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The use of sophisticated instrumentation and control strategies in mineral processing operations represents a relatively recent development. The variables that are normally targets for control and the possible variables that can be manipulated to achieve control in crushing, grinding and flotation are shown in Table 1. The important process variables encountered in mineral processing and corresponding instrument devices are shown in Table 2.

TABLE 1
Controlled and manipulated variables for crushing, grinding, and flotation

Controlled variables	Manipulated variables
Crushing	
Product fineness	Feedrate
Circulating load	Closed-side setting
Power draught	Screen area
Bin level	
Crusher level	
Grinding	
Product size	Sump-water rate
Sump level	Feedrate
Circulating load	Pumping rate
Hold-up of solids (%)	Solids in feed (%)
	Mill speed
Flotation	
Recovery (product tonnage)	Aeration
Grade	Agitator speed of rotation
Circulating loads	Pulp level
Froth levels	Reagents
Per cent solids	Frother
	Collector
	Modifier
	Depressants

TABLE 2
Mineral processing instrumentation

Measurement	Devices	Type of Circuit*
Sump level Pulp level	Bubble tube/differential-pressure cell Capacitance probe Sonic sensor	G,F
Pulp density	Gamma nuclear gauge U tube/load cell Differential-pressure cell	G,F
Volume flowrate Water Slurries	Orifice plate Turbine meter Magnetic flowmeter Ultrasonic flowmeter	G G,F.
Particle size	Sonic particle-size monitor Light-scattering size analyser	C,G,F
Mill load	Load cells Watt meter, torque meter Sound meter	
Feedrate of dry solids	Weightometer	C,G
Crusher power Mill power	Watt meter, torque meter	C,G
Particle composition	X-ray-fluorescence analyser Neutron-activation analyser	F
Bin level	Sonic sensor Capacitance probe	C
Froth level	Capacitance probe	F

* C = Crushing, G = Grinding, F = Flotation

In classical control as is current practice the aim is to maintain the control variables constant by adjusting the manipulated variables. These are essentially stabilising control loops using conventional feed back control employing PID controllers. Possible linkings of manipulated and controlled variables are shown in Table 3 The process matrix of Table 3 shows the response of the controlled variable (plus means increase) and speed of response to increase in manipulated variables. Table 4. shows the broad classes of control strategies in crushing, grinding and flotation. Typically improvements of about 10% in crushing and grinding and 5% in flotation are possible by using such classical control methods.

TABLE 3

Response of controlled variables to changes in manipulated variables as indicated by process matrices

Manipulated variable	Controlled variable				
Crushing	y_1 = product fineness	y_2 = circulating load	y_3 = power draw	y_4 = bin level	y_5 = crusher level
u_1 = feedrate	− + Fast	+ 0 Slow	− Slow	+ Slow	Fast
u_2 = closed-side setting	− + Fast	− Slow	− Fast	− Slow	Fast
u_3 = screen area	+ − Slow	− Slow	− Slow	− Slow	Slow
Grinding	y_1 = product fineness	y_2 = circulating load	y_3 = sump level	y_4 = % solids in mill	
u_1 = sump water addition rate	+ Fast	+ Fast	+ 0 Fast	− + Slow	
u_2 = fresh feed solids rate	− Slow	+ Slow	+ Slow	+ Fast	
u_3 = cyclone feed pumping rate	+ Fast	+ Fast	− Fast	+ Fast	
u_4 = feed water addition rate	+ − Slow	+ Slow	+ Slow	− Fast	
u_5 = mill speed	+ − Fast	− + Fast	− Slow	− Slow	
Flotation	y_1 = grade	y_2 = recovery	y_3 = froth depth		
u_1 = aeration rate	− Fast	+ Fast	+ Fast		
u_2 = agitation rate	0 + Slow	+ − Slow	+ Slow		
u_3 = pulp level	− Slow	+ Fast	−		
u_4 = frother	− Fast	+ Fast	+ Fast		
u_5 = collector	− Slow	+ 0 Slow	+ Slow		
u_6 = depressant	+ Slow	− Slow	− Slow		

TABLE 4

Classes of control strategies involving combinations of principal control loops

Type	Controlled	Manipulated
Crushing		
I	Product size	Closed-side setting
	Power	Feedrate
II	Product size	Feedrate
	Power	Closed-side setting
Grinding		
I	Product size	Solids
	Circulating load	Feedrate
II	Product size	Sump-water rate
	Circulating load	Solids feedrate
Flotation		
I	Recovery	Aeration rate
	Grade	Pulp level
II	Recovery	Pulp level
	Grade	Aeration

Control of Crushers

Classical control

Control of crushers may be undertaken at the following levels :

i) Interlocking and alarms

ii) Feed back control

iii) Optimisation of throughput and particle size.

The first and third levels of control are shown in Fig.1

The controlled variables are product fineness, and throughput. The manipulated variables are feed rate, closed-side setting and screen area. Circulating load, bin level, power draw and crusher level may be considered secondary controlled variables. The major disturbances are changes in ore hardness, change in flow rate from surge bins, number of machines operating and changes due to liner or screen cloth wear.

The main process variables and their measurement is summarised in Table 5.

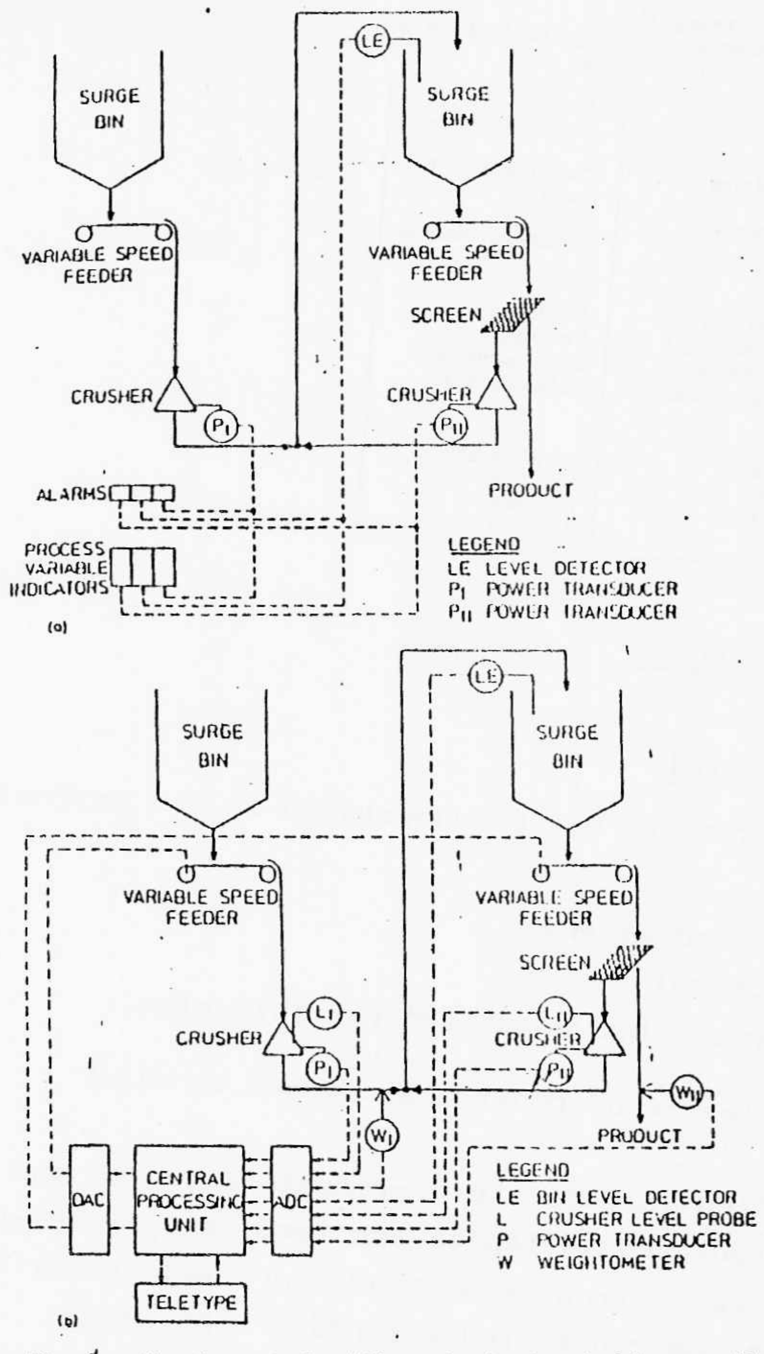

Fig. 1 Requirements for different levels of control for a crushing plant (a) local and interlocked on/off control (b) complete control using a digital computer.

The performance of a crusher strongly depends on the closed side setting, power draw and hardness and size of feed. Increase in feed rate increases the power draw and results in a finer product. At the same feed rate increased closed side setting results in a coarser product.

TABLE 5.

Instruments and techniques used in process control systems for crushing plants

Process measurement	Instruments	Comments
Mass flow of ore	electronic load-cell mounted on a weigh-length formed by suspending one or more conveyor idlers independently	output is normally by continous integrators which multiply instantaneous belt speed by belt loading to give instantaneous tonnage rates; for production control purposes, batch totalisers may also be used
Bin level	(1) ultra-sonic instruments	determine level by measuring travel time for sound reflected from the ore; problems can arise from multiple reflections from idling
	(2) gamma-ray instruments	have proved to be reliable in service; may be either on/off using single collimated beam or continuous over a limited range using wide angle source
	(3) weighing of surge bin by mounting on load cell	high initial cost but low maintenance and satisfactory service
Power draw	(1) ammeters measuring current (2) thermal convertors measuring power	care should be taken with the phase difference of current and voltage in crusher operation; generally, power measurement is more reliable when load charges are substantial; totalising kWh meters are also used for production control
Closed-side setting	(1) leading	inconvenient, time-consuming and cannot be done under load
	(2) calibration against mainshaft position	only available in crushers equipped with hydraulic mainshaft support systems

In crushers with constant closed side settings the power draw is maintained by feed rate control as shown in Fig.2. In Fig. 3 the bin level is measured and the power draw to the previous and succeding stages are controlled.

In systems having hydraulic main shaft support the CSS can be changed while in operation. In such a case the system is maintained in choke feed condition using a nuclear level gauge. The CSS is controlled with reference to power draw or hydraulic pressure. Here also cascade of power set point with surge bin level is possible.

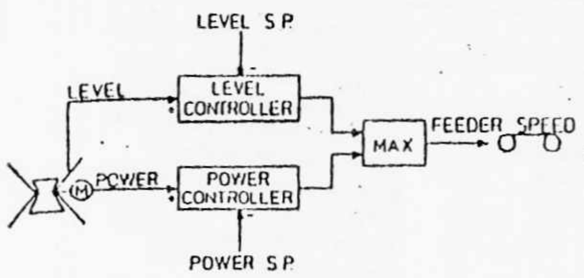

Fig. 2 Control loop for maintaining power draw or feed chamber level in a crusher without automatic setting adjustment.

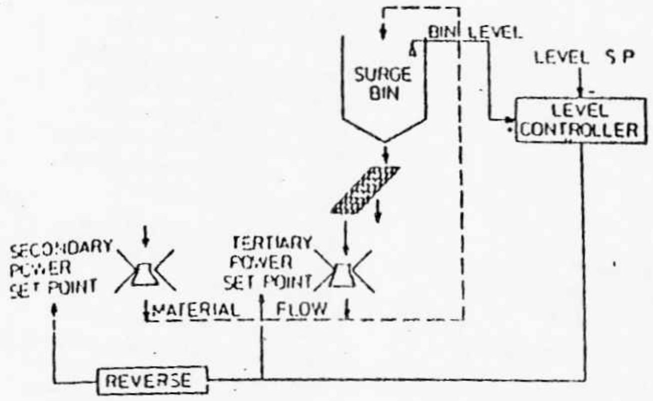

Fig. 3 Control loop for maintaining the level in the surge bin. The bin level also provides information on which stage is limiting production.

Computer control of crushers

The optimal control of an industrial crusher using a self-tuning control algorithm is presented below. A schematic diagram of the crushing plant is shown in Fig. 4 The fresh feed and crusher discharge are screened, and the oversize of each screen is fed to the crusher. The purpose of the control is to avoid stoppage of the crusher due to overloading. To maintain a high and constant power output, it is necessary to control the ore feed to compensate for variations in crushability, lump size, and crusher wear. The transportation lags in the system contribute to the difficulties of control. The lag between the feeder and the crusher is 40 to 50 s, and the lag time in the recycle loop is 70 to 80 s. The time constants of the feeder and crusher are about 12 and 20 s respectively.

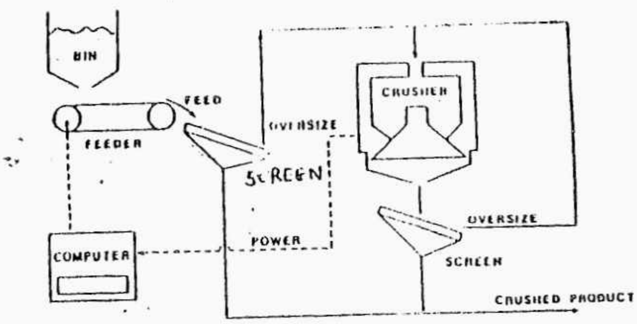

Fig. 4 A schematic diagram" of a crushing plant at Kiruna, Sweden

These investigators modelled the crusher by a regression equation of the form

$$y_{k+1} = -\alpha_1 y_k - \alpha_2 y_{k-1} - \alpha_3 y_{k-2} - \alpha_4 y_{k-3}$$

$$\beta_0 (u_k + \beta_1 u_{k-1} + \beta_2 u_{k-2} + \beta_3 u_{k-3}) + \varepsilon_{k+1}$$

where y_k represents the crusher power, u_k is the rate of fresh feed, and ε_k is the model error. All the parameters in the crusher model except β_0 are unknown. Since the observations y_k are noisy observations, the unknowns are estimated by a least squares recursive estimation scheme. The squared error function of the estimation is given by

$$J = \sum_{i=0}^{k} \lambda^{k-i} (\varepsilon_{k-i})^2 \qquad \lambda < 1$$

As the term $(\lambda)^{k-i}$ weights the model errors of recent observations more and old observations less, the algorithm adapts to variations in the process.

The computer control schemes for the crushing section of a large lead-zinc concentrator plant is shown in Figs.5,6.

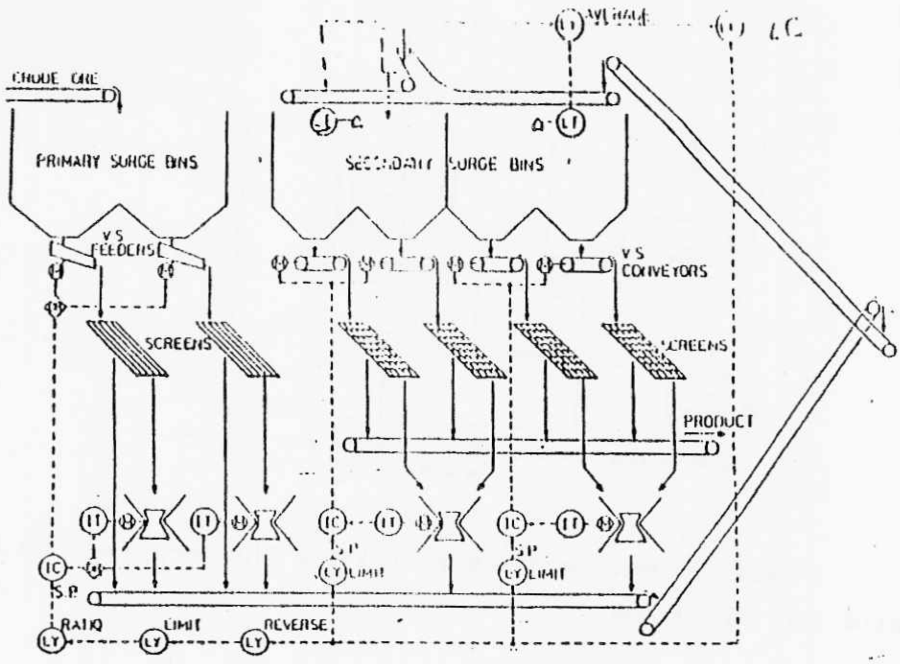

Fig. 5. Flowsheet and instrument diagram for the lead -zinc ore crushing plant,

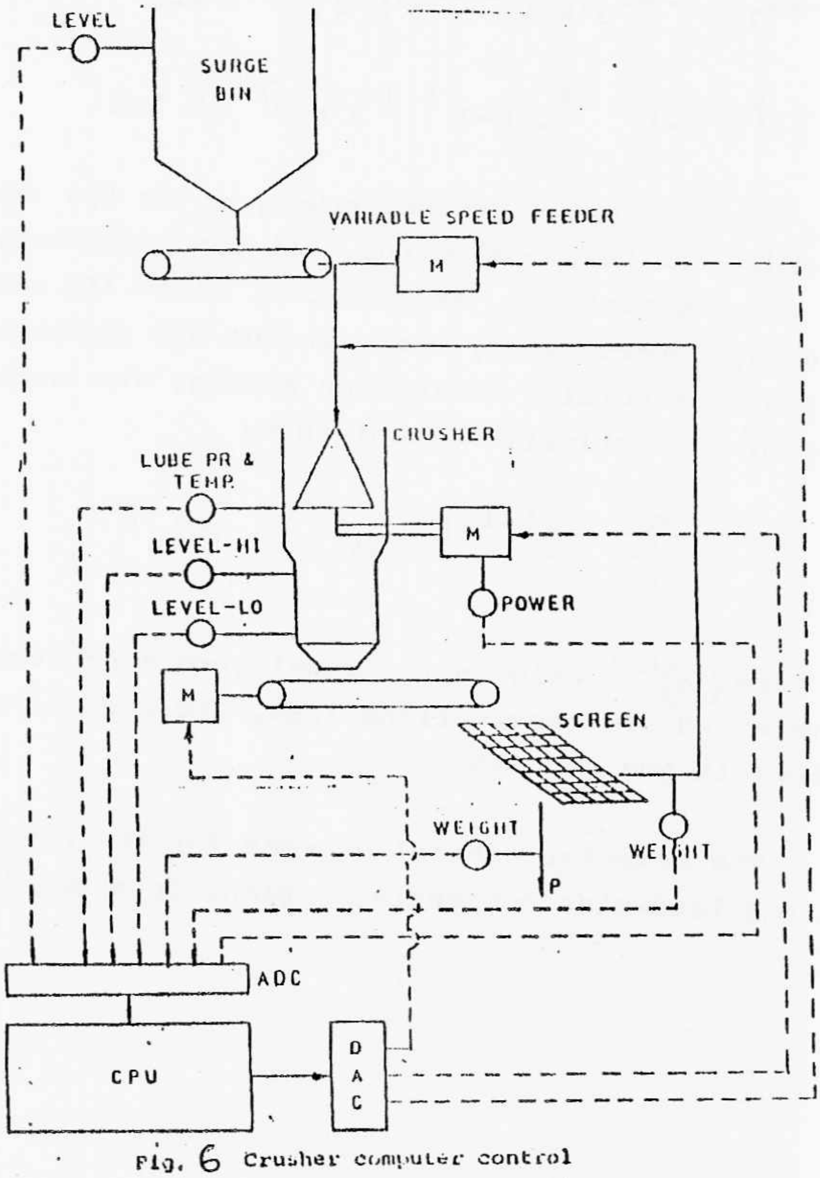

Fig. 6 Crusher computer control

Control of Tumbling Mills

Conventional control

The control and manipulated variable in grinding control are shown in Fig. 7 . The possible aims of control are maintaining product fineness constant while maintaining throughput const[illegible] or maximising it. The control systems may start with level 1 employing control of ore and water rates, level 2 controlling cyclone feeds to level 3 control on particle size and throughput

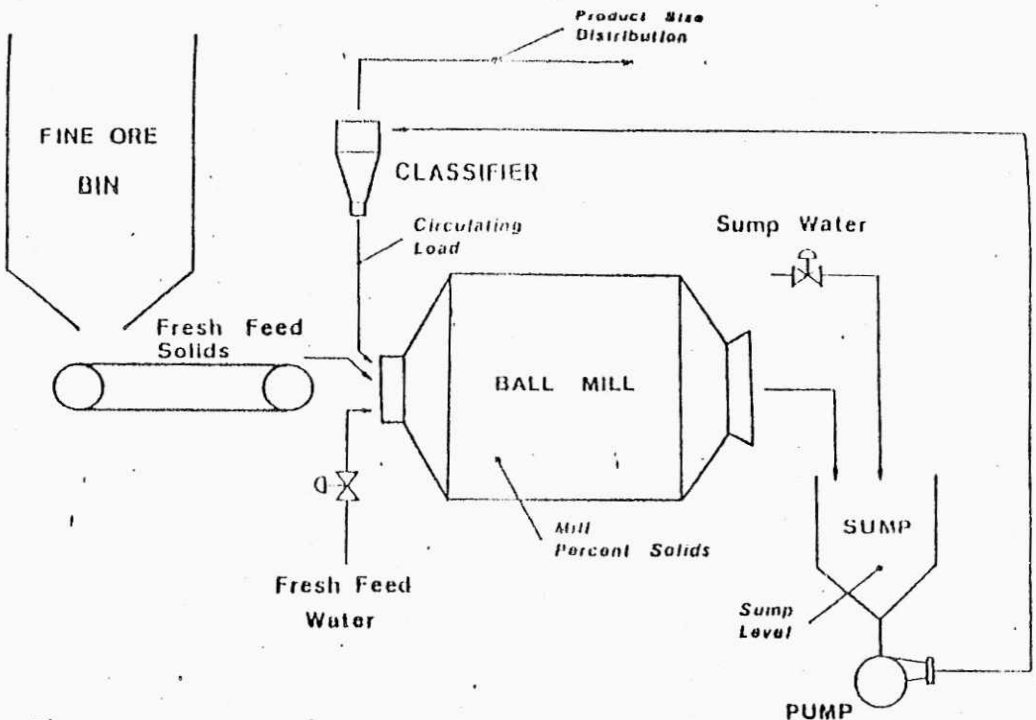

Figure 7. Schematic representation of a single stage grinding circuit showing controlled variables (Circulating load, Product size, Mill %solids, and Sump level) and manipulated variables (Fresh feedrate, Fresh water rate, Sump water rate and Pumping rate).

Two basic types of control one with constant cyclone feed rate and the second with variable cyclone feed are shown in Figs. 8 . The sensors used are listed in Table

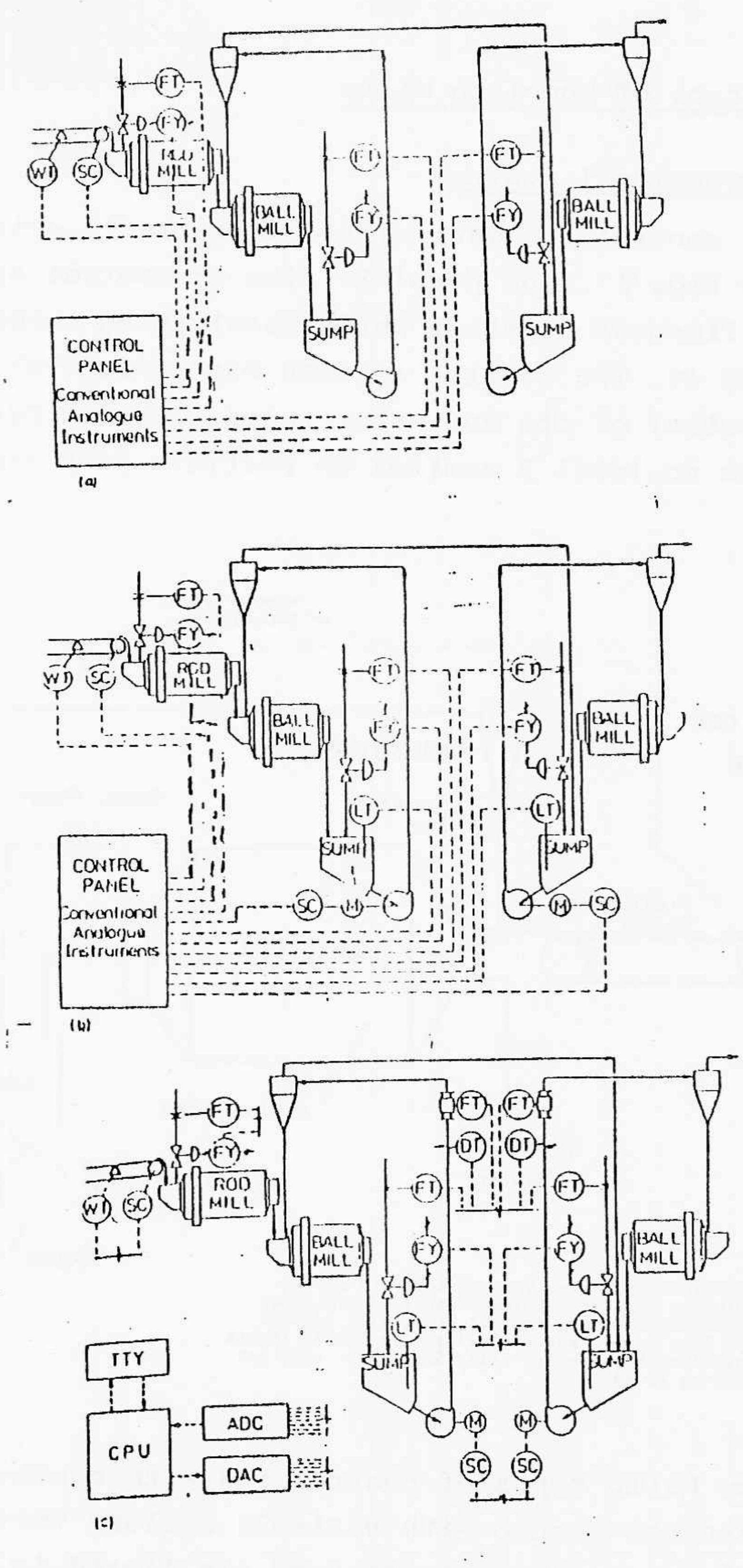

Fig. 8 Instrument requirements for different levels of control for a wet grinding circuit. a. Ore and water control to set points. b. Control of cyclone feed pumps in addition to (a.) c. Total circuit control using a digital computer.

TABLE 6

Sensing instruments which have been used in wet-grinding-circuit control systems

Instrument	Purpose
Weighers: mechanical, electrical or nuclear	mass flow rate of dry ore on a conveyor belt
Orifice, venturi meters	volume flow rate of water in a pipe
Magnetic flow meter	volume flow rate of any conducting fluid in a pipe
Gamma nuclear gauge; differential pressure cell; "Halliburton" tube	solids content or specific gravity of a slurry
"Autometrics" particle-size monitor; "RSM-Mintech" particle-size analyser	particle-size monitor
Power meter: autogenous mills	inference of load retained in the mill
Power meter: rake classifiers	inference of mass flow rate of coarse particles
Noise meter: ball mills	inference of load retained in mill
Vacuum gauge: air core of hydrocyclone	inference of solids content of cyclone underflow
Bubble tube	level of fluid in a tank such as pulp in a pump sump

The control schemes in several industrial processes are shown in Figs. 8 & 9

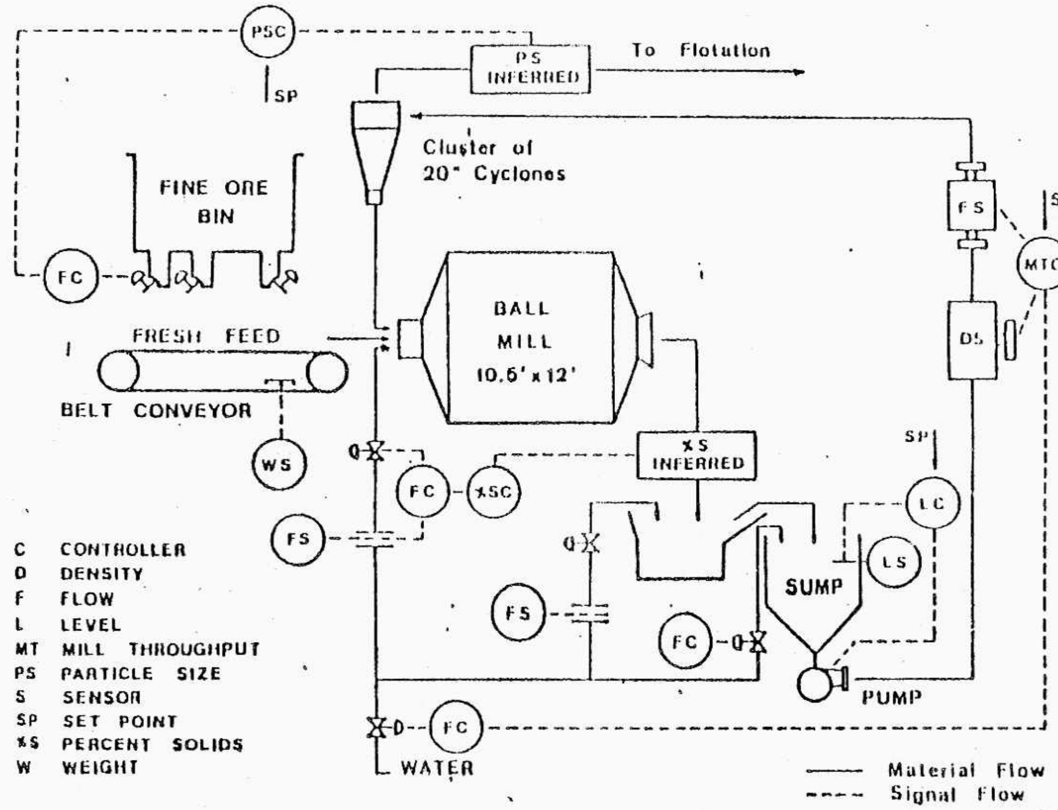

Figure 9 Instrumentation and controls at Asarco's Silver Bell unit.

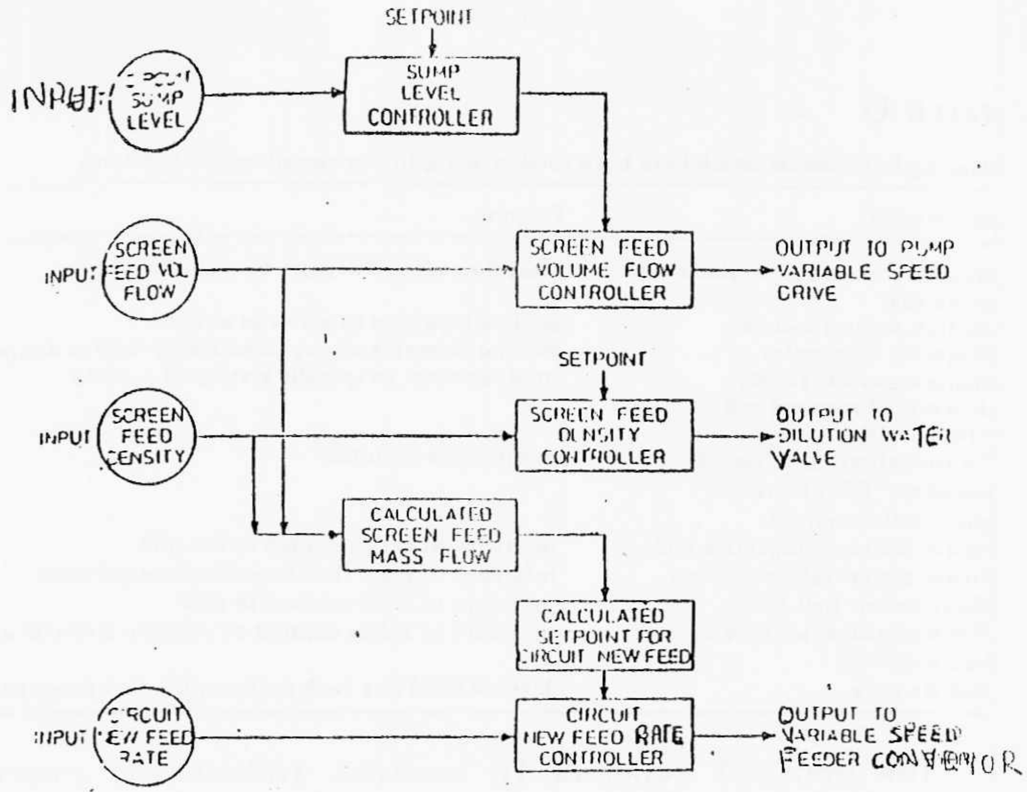

Fig. 10 Control logic for Renison Limited primary grinding circuit.

Computer control of tumbling mills

In recent years considerable interest has been developing in computer control of grinding systems. Other than conventional control implemented using computers recent work has shown new approaches to control. Decoupling control, optimising control and set poing supervisory control are some of these approaches. In the optimising or supervisory systems the plant throughput is maximised at a constant product size. Fig. 10,11,12 illustrate these systems.

Control of Flotation Circuits

There is considerable economic incentives for control of flotation circuits. Typically 6 months pay back period has been reported (Table . 7). Flotation control may consist of feed forward ratio control of reagents, circuit stabilisation control (levels and circulating loads), feed back control of

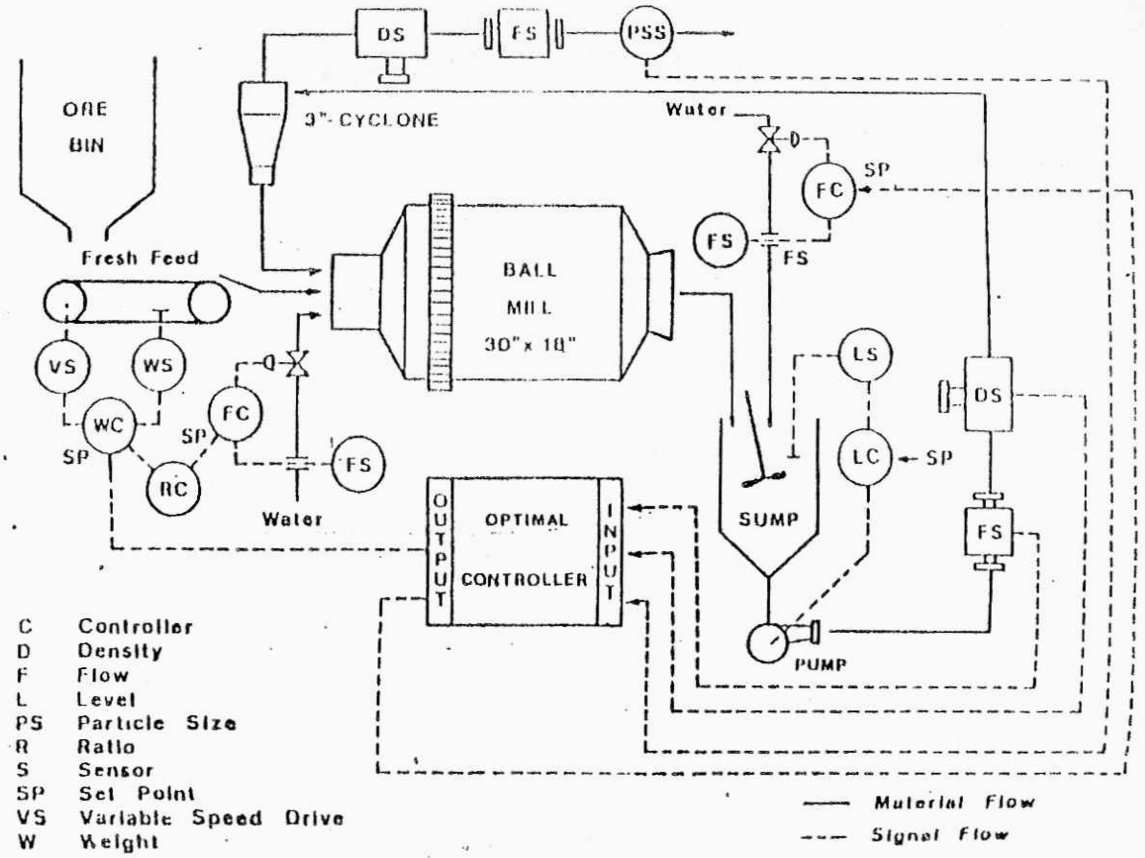

Figure 11 Instrumentation and controls at University of Utah Pilot Plant

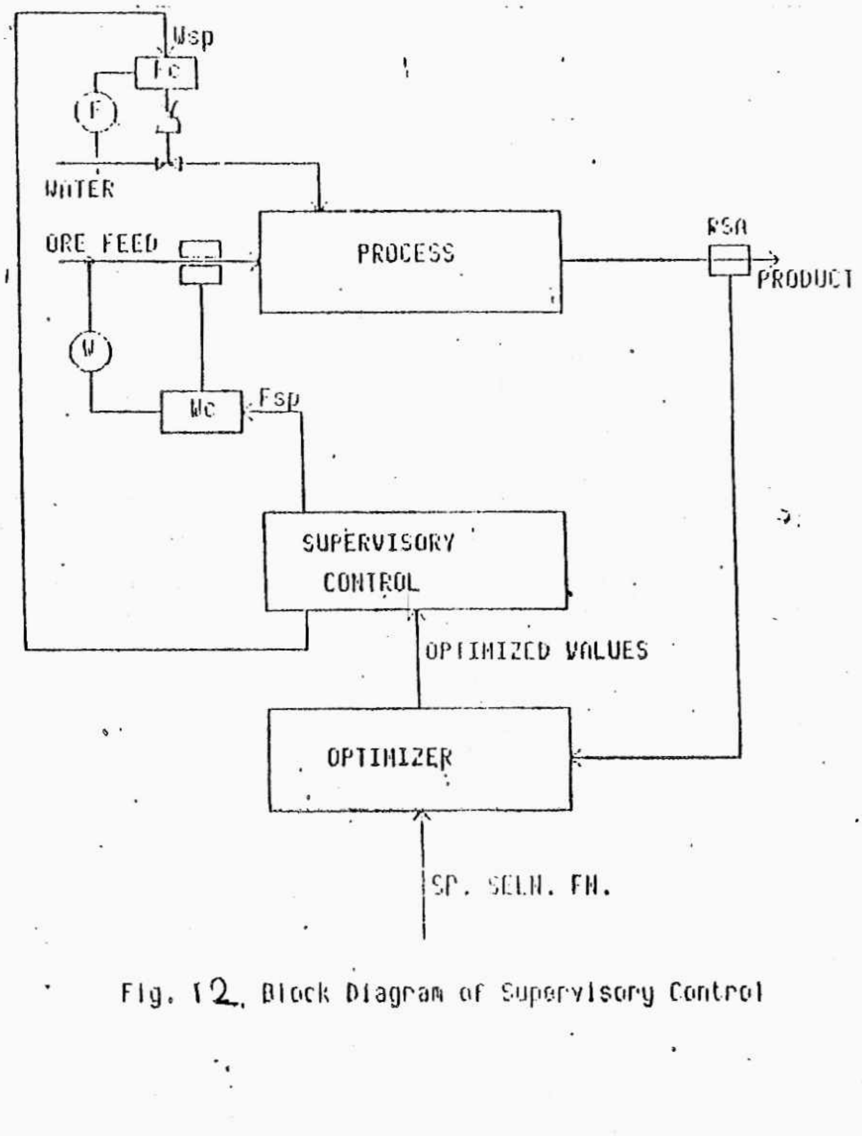

Fig. 12. Block Diagram of Supervisory Control

grades and finally optimising control of the system for maximum economic recovery.

Table 7 - Outokumpu Concentrators using Computer Control

	Capacity	Concentrates	Mode of Controls	Results
Pyhasalmi	2500 TPD	Cu,Zn,Fe	Stabilizing control 1969 and optimizing control 1972	2.5% increased metallurgical values from optimized control ($[illegible] annual gain)
Vuonos	1500 TPD copper circuit; 4500 TPD nickel circuit	Cu,Ni	Direct digital control in 1972 for operation of crushing and materials handling; stabilization control in 1974	1.5% increased Cu recovery
Kotalahti	2200 TPD	Ni,Cu	Proscon 103 stabilizaing control in 1974	2% increased nickel and copper recovery
Vihanti	2500 TPD	Zn,Pb,Cu	Proscon 103 stabilizing control in 1974	2% increased zinc recovery
Keretti	1500 TPD new feed; 3000 TPD reclaimed tails feed	Cu,Zn,Co	Proscon 103 stabilizing control in 1975	about $200,000 annual gain in reagents savings improved recovery

The important load disturbances are feed slurry density, feed assay, feed size distribution, feed mineralogy and liberation.

In feed forward control (Fig.13) the reagent addition is controlled with reference to significant variation in the feed characteristics. Fig. 13 shows a typical feed back control system.

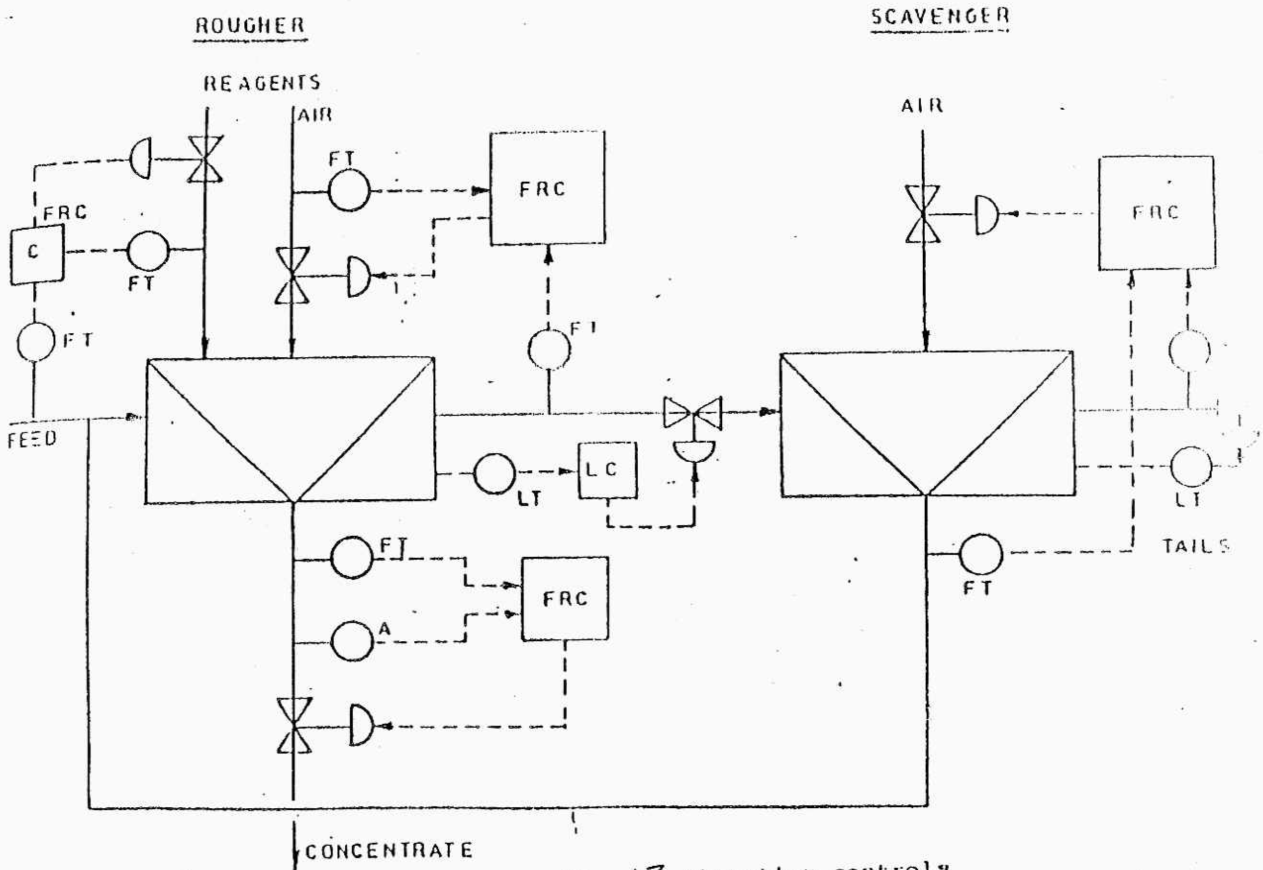

Fig. 13 Flotation controls

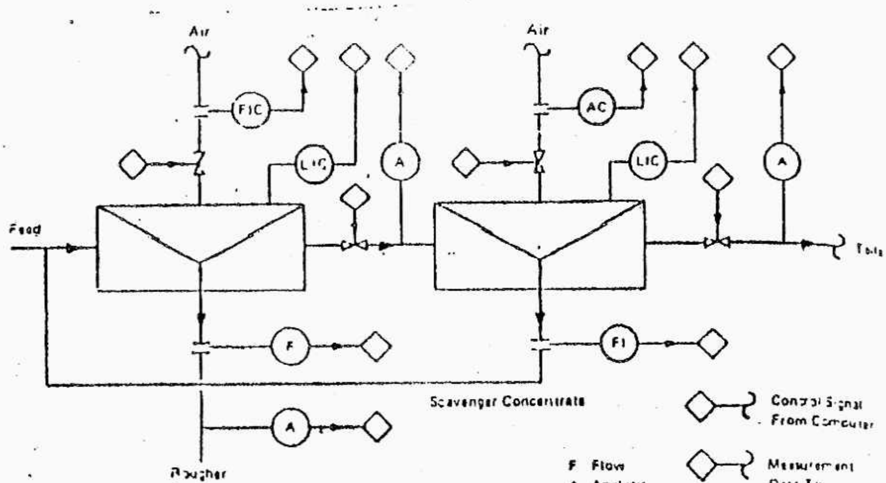

Figure 14 Flotation circuit stabilizing control scheme.

In the system of Fig. 13, the variables measured are,

*feed assay

*tails assay

*concentrate assay

*aeration rate

*rougher concentrate flow rate

*rougher cell level

*scavenger concentrate flow rate

*scavenger cell level.

Superimposed on the control system is feedfowrard reagent flow control made in ratio to ore tonnage rate. Control points in Fig. 13 are :

*flow rate of rougher concentrate, controlled by aeration rate in the rougher cell bank

*assay of rougher concentrate, controlled by flow rate of rougher concentrate

*level in rougher cell bank controlled by dart valve position

*flow rate of scavenger concentrate, controlled by aeration rate in scavenger cell bank

*assay of tails, controlled by flow rate of scavenger concentrates

*level in scavenger cell bank controlled by dart valve position.

The control system described in Fig. 13 comprises two cascade control loops, one for each cell bank, in which the assay value provides a correction to the concentrate flow rate set-point from the aeration rate. Level variations are compensated by flow control of the tails from the cell bamk. A convenient diagram to display the interaction of measurements and control loops is given in Fig. 15

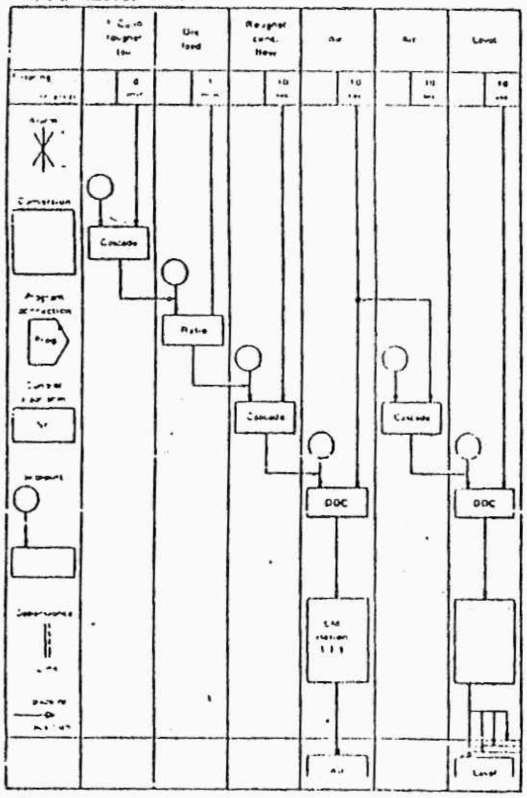

Figure 15 Process computer measurements and controls at Pyhasalmi concentrator, Cu rougher flotation.

Computer control of flotation cells

A typical computer control logic is described below. The ore is processed for Cu, Zn and pyrites. The control objective is to optimise Cu + Zn production.

The ore feed is processed for copper, zinc and pyrite concentrates. Copper flotation is subject to disturbances due to a variable quantity of oxidized (activated) zinc. Optimized control seeks an economic balance of copper and zinc recovery accounting for metal impurity in each concentrate. The optimizing systems gives the set-points for on-line direct digital control. Copper rougher concentrate flow is the control variable determined by set-point on copper losses in tails according to the control loop diagram (Fig. 16). The rougher concentrate flow set-points are maintained by aeration control with backup by pulp level control, linking cascade loops through direct control as described on the diagram.

The optimization scheme determines the set-points for copper in copper tails, rougher concentrate grade, and reagent feed rates to compensate for process disturbances. The influence of oxidized zinc is found through feed back of process information. The scope of the system incorporates statistically derived mathematical model having coefficients periodically updated to reflect changing influences of process disturbances such as the mineral oxidation.

The optimizing control scheme, schematically described in Fig. 16 is correlated to relative values of copper and zinc concentrate production, accounting for grade as determined by smelter feed economics. Fig. 16 indicates the comparative distribution of results under manual control and optimizing computer control. Before optimizing control of the metallurgy was put into effect, the flotation process was operated to favor recovery of copper and unnecessary losses of zinc were experienced. Optimized control increased total metallurgical values, or productivity, by 2.5% at only a slight decrease in copper recovery.

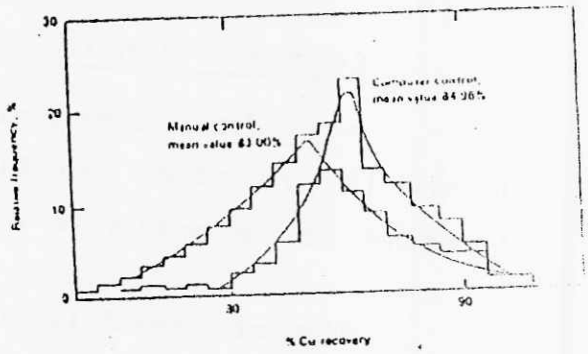

Figure Recovery economics

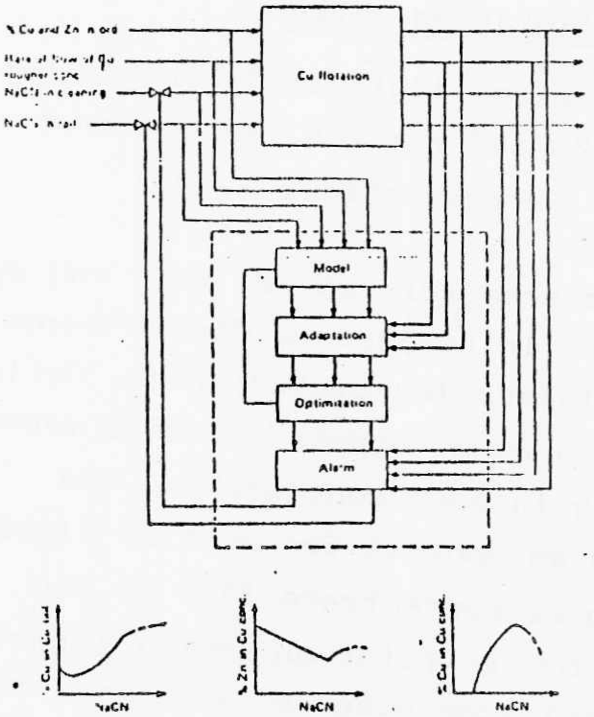

Figure 16 Principle of Cu flotation .. i

Typical control schemes for rotary filters and thickeners are shown in Fig. 17 and 18. These are simple feed back loops with safety interlocks and protective relays.

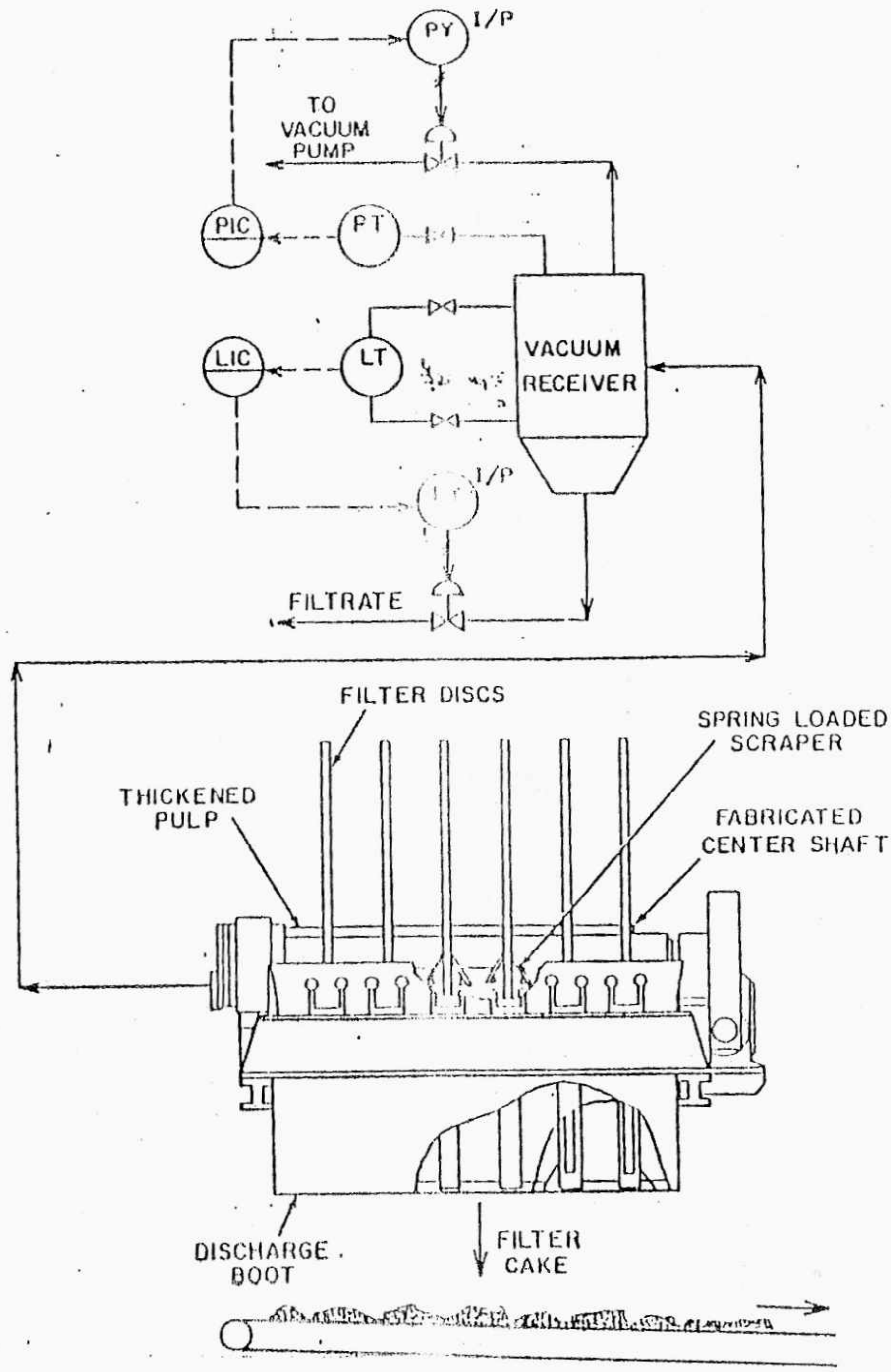

Fig. 17 Filtration controls

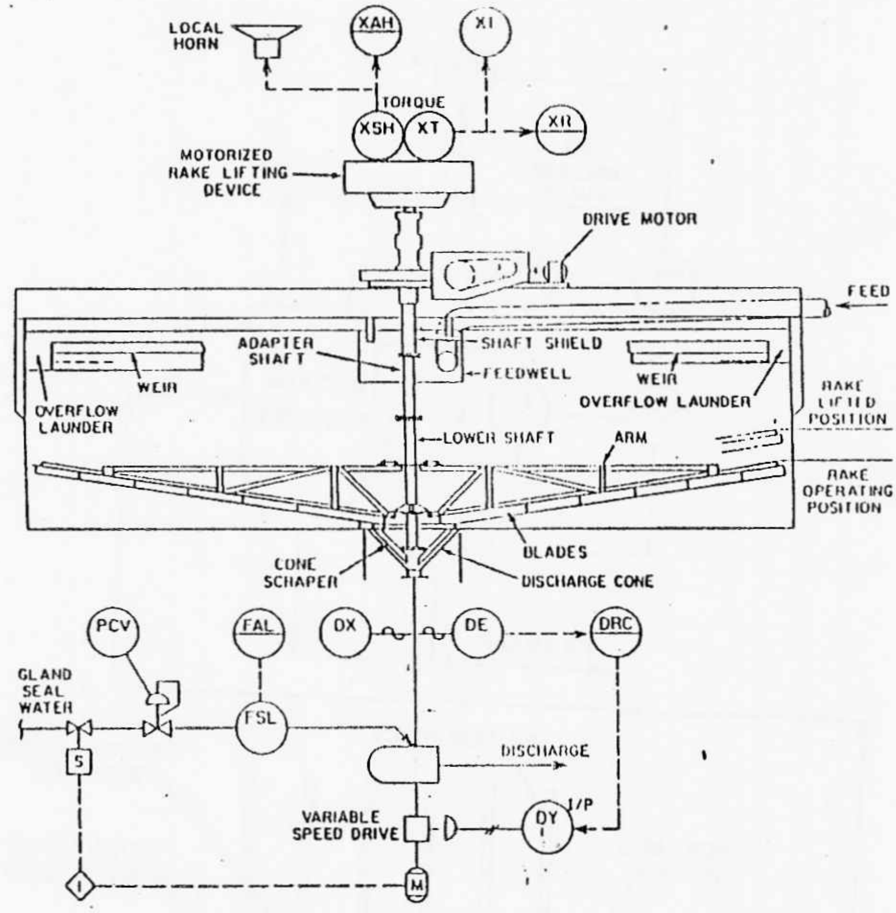

Fig. 48. Thickener controls