves are the traditional means of manipulating the flow rate of pulp in a pipe line. They have the advantage of simplicity and robustness, but the disadvantage of having a large mass and a high actuator power requirement. Pinch valves consist of a length of rubber pipe which is pinched between bars or, preferably, by a pneumatic or hydraulic sphincter. The rubber pipe has significant hysteresis, and a valve that provides a pull action as well as a pinch action may be required to overcome this. Where the flow rate of an exceptionally abrasive or viscous substance must be manipulated, a proportioning diverter valve operated in on/off mode by a variable mark-to-space ratio pulse signal is often the only solution.

Various linear and non-linear flow versus valve stem position

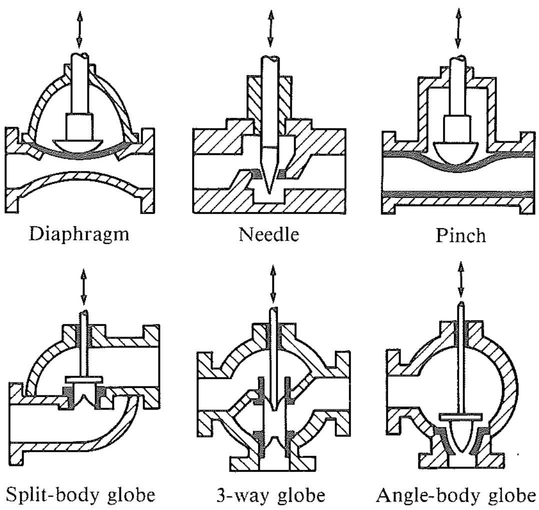


Figure 14.15. Typical control valves.

characteristics are available (Hutchinson and Merwick, 1976), and manufacturers' catalogues should be consulted for details. It should be borne in mind that such characteristics apply for a constant pressure drop across the valve. The characteristic for the valve installed in a line is quite different. For automatic control applications, the valve characteristic should be chosen so that the overall system flow versus valve stem position characteristic is approximately linear.

Actuators for control valves are of three main types, namely, electrical, pneumatic, and electro-hydraulic. Pneumatic actuators have remained the most popular type over many years because they offer the combined advantages of a high power to mass ratio, simplicity, low cost and a well-defined fail-safe action. The electrical output signal from the control centre is initially processed by a current-to-air-pressure transducer which can be either separate or integral with the actuator. Clean compressed air is supplied to this transducer at pressures of up to 700 kPa, and the actuator action on air supply failure can be either to open or to shut the valve, depending on design. Actuators are available to operate valves in both infinitely variable and fully open/fully shut modes. Valve positioners which operate as local position feedback controllers can be attached to pneumatic actuators. These are used when the valve is the final control element in a critical control loop and a strictly linear relationship between electrical current and valve position is required.

Electrical valve actuators are available but are usually powered from the main electricity supply and therefore suffer from the problem that in the event of supply failure, the valve stays put. Electro-hydraulic actuators can overcome this problem by provision of a reserve accumulator to permit a definite failure action. These have the advantage of not requiring an air supply but are relatively expensive due to the high-precision construction required.

14.5.3 Material feeders

The main types used for automatic process control are vibrating feeders and belt feeders. The vibrating feeder is a flexible device and can be used for feeding material ranging from fine powder to run-of-mine rock. It has the advantages of high throughput capability, low cost and low maintenance requirements. Its chief disadvantage, however, is that the actual mass feed rate varies erratically when the material changes through the range from wet fines to dry rock. Attempts to overcome this problem by measuring the actual mass feed rate with a belt weigher and using this as a feedback signal to control the vibrating feeder have not been successful, probably due to the transport delay caused by the distance between the feeder and the belt weigher.

Belt feeders are more expensive and require more maintenance than vibrating feeders, but have the advantage that a belt weighing action can be directly incorporated to give a so-called weighfeeder. These give a precise measurement and control of material mass feed rates.

14.5.4 Variable-speed drives

The two main types encountered use hydraulic control and solid-state electronic control. Hydraulic variable-speed drives are based on a variable fluid coupling which is placed between a conventional a.c. induction motor and its load. These are reliable, robust and are available in a range of power ratings. Versions specially designed for use with pumps are also available.

Electrical variable-speed drives can be subdivided into a.c., d.c. and eddy current types. A.C. drives consist of a solid-state rectifier/inverter which controls the frequency and voltage supplied to a specially rated a.c. induction motor. These are very reliable, but can be more expensive than other types.

D.C. drives have simpler electronics than a.c. drives, but require a d.c. motor. The brushes in such motors have specialized maintenance requirements.

Eddy-current drives comprise a variable magnetic coupling which is placed between a conventional a.c. induction motor and its load. These have very good reliability and are easy to maintain.

14.6 Process Control System Design Philosophies and Architectures

14.6.1 Introduction

Preceding sections of this chapter have discussed certain important components of process monitoring and control systems. In this section, we examine the issue of integrating these components into overall working systems.

One of the most basic tasks of all process control systems is to gather data from the various process instrumentation subsystems and to present this data in a suitable form to the process operating personnel or, in other words, to provide the interface between the process and the operator. Various alarm monitoring and reporting tasks together with the task of implementing automatic safety shut-down logic are also usual basic requirements. Closely connected with basic data presentation are the tasks of storing and subsequently processing the data and presenting information for process study, evaluation and problem investigation purposes. There is a trend towards providing sophisticated process study, development and management tools in modern plants. Another important task of process monitoring and control systems is to implement the various kinds of real-time automatic control referred to earlier in this chapter such as single-variable or multi-variable continuous control, supervisory control, process optimization and logic or sequencing control.

Although the above tasks are relatively standard, there are a number of different ways in which they can be implemented, based on different process operating philosophies and to a certain extent, process design. There are also a number of different technologies available for implementing process control functions.

Some of the most important issues relating to process control system architectures and design philosophies will be discussed under the subsequent headings. Considine (1985) provides additional information.

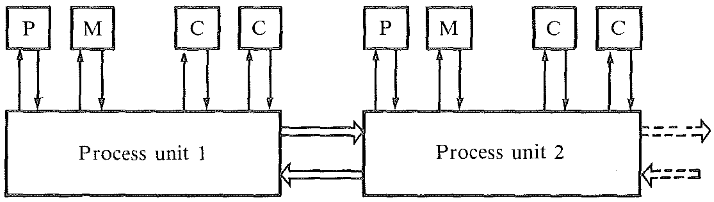
14.6.2 Manual control versus fully automatic control

Manual control is one major process operating philosophy. This is not, as is sometimes mistakenly supposed, inherent in process control systems implemented with the older technologies. Some of the very latest systems based on digital computer technology with sophisticated colour video displays are designed to support essentially a manual control philosophy.

The basic idea in this approach is to provide the process operator with a good dynamic display or other means of visualizing the state of the process, but then to leave the manipulation of many of the most important process control inputs to the operator. The process control system would also be designed to provide him with convenient means of making the necessary adjustments. In other words, the human operator is retained as an essential element of most feedback and feedforward loops. A recently proposed variation on this idea is that the process monitoring system should calculate and display some meaningful measure of economic performance for the entire process. This, in theory, then enables the human operator to use his ingenuity to find the most profitable operating condition for the plant.

Perhaps the major advantage of manual control is that the design of the process control system is relatively simple. It can also be argued that there are certain tasks where it will always be virtually impossible to better manual performance. On the other hand, a major disadvantage is that the use of manual control where a simple automatic device could do the job is a gross misuse of the human resource. Human performance is especially bad when simple tasks requiring concentration over extended time periods have to be performed. Human concentration is a limited resource and should only be called on in short bursts.

At the opposite end of the spectrum one finds the fully automatic process control systems. Here the philosophy is that whenever an adjustment based on a measurement needs to be made to a control input, this should be done automatically by the process control system without requiring human intervention. This particular philosophy can also extend upwards to the super-



Key: P Protection logic system
M Monitoring system
C Controller

Figure 14.16. Process control using hard-wired analogue equipment.

visory control of overall process unit operation, the coordination of the operation of the various units and overall plant economic optimization. Implementation of fully automatic process control has only really become feasible due to the flexibility of the newer programmable computer-based process control technologies, although it is certainly possible to implement a large number of the individual tasks required using hard-wired analogue process control technologies.

Most present-day mineral processing plant control systems fall somewhere between the above two extremes, with a steady trend towards the use of fully automatic control.

14.6.3 Hard-wired analogue control systems

Process control equipment that was available commercially during the period from the early 1960's to the mid 1970's utilized analogue electronic or occasionally pneumatic technology. The form of control system architecture that can be readily implemented using this kind of equipment is shown in Figure 14.16. Signals from field transmitters are individually wired to a central control room where various items of equipment such as measurement read-out displays, recorders, controllers, alarm monitoring systems and safety shut-down logic systems are located. These are often mounted on or behind a large control panel to give a degree of centralization of information display in order to aid the operator in visualizing the state of the process.

An important property of such systems is that although the various functions are physically centralized in the control room, there is a logical separation or distribution of the various process control tasks among individual standard modules such as process controllers and shut-down logic modules. This gives these systems their major advantage of simplicity and a degree of fault tolerance. If one controller fails, for example, it does not necessarily affect the operation of the rest of the process control system seriously.

The disadvantages of hard-wired analogue systems are, firstly, that it is difficult to achieve a properly coordinated and centralized display of the state of the process. The various measurement read-outs are usually spread over a fairly large section of the control panel and the operator has to move

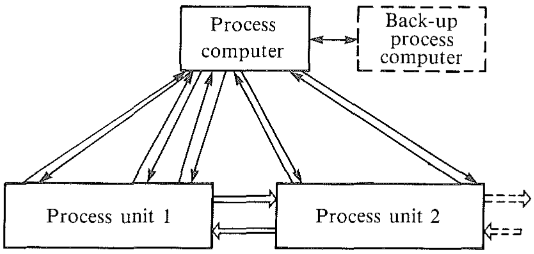


Figure 14.17. Process control using centralized process control computer.

from one read-out to another in order to build up a mental picture of the operation of the plant. Secondly, with these systems, flexibility to implement advanced control functions for process study, development and management information tools and to modify and improve existing functions as the needs of the process evolve is severely limited. A third disadvantage is that the cost of such equipment is relatively high.

In spite of these disadvantages, hard-wired analogue control systems have a good reputation for reliability and remain the most appropriate choice for many present-day applications.

14.6.4 Centralized computer control systems

The development of relatively cheap and powerful minicomputers during the early 1970s meant that by about 1975 it had become feasible and attractive to incorporate such minicomputers into process control systems. The benefits of doing this were that the vast data manipulation and storage capabilities of the digital computer could be used to collate and present data and derived information to the process operator in a far more meaningful and powerful way than had previously been possible. Historical process operating data could also be stored, retrieved on demand and processed into management reports and information concerning process dynamic behaviour. In addition to this, many of the previously hard-wired control tasks could be programmed, thereby giving far greater flexibility and the opportunity to implement more integrated and advanced forms of control. Details of how the controllers discussed in Section 14.3 can be implemented in digital computer form are available in the texts by Astrom and Wittenmark (1984) and Bennett and Linkens (1984).

The way that these early computer process control systems are arranged is illustrated in Figure 14.17. Because of limitations of the early computing hardware and cost considerations, individual signals from all the field transmitters are wired into one central computer. This computer carries out all of the process control tasks including information display, continuous control, alarm monitoring and safety shut-down logic functions. These systems certainly allow the implementation of more integrated and coordinated information display and control strategies, but suffer from the major problems of very high complexity and poor reliability, particularly

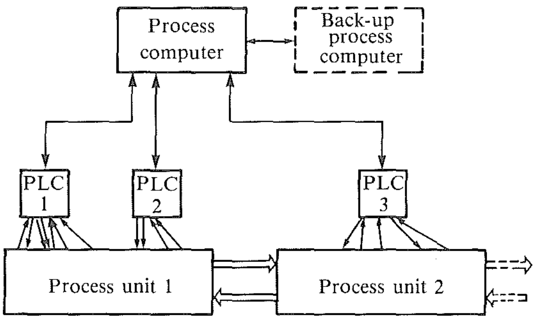


Figure 14.18. Process control using centralized process control computer with distributed PLCs.

in connection with the programs or software. There has consequently been a trend in more recent computer control systems to return to a greater distribution of the process control tasks among simpler separate hardware modules. In modern systems these separate modules themselves would typically be computer-based. Figure 14.18 shows the structure of an example of this kind of system which uses a number of programmable logic controllers (PLCs) connected to a minicomputer in a star arrangement. The PLC modules would perform scanning and pre-processing of the sensor inputs and also possibly logic control and PID control tasks. The minicomputer would typically provide information display, historical data logging, report generation and process study tools.

14.6.5 Distributed computer control systems

The problems experienced with earlier centralized computer control systems based on a single minicomputer and the increasing availability of cheap digital computing power in the form of the microprocessor have spurred on the development of a new generation of digital computer-based process control systems since the late 1970's. Known as distributed computer control systems, these attempt to combine the advantages of hard-wired analogue process control systems and centralized computer control systems without incurring the disadvantages of either. The basic idea is that a distributed computer control system consists of a collection of fairly autonomous digital processing modules which communicate with one another over a shared high-speed digital communication channel or data highway.

There is no master module that directs overall system operation in a truly distributed system. System co-ordination relies instead on co-operation among the autonomous processing modules. This eliminates the possibility of total system failure due to the failure of any one processing module.

The various tasks of process control are distributed among these separate modules which may also be physically distributed along the data highway as shown in Figure 14.19. This retains the advantages of simplicity and fault tolerance through the use of simple standard modules and also gives a potential reduction in wiring costs. On the other hand, the data collation and pro-

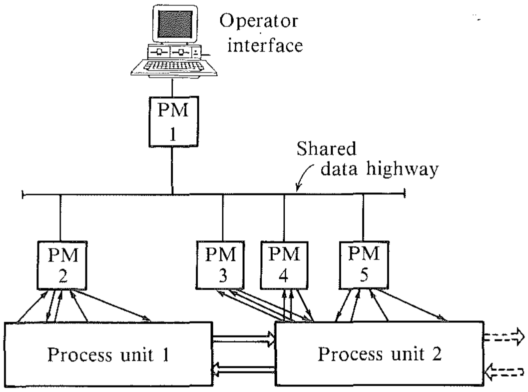


Figure 14.19. Distributed computer control system using five autonomous processing modules. Note that processing modules 3 and 4 operate in active redundancy to give fault tolerance.

cessing capabilities of the digital computer are retained, allowing powerful centralized information display and operator interface capabilities and the implementation of advanced and flexible forms of control.

14.6.6 Examples of current practice

This section discusses three examples of recently installed process control systems. Figure 14.20 shows the overall structure of the control system for

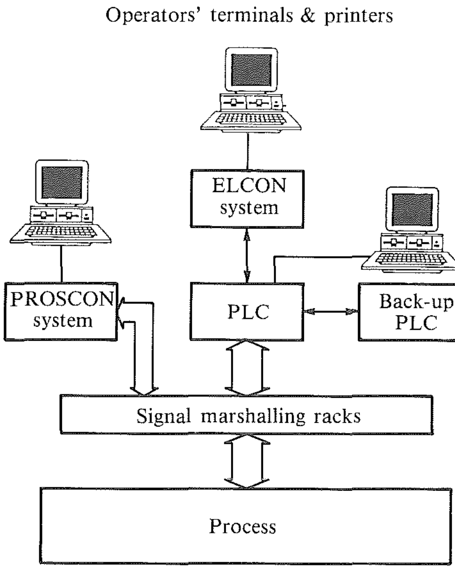


Figure 14.20. Structure of control system for East Driefontein gold plant.

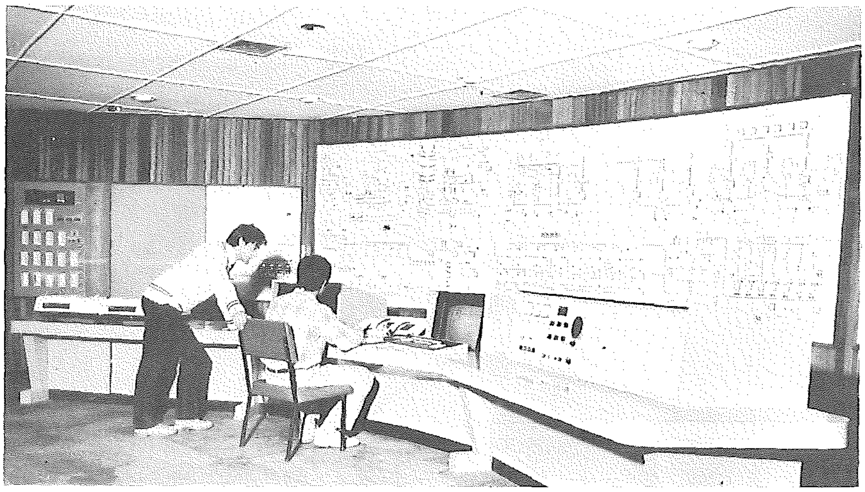


Figure 14.21. East Driefontein gold plant control room.

the East Driefontein gold plant. The system includes two process computers and two PLCs. Although similar to the basic centralized computer control concept discussed in Section 14.6.4, this system actually incorporates a greater degree of task distribution. The tasks performed by the various sub-systems are described in the following paragraphs.

The PLC monitors all measured plant variables and controls all major plant motors. It carries out alarm checking, logic control for interlocking of equipment, on/off control according to measured values and mill pebble feed control according to a special program. It also produces reports once per shift. A back-up PLC is available to take over in the event of failure of the primary PLC.

The ELCON supervisory system provides colour graphic displays of the process which can include dynamically up-dated indications of motor states, measured variables and the status of alarms. It also provides displays of control loop details, plots of measured variable trends and hourly, once per shift, daily, weekly, monthly and yearly reports.

The PROSCON system shown in Figure 14.20 is a process computer which is dedicated to the task of multi-variable control of the three milling circuits. This is an implementation of the multi-variable controller referred to in Section 14.3.3.

The control room of the East Driefontein plant is shown in Figure 14.21. The operator's colour video terminals can be seen in the foreground. This is typical of the kind of operator interface found in many modern systems. The old control panel, showing a mimic diagram of the plant, can be seen in the background.

Figure 14.22 shows the overall structure of the control system for the Cooke gold plant of Randfontein Estates. This is a PROSCON proprietary

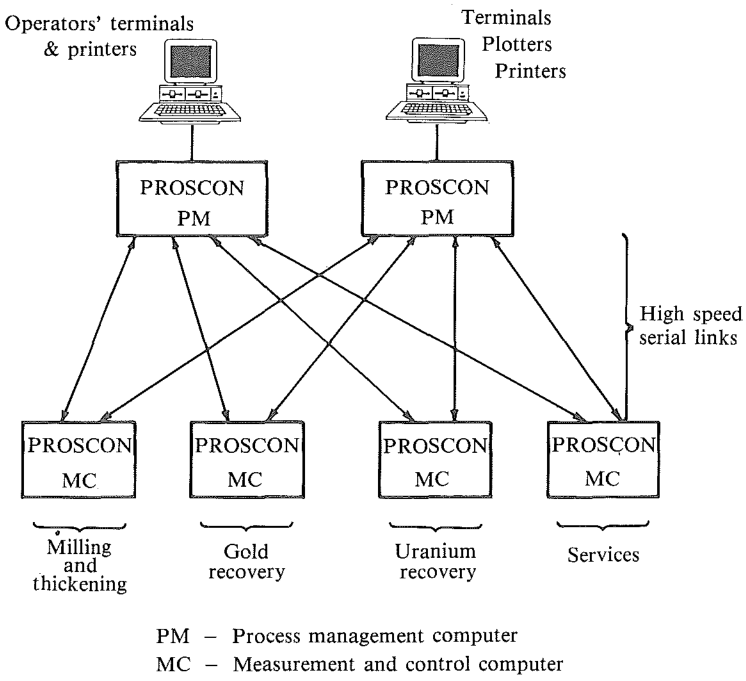


Figure 14.22. Structure of the control system for the Cooke gold plant.

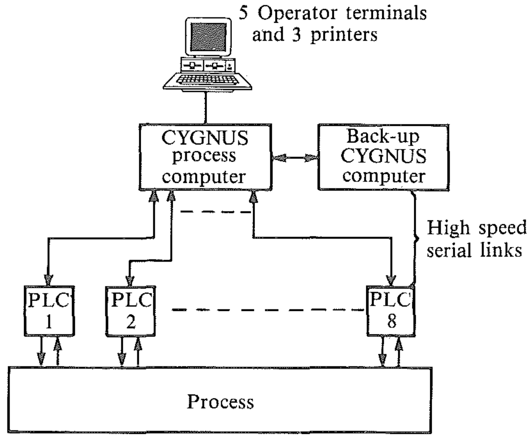


Figure 14.23. Structure of control system for Vaal Reefs No. 9 Shaft gold plant.

system and includes six process computers. Four of the computers interface directly with the process instrumentation and actuators and have relatively low processing power. These perform averaging and processing of measured variables, alarm checking, PID control and logic control. The remaining two computers are much more powerful and provide colour graphic displays for the plant operators, data collection and storage functions, reporting, event recording and a process study package. Note the regular structure of the system. This is typical of systems supplied by a single manufacturer. Again there is a moderate degree of task distribution in this system and redundancy to provide fault tolerance.

The process study package provides comprehensive facilities for interactively manipulating the stored historical data. There are data preparation facilities (plotting, selection and rejection, manual input, mathematical functions and filtering), statistical analysis facilities (histograms, autocorrelation, crosscorrelation and power spectra) and static and dynamic model identification facilities. In addition, there is a facility for designing multi-variable controllers.

The CYGNUS process control system installed at No. 9 Shaft gold plant of Vaal Reefs is shown in Figure 14.23. Note the close correspondence with the structure shown in Figure 14.18. The eight PLCs each interface with a section of the plant. The division into sections is as follows: ore handling, Mill 1, Mill 2, thickeners, leaching, CIP A, CIP B, and residue. The functions performed by the centralized supervisory computer are as for the PROSCON system discussed above. An interesting feature of the CYGNUS system is the special keyboard shown in Figure 14.24. This allows very easy access to comprehensive system functions by the plant operator.

An interesting aspect of the Vaal Reefs No. 9 Shaft system is that a KENWALT process simulator was used for testing of the process control system and for operator training during plant commissioning. A special-purpose computer which runs the simulator is connected to the instrumentation wiring in place of the real plant instruments to allow this. The KENWALT simulator can also be used during design and provides the following:

- A facility for the plant designer to optimize flowsheets by allowing the quick evaluation of the effects of a large number of plant variables.
- A facility for sizing equipment and designing a maintenance schedule to minimize the effects of equipment down-time on plant performance.
- A facility for the control system designer to design and pre-tune the control loops using the dynamic models inherent in the simulated process flowsheet. Any problematic loops which may entail plant redesign will become apparent at this stage.
- A facility for designing and testing the stop/start sequence logic.

This illustrates the opportunities for optimizing the interaction between process design and control system design that are now available.

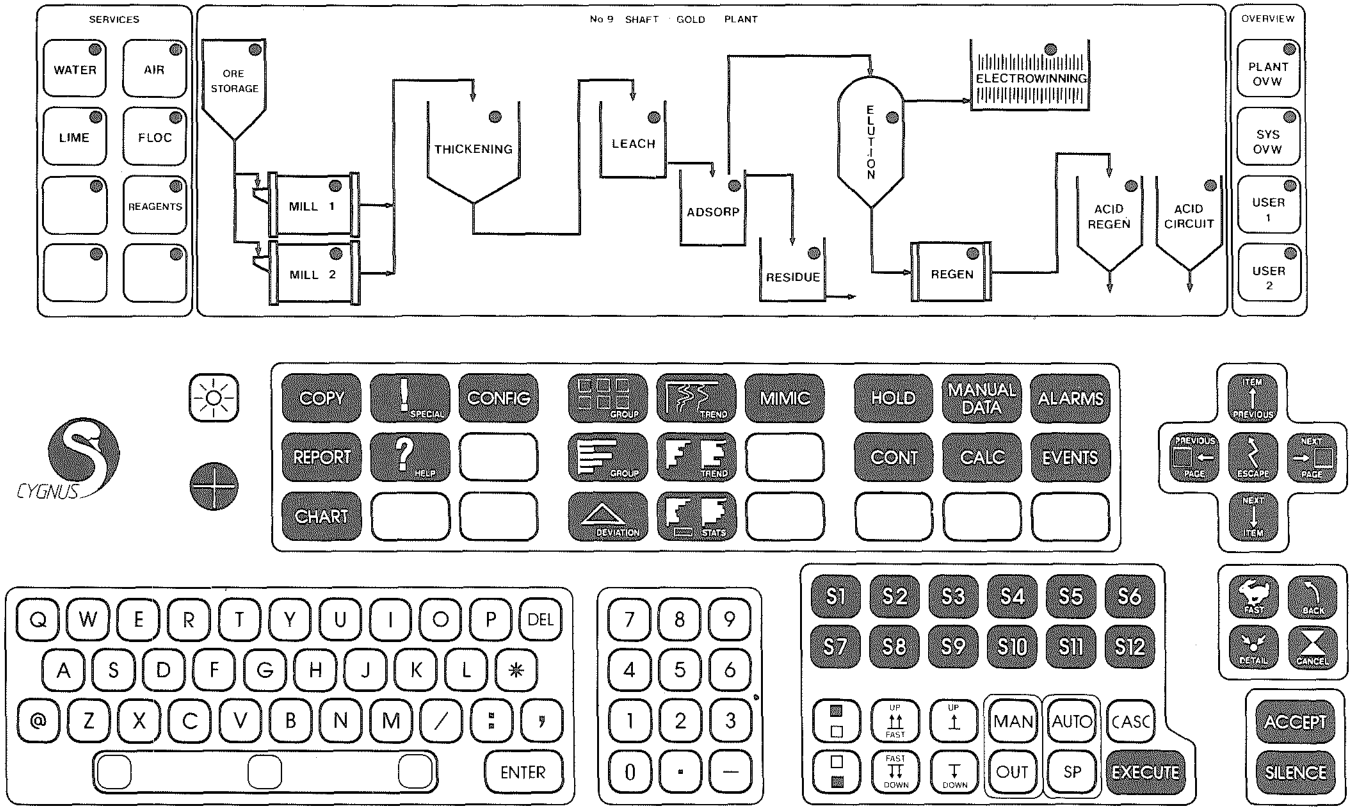


Figure 14.24. Example of special keyboard used in CYGNUS process control system.

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