



TOGETHER
for a sustainable future

OCCASION

This publication has been made available to the public on the occasion of the 50th anniversary of the United Nations Industrial Development Organisation.



TOGETHER
for a sustainable future

DISCLAIMER

This document has been produced without formal United Nations editing. The designations employed and the presentation of the material in this document do not imply the expression of any opinion whatsoever on the part of the Secretariat of the United Nations Industrial Development Organization (UNIDO) concerning the legal status of any country, territory, city or area or of its authorities, or concerning the delimitation of its frontiers or boundaries, or its economic system or degree of development. Designations such as “developed”, “industrialized” and “developing” are intended for statistical convenience and do not necessarily express a judgment about the stage reached by a particular country or area in the development process. Mention of firm names or commercial products does not constitute an endorsement by UNIDO.

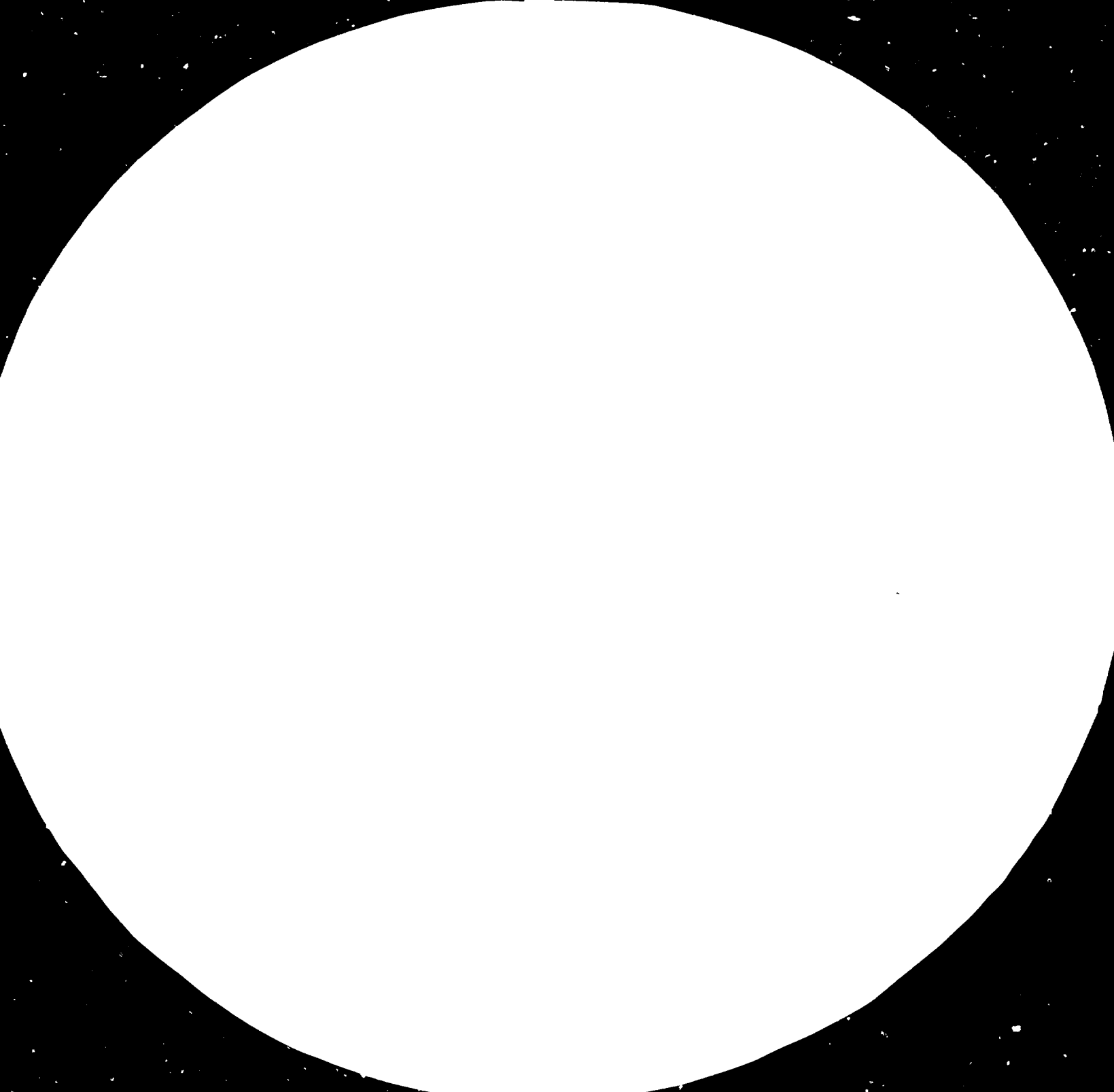
FAIR USE POLICY

Any part of this publication may be quoted and referenced for educational and research purposes without additional permission from UNIDO. However, those who make use of quoting and referencing this publication are requested to follow the Fair Use Policy of giving due credit to UNIDO.

CONTACT

Please contact publications@unido.org for further information concerning UNIDO publications.

For more information about UNIDO, please visit us at www.unido.org





MICROCOPY RESOLUTION TEST CHART

NATIONAL BUREAU OF STANDARDS
STANDARD REFERENCE MATERIAL 1010a
(ANSI and ISO TEST CHART No. 2)

→ Marie Dietrich



14147

United Nations Industrial Development Organization

PROCESS INSTRUMENTATION AND CONTROL
IN ALUMINA PLANTS.

G. Lang

2962

ALUTERV-FKI

BUDAPEST, 1984

UNITED NATIONS INDUSTRIAL DEVELOPMENT ORGANIZATION

PROCESS INSTRUMENTATION AND CONTROL
IN ALUMINA PLANTS

by

G.Lang
Hungarian Aluminium Corporation

The opinions expressed in this volume are those of
the author, and do not necessarily reflect the views
of the secretariat of UNIDO

Printed by
ALUTERV - FKI
Budapest, 1984

CONTENTS

	Pages
1. Development of instrumentation and process control	1
2. Special instruments and final control elements	5
2.1 Feed rate measurements of solids	5
2.2 Flow rate measurements of liquids	5
2.2.1 Head flow meters in open channels	5
2.2.2 Head flow meters in closed lines	8
2.2.3 The magnetic flow meter	12
2.3 Level measurements in tanks	13
2.3.1 Level measurement in open tanks	13
2.3.2 Level measurement in closed vessels	14
2.4 Density measurements of fluids	16
2.4.1 Density metering by bubbler assembly	16
2.4.2 Density measurement by gamma-ray absorption	18
2.5 Measurement of boiling-point rise	19
2.6 Oscillometric conductivity measurement	20
2.7 Control valves and actuators	22
2.8 Sizing of control valves	25
2.9 Variable speed pump drives	30
3. Basic instrumentation of unit operations	32
3.1 Instrumentation and control of wet grinding	32
3.2 Instrumentation and control of digestion	33
3.3 Instrumentation and control of clarification	34
3.4 Instrumentation and control of evaporation	35
3.5 Process instrumentation of calciner kilns	37
4. Advanced control concepts	39
4.1 Molar ratio control in digestion	41

	Pages
4.2 Concentration control of aluminate liquor	46
4.3 Seed ratio control for precipitation	52
4.4 Control of hydrate wash filtration	56
4.5 Concentration control of strong liquor	58
4.6 Concentration control of digestion liquor	62

FIGURES

	Pages
Fig.1 The approximate linear weir assembly	7
Fig.2 Concentric orifice plate with flange taps	9
Fig.3 Eccentric orifice plate with pipe taps	10
Fig.4 Air purged level and density meter	17
Fig.5 Oil seal pot applied for level measurement	15
Fig.6 Measurement of boiling point elevation	20
Fig.7 Oscillometric conductivity measurement	21
Fig.8 Pumped system with throttling valve	28
Fig.9 Pumped system with speed regulation	30
Fig.10 Instrumentation and control of wet grinding	32
Fig.11 Process instrumentation and control of digestion	33
Fig.12 Process instrumentation and control of clarification	34
Fig.13 Instrumentation and control of evaporation	36
Fig.14 Process instrumentation of a calciner kiln	37
Fig.15 Concentration controls in the Bayer-process	39
Fig.16 Molar ratio control in digestion	41
Fig.17 Concentration control of aluminate liquor	46
Fig.18 Seed ratio control for precipitation	52
Fig.19 Control of hydrate wash filtration	56
Fig.20 Concentration control of strong liquor	58
Fig.21 Concentration control of digestion liquor	62

1. DEVELOPMENT OF INSTRUMENTATION AND PROCESS CONTROL

Modest local instrumentation and manual operation was the state of the art in alumina plants, as long as the digestion process was batch operated.

With the introduction of continuous digestion the need of automatic control arose. In fact, reliable control techniques were a precondition for making the process continuous at all. Temperatures, pressures, flow rates, liquid levels had to be constantly adjusted by many valves and other final control elements, that was unthinkable to be operated manually, even with one man at each control valve.

Process fluids in alumina plants (slurries, caustic liquor) tend to scaling, clogging when stagnant, but abrading at higher flow velocities. Special designs had to be found for elements in direct contact to these fluids, to ensure reliable operation. Primary elements and control valves successfully used in alumina plants are to be dealt with to some detail later.

The equipment used for a long time were the various pneumatic instruments and control valve actuators, which earlier were locally mounted. Later instruments for supervision were grouped into operators panels, located in special control rooms.

The method used exclusively in the first period and still applied in the overwhelming majority is the negative feed-back loop to control a single variable.

The controlled variable may be any physical or chemical quantity measured. The manipulated variable is mostly a flow changed by the action of a control valve. In negative feed-back the action taken is manipulated negatively in its effect to the direction of the controlled variables deviation from the desired set point. When the temperature for instance in an autoclave rises over the set point, the heating steam flow will be throttled to make the temperature sink.

By application and extension of this technique many basic measurements and automatic controls came into use for every major unit operation as is to be shown later. A traditional instrumentation of an alumina plant comprises about 150 to 250 individual measurements and controls. A semi-centralized supervising system is prevalent with some 4 to 6 plant section instrumentation rooms. Summarizing counters to integrate flow- and weight rates, analog strip chart recorders for other important variables are widely used. Many plants are still employing this kind of process control.

Essentially the purpose of instruments and controls is not only to run the process smooth and safe, but also to turn out the best economy. That means to manufacture alumina of better quality and at a lower cost. The cost of raw materials and energy can be lowered and the quality of alumina improved by keeping the ideal process conditions within narrower tolerances.

Deviations of the controlled variable can only be kept within a narrow margin if the physical and the chemical quantities involved can be measured with high accuracy and if advanced control methods are applied.

The magnetic flow meter and the nuclear radiation type of density meter proved more accurate than earlier methods. Fluid density, electrical conductivity, boiling point elevation were found useful as inferential measurements. The application of thermal titrimetry and the neutron activation analyser meant faster analysis of caustic liquors and bauxite.

One of the new control concepts to be introduced was what is called feed-forward control. All principal factors affecting the process are measured, and along with the set point are used in computing the correct manipulated variable. The corrective action starts immediately, to cancel out the disturbance before it affects the controlled variable. Molar ratio control is an example to be dealt later for this concept. Another concept, namely the sampled-data feed-back trim is also used there, relying on a periodic analysis of output composition.

Whenever a single manipulated variable affects not one but two or more controlled variables, the loops are said to be coupled. To avoid one control loop to disturb the other, decoupling computing systems can be conceived, as shown later in the case of alumina liquor concentration controls. The idea is, that as a single valve can affect two controlled variables, so a single controller can be made to adjust two valves to decouple the system.

The computations made by above advanced control systems are material balances of the process (mass-, caustic soda- and alumina balances), and the manipulated variables are accurately regulated flow rates. A thorough

knowledge of the process is consequently required, which has been gathered by extensive work on mathematical process simulations.

The successful implementation of these advanced control concepts requires a lot of calculations. Analog computing elements, either pneumatic or electrical can be applied but for modest accuracy. The advent of microprocessors and digital signal processing could bring about a break-through in computerized process control.

Digital data storage and handling is already changing the scene of plant supervision and accounting as compared to integrating counters and strip-chart recorders still widely used in alumina plants.

2. SPECIAL INSTRUMENTS AND FINAL CONTROL ELEMENTS

2.1 Feed rate measurements of solids

The weight rates of bauxite and alumina hydrate can be measured with belt scales. With constant speed belts the flow is controlled by a gate to modify discharge of the feeder from the hopper to the belt. The gate can be controlled automatically so that a continuous regulation makes the feeder to deliver material at a desired weight rate.

Belt scales complete with automatic feeders of reputable make may best be purchased, but pick-up units to be built into existing conveyors are also available and can be fitted to the plant instrumentation system.

2.2 Flow rate measurements of liquids

2.2.1 Head flow meters in open channels

Higher flow rates have long been measured in open channels with no other method available for the time. Earlier these were the only means for metering liquor and slurry flows, and where high accuracy is not imperative some types can still be recommended. Among many known designs best results have been made with the V-notch and the approximately linear weirs. Latter having quadrant cutouts for side contraction over a horizontal crest, will be discussed in more detail, because they could easily be constructed in the plant.

Generally speaking, a weir plate is an obstruction in a flowing open stream over which the liquid is made

to pass. As a result the head on the upstream side is raised. This height of the stream decreases somewhat in the direction of flow, as the sheet of liquid passing over the crest is drawn downwards. The head is therefore measured some distance upstream of the plate, where this draw down effect is negligible.

The theoretical relationship between the discharge rate and the head can be derived for the various shapes of openings from basic laws of hydraulics. Practical weir flow formulae, however, are best used, because they have been suggested on results found in a large number of hydraulic experiments with the various standard designs of weirs.

The sizing formula for the approximate linear weir is given below because it is not easily available:

$$V=K(1,5\sqrt{2}\cdot g\cdot b\cdot h^{1,5}-Z\cdot d^{2,5}) \text{ dm}^3/\text{s}$$

where

V	is the flow rate in dm^3/s								
K	is the coefficient of discharge, dependent on h/d								
g	is the acceleration of gravity (a standard value of $g=98,9 \text{ dm}/\text{s}^2$ can be assumed)								
b	is width of the notch in dm								
h	is the head of flow over the crest in dm								
d	is the diameter of the cut out circle in the plate in dm								
Z	is a function of the ratio h/d as given below								
h/d	0,10	0,15	0,20	0,25	0,30	0,35	0,40	0,45	0,50
K	0,640	0,630	0,620	0,615	0,612	0,610	0,610	0,610	0,610
Z	0,108	0,238	0,416	0,643	0,912	1,222	1,571	1,956	2,373

The construction is shown in Fig.1.

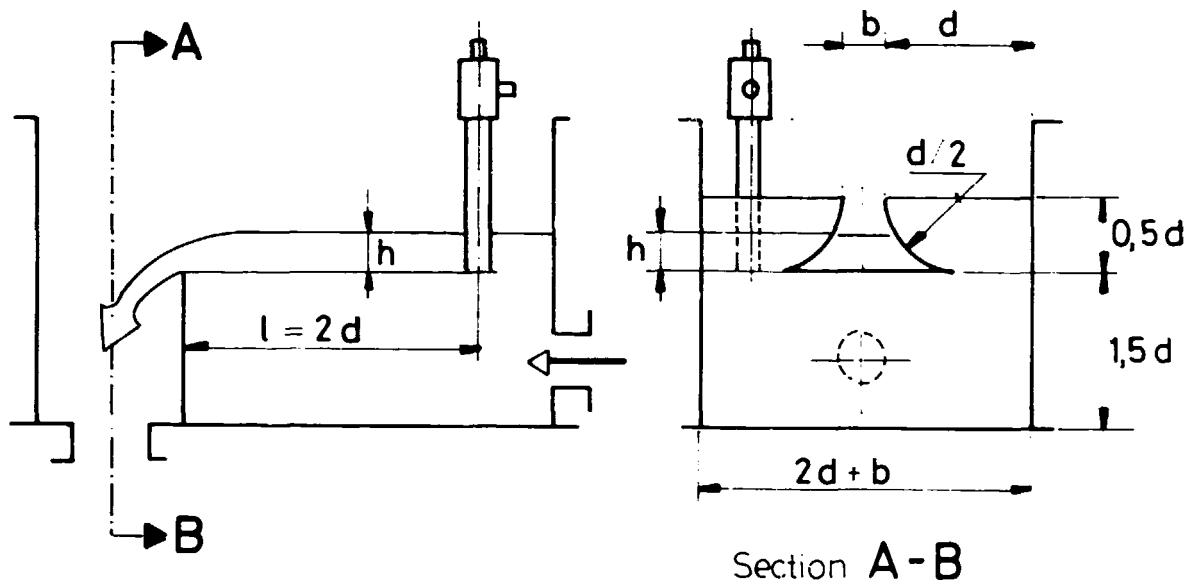


Fig.1 The approximate linear weir assembly.

The weir box or gauging tank holding the weir plate and the dip tube are considered as primary elements. There are a number of requirements to be observed to ensure satisfactory performance. The main dimensions of the weir are derived from recommendations as following.

Approach velocity to the plate must be low to minimize loss of head. Area of approach should therefore be at least six times that of the weir notch at maximum head.

The channel of approach should be deep enough for proper bottom contraction: A crest height 3 times of head is recommended. The horizontal distance from the side of aperture to the side wall of tank should be at least twice that of head to ensure adequate side contractions.

The point of head gauging should be at least 4 times of maximum head upstream from weir plate, to reduce draw-down of the liquid surface.

The outlet of the supply pipe should be under the liquid surface in the inlet chamber of the tank, to avoid splashing and bubbling. Baffle plates can be used between supply pipe outlet and gauging plate. The effects of disturbances are reduced and a more uniform velocity distribution is produced.

The bottom of the weir box should be flat, the side walls vertical and parallel to each other. The weir plate must be set vertically in the tank and perpendicular to the tank axis. The crest has to be kept level within 1,5 to 3,0 millimeters. The down-stream edge of the notch is mostly beveled off at a slope of 45° to the vertical. The upstream face should be smooth without projections.

2.2.2 Head flow meters in closed lines

Head meters operate by measurement of the pressure differential or head across a suitable restriction to flow in the pipeline.

The sharpe-edge concentric orifice plate is used for clear liquids like water, steam and compressed air. The eccentric orifice plate is the recommended flow restricting element to be used in horizontal pipe runs for metering various liquor flows. It can also be applied for slurry service because of low cost, and acceptable accuracy.

Orifice plates are made of corrosion resistant stainless steel. The upstream face of plates should be as flat as commercially available. The thickness at the orifice edge be about 3 to 5 millimeters. The edge shall be bevelled downstream at an angle of 45 degrees in case the plate would be thicker than permitted for its edge.

Bores of orifices must be sharp and round, and placed concentric to pipe diameter. The eccentric orifice, however, is constructed within a circle diameter about 0,98 times the internal pipe diameter, and placed at the bottom of the circle line.

Pressure taps directly at the up-and downstream faces of the orifice are called flange taps, and are preferred for use with concentric orifices as shown in Fig.2.

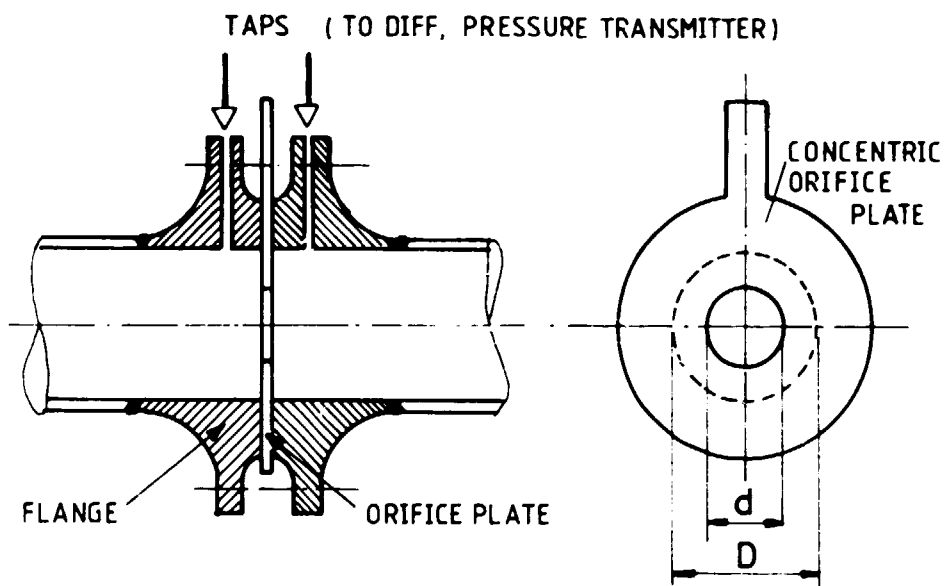


Fig.2 Concentric orifice plate with flange taps.

Pipe taps (or full pressure recovery taps) are used with eccentric orifice plates in alumina plant practice for metering caustic liquors and also slurries. The upstream and downstream taps are located 2,5 pipe diameters and 8,0 pipe diameters from the orifice plate. The tappings are suitably enlarged for the adaption of a sealing device, to be discussed later. The arrangement is shown in Fig.3.

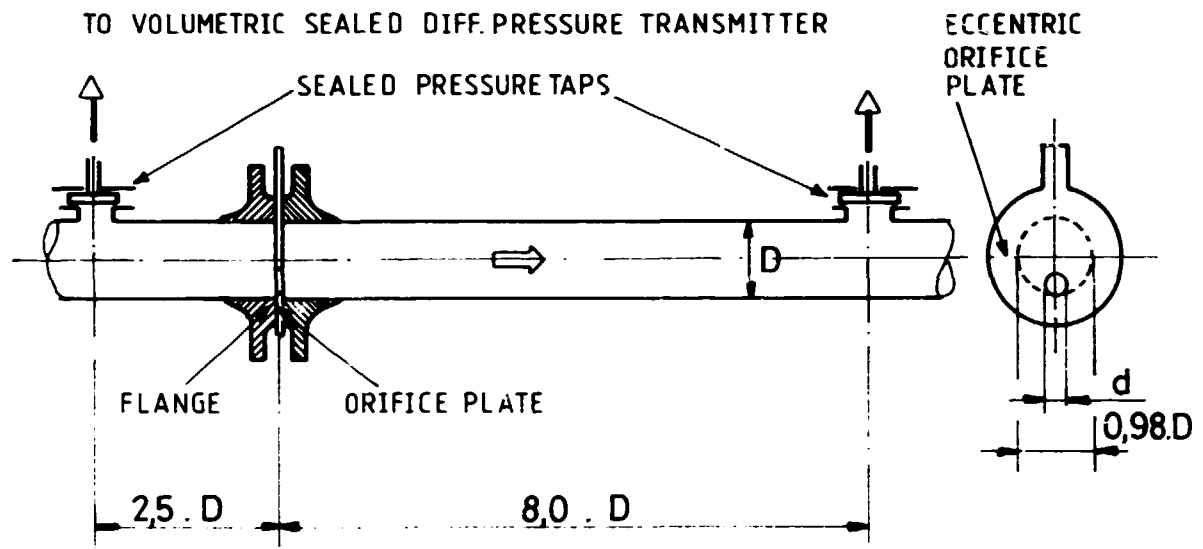


Fig. 3 Eccentric orifice plate with pipe taps.

The fundamental equation for orifice sizing is:

$$\alpha \cdot \beta^2 = \frac{V_{\max} \cdot C_o}{0,01252 \cdot F_a \cdot D^2 \cdot C_f \cdot h_{\max}}$$

were V_{\max} is the maximum flow rate in m^3/h
 h_{\max} is the maximal head in kp/m^2
 ρ_0 is the density at base temperature in kp/m^3
 ρ_f is the density at flow temperature in kp/m^3
 D is the internal pipe diameter of the line
in mm
 Fa is the thermal correction factor

The thermal correction factor can be calculated for the temperature ranges as met in alumina plants and for steel as material of the primary element from the expression:

$$Fa = 1 + 2\lambda(T - T_0)$$

T being the actual flow temperature in $^{\circ}C$
 T_0 is the base temperature, mostly $15^{\circ}C$
 λ is the linear thermal expansion factor, for steel it is 0,000012 between flow temperatures 0 to $250^{\circ}C$, and 0,000013 between flow temperatures 250 to $450^{\circ}C$.

With eccentric orifices and pipe taps it is desirable to arrive at a value of $\beta^2 \cdot \alpha$ between 0,025 and 0,2900 for acceptable accuracy in a broad metering range. First h_{\max} could be chosen freely to meet this requirement, because most differential pressure transmitters can widely be adjusted, mostly between 500 and 10000 kp/m^2 pressure differences. In case this would not suffice, the line diameter D could be changed, mostly by reducing to a suitably narrower pipe diameter.

After having settled $\beta^2 \cdot \alpha$, its components β and α can be found by trial using the expressions

$$\alpha = 0,58925 + 0,2725\beta - 0,825\beta^2 + 1,75\beta^3, \text{ or}$$

$$\beta^2 \cdot \alpha = 0,58925 \cdot \beta^2 + 0,2725\beta^3 - 0,825\beta^4 + 1,75\beta^5$$

Having found β the bore diameter can be calculated by

$$d = \beta \cdot D$$

For concentric orifices with flange taps the expression for resolving $\beta^2 \cdot \alpha$ into its components is somewhat different:

$$\alpha = 0,598 + 0,01\beta + 0,0001947(10 \cdot \beta)^{4,425}, \text{ which would}$$

result in lower α contraction coefficients than with pressure taps. That means larger d constriction diameters with flange taps would cause equal differential pressures than smaller bores with pipe taps.

2.2.3 The magnetic flow meter

Head meters generate differential pressures which are conveyed by means of pressure taps to the secondary element. Their reliability in alumina plant service is impaired by the fact, that liquors and slurries tend to clog the taps by sediments and scale. To overcome this difficulty other types of flow meters have been tried in the field, which do not require any pressure taps. One of them may be considered as the best answer till now for liquor and slurry service in alumina plants.

The magnetic flowmeter measures volume rates of flow without any restriction and tappings. It operates on the principle of an electrical power generator, based on Faraday's law of electromagnetic induction. Simply stated the principle is, that when a moving conductor cuts across lines of force in a magnetic field, a voltage is induced in it. The conductor is the flowing liquid. The length of the conductor is determined by the spacing between the two electrodes across which the output voltage appears. The magnetic field is provided by a current flowing in coils, wrapped around the flowmeter tube.

First experiments run with early magnetic flow meters were not fully satisfactory. But after more than a decade of extended factory use the magnetic flow meter proved to be the most reliable device with best long-time accuracy for slurry and caustic liquor service.

2.3 Level measurements in tanks

2.3.1 Level measurement in open tanks

Level measuring instruments operating on open vessels find wide application in the plant. The pneumatic bubbling type employing a flow of air is used with advantage for head measurement.

Theoretically the height of liquid above a pressure gauge can be inferred from the reading of the gauge, provided the density of the liquid is known and constant. In alumina plants, however, the density of liquors and slurries is changing to a certain degree, and only average figures are known. Accuracy in terms of level is modest, but application of the method is still considered adequate for practical purposes.

A supply of compressed air is piped to the dip tube by a pressure reduction valve, to reduce air pressure a little above that of the maximum head being measured. A flow adjusting valve and an air flow gauge facilitate adjustment of a slow bubbling air purge. Air pressure builds up exactly to balance the head pressure, any surplus escapes through the liquid.

The back pressure is piped to the input connection of a suitable pressure transmitter. With this system the location of the pressure instrument in relation to the point of measurement is not limited. The longer the distance, however, the slower the response to changes in level.

Levels and liquids with large amounts of suspended matter can be measured. The probes are best immersed only to the expected minimum level in the liquid, not too near to tank bottoms, where sediment build-up may clog the end of the bubbler pipe. (Refer to Fig.4)

2.3.2 Level measurement in closed vessels

Liquid levels can also be measured in closed tanks under pressure using the head gauging method. Instead of one dip-tube and a pressure transmitter as in open tank service, two taps are piped to a differential-pressure transmitter. The upper tapping point is made at least as high as the maximum level to be measured, and the lower one at least at the height of the expected minimum level.

A constant reference head pot is applied at the high level tapping for service with clear liquids. The

differential between the constant reference head and the variable level remains active on the transmitter input only, because the static pressure in the tank above the liquid is added to both sides of the transmitter and so its effect is canceled out. The differential falls as the tank level rises, and the normal action of the transmitter must be reversed if a rising level is to give an increasing output signal.

The arrangement described is quite satisfactory with clear liquids, but for slurry or liquid service a pair of oil seals as shown in Fig.5 are required to prevent clogging.

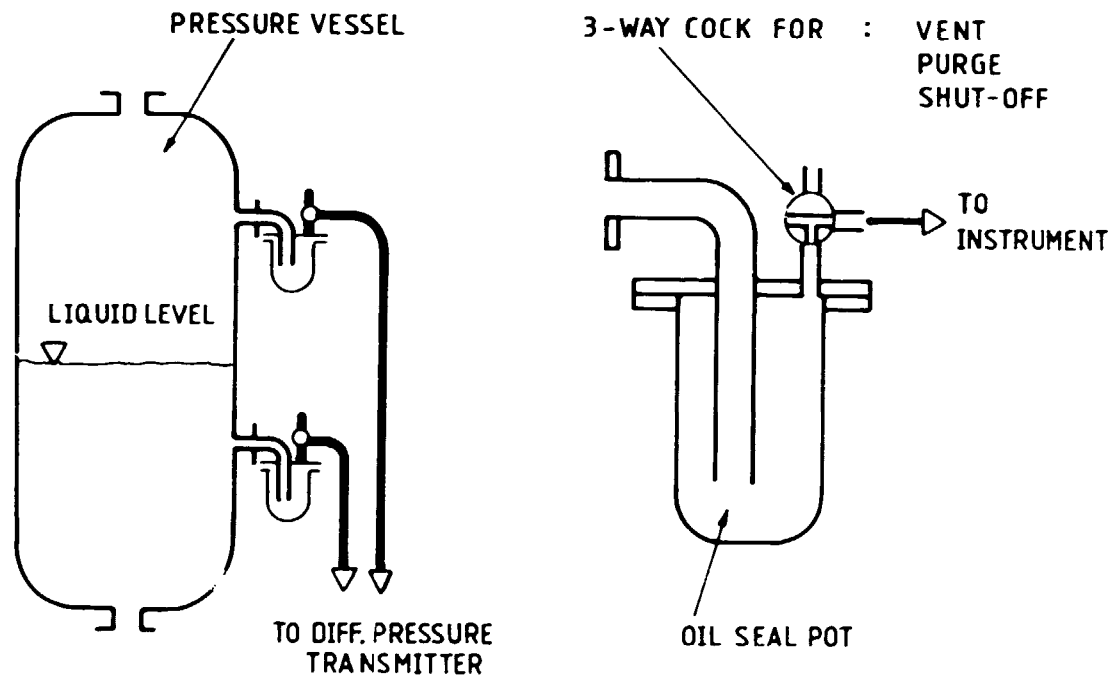


Fig.5 Oil seal pot applied for level measurement.

For the latter 40 mm diameter taps are made, with 20 mm shut-off valves in the oil line. For water service 15 mm taps and shut-off valves are adequate.

2.3.3 Radiation-type level measurement

The nuclear radiation-type level gauge has gained popularity in recent years. Its chief advantage is that no part of the device comes into contact with the material being measured. For this reason it is the best method for level measurements in digester flash tanks, where the properties of the fluid do not permit a trouble-free use of the differential-pressure gauging. It finds application also on bins for hydrate and alumina storage.

The nuclear level gauge uses a source of gamma radiation, mostly C_{54}^{137} and C_{27}^{60} cesium- and cobalt isotopes, located on one side of the vessel and a detector mounted opposite. The detector is a Geiger counter, which converts gamma radiation received into electrical quantities by amplification, integration and rectification. The intensity of the radiation received by the detector depends upon the mass of material between the source and the detector. An increase in level will thus cause a corresponding decrease in the radiation received and in the resulting d-c current produced in the detector.

2.4 Density measurements of fluids

2.4.1 Density metering by bubbler assembly

The arrangement used for density measurement includes elements for drawing the sample, creating the heads

of primary differential and of compensation for zero suppression, as shown in Fig.4 along with the bubbler dip tube for liquid level measurement as described earlier.

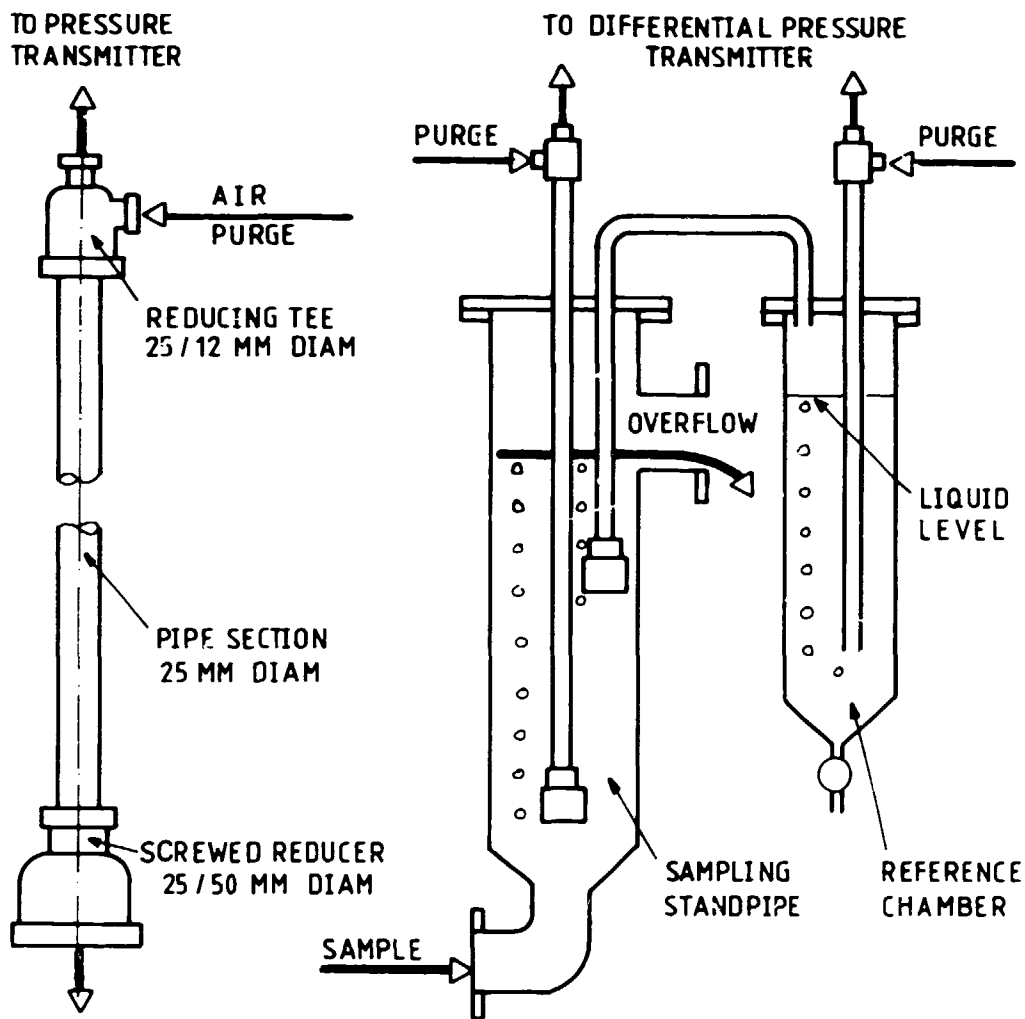


Fig.4 Air purged level and density meter.

Sampling is made by applying a standpipe of large diameter across a low fluid head, to ensure that a low fluid velocity will flow through it. Both full flow and by-pass arrangements are used. A pair of air purged dip tubes immersed to different elevations are mounted on the standpipe.

A third bubble pipe is engaged within the so called reference chamber. Oil in this chamber adds a new head to the back-pressure of the shorter dip tube. The chamber itself is a closed vessel having its own bubbler tube, pressure-taps and drain plug.

Air is purged through a feed sight to the reference chamber, the tubing connected to the inlet for the bubbler pipe. The back pressure is lead from a connection at the air filled portion of the chamber to the shorter dip-tube immersed in the process fluid.

Another sight feed bubbler and the longer dip-tube create the head which is taken to the high-pressure input of the differential pressure transmitter.

2.4.2 Density measurement by gamma-ray absorption

Density is mostly the inferential variable for solids concentration in liquids, as in cases of diluted digester blow-off or settler underflow. Recently the gamma-ray absorption measurement as described under radiation type level gauging is also used in alumina plants for density metering, and this method is gaining popularity. The radiation source is mounted on one side of the pipe line, the detector unit on the opposite side. A vertical line

section is good to ensure full cross section flow through the pipe. The absorption is proportional to fluid density and also to concentration of solids, so the output signal range can be calibrated to any of these variables.

2.5 Measurement of boiling-point rise

The boiling-point elevation of liquors is inferred to their caustic soda concentration in the arrangement shown in Fig.6. The temperature of the boiling liquor in the last evaporator stage is compared with that of boiling water at the same pressure. The reference chamber has a pressure equalizing and an overflow connection to the vapor space of the evaporator vessel. A small flow of condensate is flashed into the chamber. One resistance thermometer is immersed into the boiling water, the other one into the boiling liquor. The output signal of the differential temperature transmitter can be calibrated in terms of caustic soda concentration. The measurement is used to monitor the end point of evaporation.

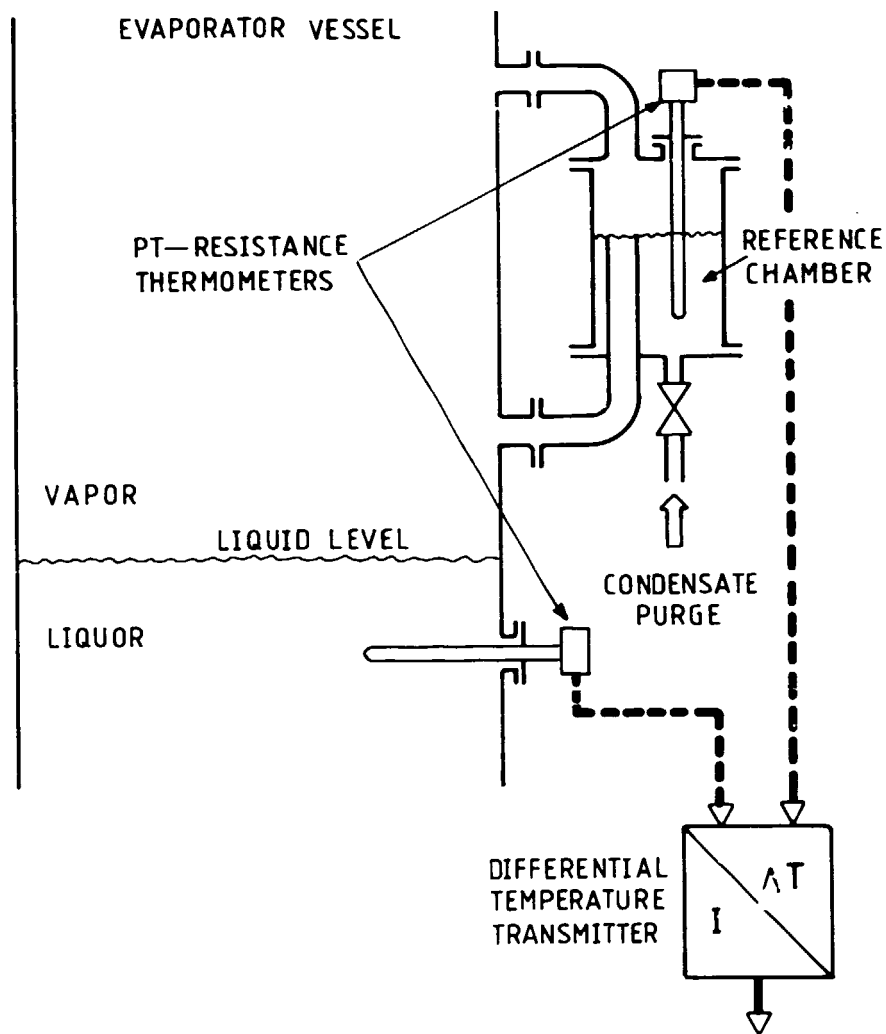


Fig.6 Measurement of boiling point elevation

2.6 Oscillometric conductivity measurement

The arrangement shown in Fig.7 is applied to monitor caustic soda and alumina concentrations of plant liquor. The operation is semi-continuous, because the two con-

centrations are metered sequentially.

The liquor sample is made to pass the overflow vessel. Upon opening of valve 3 the sample flows through the cooler 2. The temperature of the liquor at the cooler outlet is controlled automatically by adjusting the water valve 4. The sample fills the cell 7, which contains a circuit coupled to a 4 megacycle oscillator of the transmitter 6. The ratio of circuit reactance to circuit resistance will change with the conductivity of the sample in the cell.

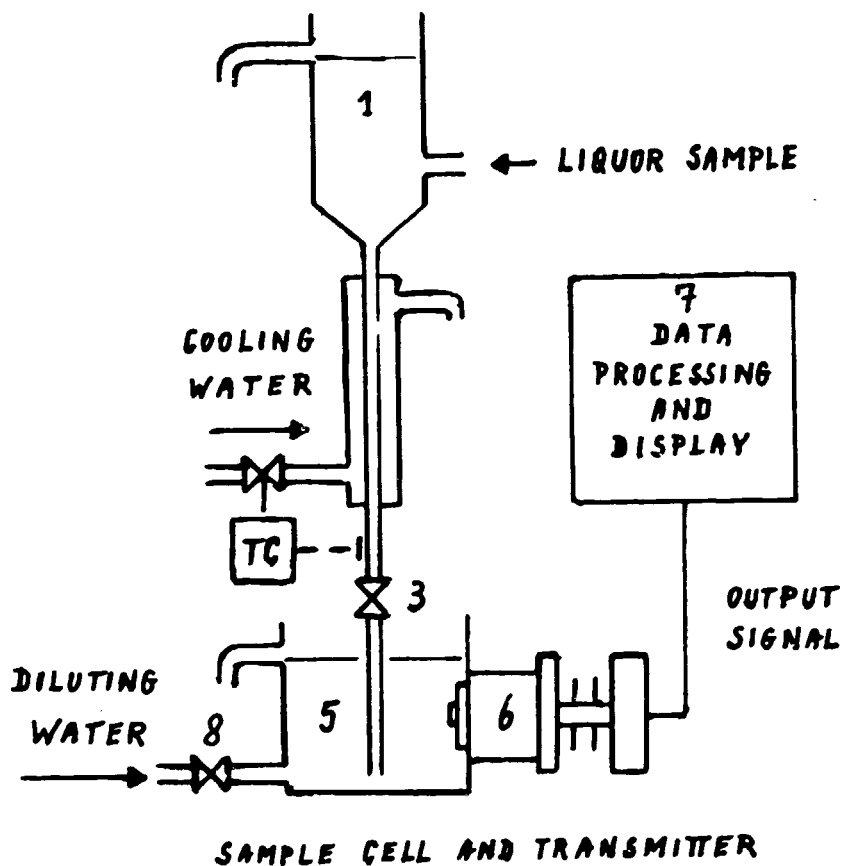


Fig.7 Oscillometric conductivity measurement

The conductivity of the undiluted sample is measured and stored. Upon shutting valve 3 and opening valve 8 an isothermal dilution is taking place in the sample cell 5, which would slowly change the conductivity of the original sample. At a certain dilution conductivity would reach its highest value, and decrease on further addition of water. The peak in conductivity is also stored. The data processing unit calculates caustic soda and alumina concentrations from the stored data of the two measurements.

Variations of the method described are being used in applications for monitoring digestion and alumina liquors, and even digester blow-off concentrations.

2.7 Control valves and actuators

The control valve is perhaps the most critical part of the control loop. Great care is required at the selection of a valve for any particular application. Different types of valve body designs can be recommended for typical applications in alumina plants.

Single and double port globe valves are widely used with clear fluids, like water, steam, compressed air, natural gas and fuel oil. While single port valves are adequate for smaller flow rates and moderate pressure drops, double seated types can handle higher flow rates and pressure drops. Latter types are reversible, so that plugs may move to open or to close with increasing loading pressure on the same type of actuator, what can be used to advantage.

In larger sizes, however, globe valves become expensive, so the butterfly valves may be used, especially for air or gas service. Their installation requires minimum

space, but operating torques may be high and huge actuators would be needed.

Angle body valves are self-draining single ported valves and are extensively used to handle slurry flows. Flashing slurries in the heat recovery section of digester string present severe operating conditions. Flow direction is from the plug side through port to discharge end of valve. The outlet section should be lined to withstand eroding effects. Hard surfacing of port and throttle plug is considered imperative. Flashing calls for larger port and plug diameters, which combined with appreciable pressure drops results in high stem forces. Large size pneumatic actuators and higher loading pressure ranges are required.

The widely used slant port valves are also self-draining. They can be mounted into straight pipe runs, and are applicable for slurry and liquor service under modest pressure drops and non-flashing flow conditions.

Another type of special control valve found in alumina plants for applications where low line pressure and small pressure drops prevail is the Saunders-valve. A diaphragm forced down by the stem toward an internal wedge restricts the flow of the process fluid, resulting in a non-clogging design. Operating temperature is limited by availability of diaphragm material. High stem forces needed for closure require larger pneumatic actuators and higher loading pressures.

Recently Vee-ball valves have gained some popularity in alumina plant applications. Basically similar in con-

struction to conventional ball-valves, they exhibit a V-notch cut in the ball and a specialized body for hard to handle fluids. The notch results in a modified parabolic flow characteristic.

All valve assemblies have movable parts to provide a variable restriction to flow. The inherent flow characteristic of any movable part or plug is the relation between flow rate through a valve and valve plug travel, when a constant pressure differential is maintained across the valve.

One common style is the linear flow characteristic, where the flow rate is directly proportional to stem travel. The change of flow rate with respect to stem travel is constant.

The other widely used type is the equal percentage flow characteristic. For equal increments of stem travel the change in flow rate with respect to stem travel can be expressed as a constant percent of the flow rate at the time of change. So the change of flow rate to stem travel will be smaller when the valve plug is near to close the port, and higher with the valve nearly open.

It is a good rule to select linear flow valves for applications where flow rates may widely change, while corresponding pressure differentials remain nearly constant, as for example between flash tanks of digester string.

Wherever pressure drops are due to change appreciably while flow rates handled remain within normal limits, as in systems with centrifugal pumps, the equal percentage characteristic should be chosen.

By selecting the proper characteristic for the process application, a uniform gain of the control valve will result for the whole operating range, which is important for control loop stability.

The most common valve actuator in the alumina plant is the air operated diaphragm. Mostly the direct acting type is used, where air pressure moves stem down against the spring. In rare cases the reverse acting type is needed. In any case the actuator will move the stem until the force developed by the air pressure on the diaphragm is equal to the sum of the forces developed by the compression of the spring, the process fluid acting on the plug and by stem friction. If the stem fails to move exactly to the position as called for by the input signal, the positioner regulates the loading air pressure to overcome extra forces mostly caused by stem sticking.

2.8 Sizing of control valves

The selection of the proper valve type depends on the nature of the fluid to be controlled and the circumstances of the application as discussed formerly. But to decide what size the valve body should be and exactly what port diameter the valve should have is also a very important task, because the performance of the control loop will greatly be affected by the correct sizing.

The following information about process conditions are required for sizing:

- 1.) Nature of fluid (caustic concentration, solids content),
- 2.) Maximum flow rate in m^3/h or t/h ,

- 3.) Valve inlet pressure in kp/cm^2 ,
- 4.) Valve outlet pressure in kp/cm^2 ,
- 5.) Temperature at valve inlet in $^{\circ}\text{C}$,
- 6.) Specific gravity or specific volume in t/m^3
resp. m^3/t ,
- 7.) Vapour pressure at flowing temperature.

The basic equations used for sizing are:

$$V_{\max} = C_f \cdot k_v \cdot \sqrt{\frac{P_1 - P_2}{\rho}} \quad \text{m}^3/\text{h}, \text{ and}$$

$$m_{\max} = C_f \cdot k_v \cdot \sqrt{\frac{P_1 - P_2}{V}} \quad \text{t/h},$$

where V_{\max} is the maximum volume rate in m^3/h
 m_{\max} is the maximum mass rate in t/h
 P_1 is the pressure at valve inlet in kp/cm^2
 P_2 is the pressure at valve outlet in kp/cm^2
 ρ is the specific gravity in t/m^3
 V is the specific volume in m^3/t
 k_v and C_f are empirical coefficients determined experimentally for flow conditions of the open valve

k_v is the coefficient indicating the flow rate capacity of the valve, depending mainly from the port size of the open valve. Its magnitude states, how many m^3/h of water would flow through the valve at a pressure difference of 1 kp/cm^2 .

The application of the Bernoulli theorem for the conditions given in the flow through the valve port results in:

$$k_v = \alpha.5,04. \frac{d^2 \pi}{4}$$

where d is the port diameter in cm, and α is the coefficient of flow contraction, which is about 0,61 for sharp edge restrictions. So k_v turns out to be closely related to the port area and the diameter of the valve port:

$$\frac{k_v}{3,08} = \frac{d^2 \pi}{4} \quad \text{or} \quad k_v = 2,42 \cdot d^2,$$

which are very useful equations, when the k_v coefficient is not available for a control valve of known port size, or to determine the port diameter for any k_v required.

C_f is the pressure recovery coefficient, which depends mainly upon the flow pattern through the valve. By definition

$$C_f^2 = \frac{P_1 - P_2}{P_1 - P_{vc}}, \quad \text{where}$$

P_1 and P_2 are the pressures at valve inlet and outlet, and P_{vc} is the pressure in the "vena contracta" i.e. at the highest flow velocity in the port section. P_{vc} is a lower pressure than P_2 , as expressed by

$$P_{vc} = \frac{P_2 - P_1(1 - C_f^2)}{C_f^2},$$

and eventually might become lower than the saturation pressure related to the fluid temperature T_1 , which would cause flashing. C_f becomes smaller as the flow pattern through the valve becomes more streamlined, as visible from typical C_f -values shown below for orientation:

Body design	C_f
globe valve	0,80
angle body	0,72
slant port	0,62
Vee-ball	0,58
butterfly	0,50

To clarify what differential pressure $P_1 - P_2$ has to be used in the sizing formula, a pumped fluid flow system is to be considered. The relationship between pressure and flow rate is visible in Fig.8.

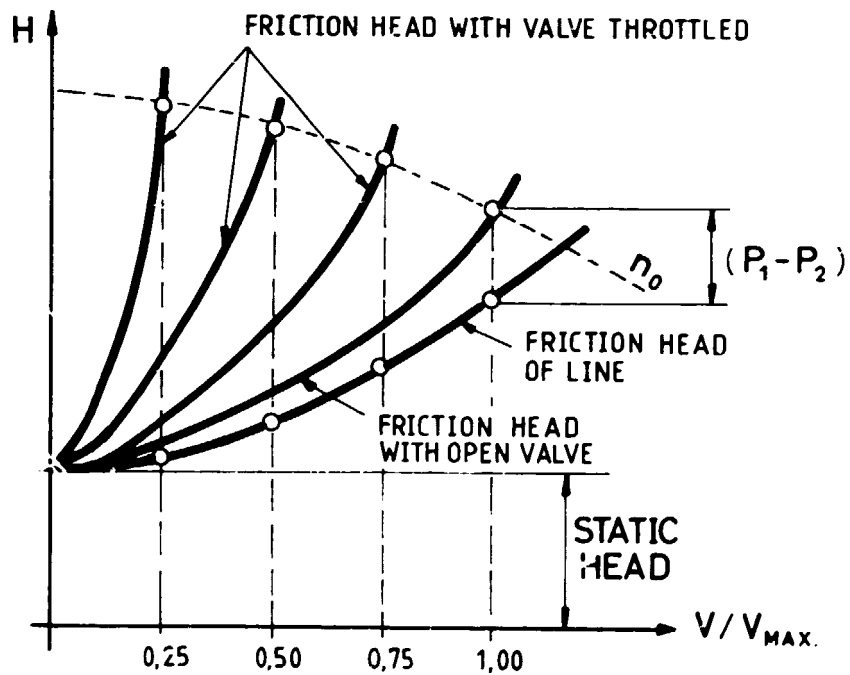


Fig.8 Pumped system with throttling valve

From the plot it becomes clear that a lot of information is needed to arrive at the actual $P_1 - P_2$ at V_{\max} . The actual pump characteristic, the friction head of the connecting piping and the static head of the elevation must be known fairly well.

The open valve pressure drop $P_1 - P_2$ is the difference between the pump head and the friction head at V_{\max} rated flow. The combined friction head curve of piping and open valve cuts the actual pump characteristic just at the maximum flow rate. The throttling curves become ever steeper as the valve is closed successively. The pump line is cut at lower flow rates, while the pressure difference absorbed by the valve increases.

Flashing will occur, whenever the pressure drops below the saturation pressure of the fluid in its passage through the throttling restriction. The sizing of valves for flashing fluids is based on empirical formulas.

In applications for level control of saturated fluids as in flash tanks of digester strings, with the control valve in the discharge line, the maximum mass rate will be

$$m_{\max} = k_v \cdot C_f \cdot 0,218 \cdot P_1^{0,672}$$

for a wide range of relative pressure drops of

$$0,14 < \frac{P_1 - P_2}{P_1} < 1,00$$

When, however, the relative pressure drop is as low as

$$0,05 < \frac{P_1 - P_2}{P_1} < 0,14$$

the mass flow rate can be computed from

$$m = m_{\max} \cdot 2,246 \cdot \left(\frac{P_1 - P_2}{P_1} \right)^{0,407}$$

- where m_{\max} is the maximum flow rate in t/h,
 m is the mass flow rate in t/h,
 P_1 is the pressure at valve inlet in kp/cm^2 gauge,
 P_2 is the pressure at valve outlet in kp/cm^2 gauge.

2.9 Variable speed pump drives

The energy wasted by throttling can be conserved by applying speed regulation for pump drives instead of control valves. Pump characteristics at lower speeds cut the friction head curve of the piping at decreasing flow rates as shown in Fig.9. So the same flow rates

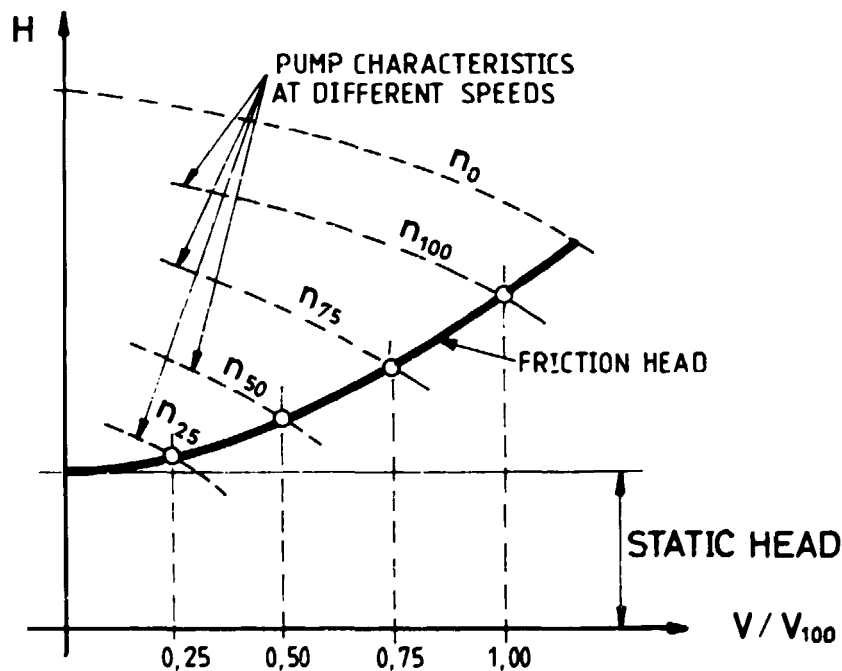


Fig. 9 Pumped system with speed regulation

can be adjusted by speed regulation of the pump as by valve throttling, but with an average of about 20 percent less energy consumed.

Both the hydraulic couplings and the variable frequency drives were applied to advantage in combination with squirrel cage induction motors. Hydraulic couplings are rugged and cheap, but their efficiency deteriorates with increasing slip at lower output speeds. A disadvantage is, that separate couplings must be used for each motor.

With variable frequency drives one control device can be used for a pair of pump motors, where one is operated at a time with the other one as stand-by unit. All though costlier than the couplings, variable frequency drives result in high efficiencies over a wide speed range.

By the elimination of control valves their maintenance costs can be saved, which might be considerable with abrasive slurries. Additionally the life time of pump impellers would double in the average due to reduced speeds.

3. BASIC INSTRUMENTATION OF UNIT OPERATIONS

3.1 Instrumentation and control of wet grinding

Purpose of the system shown in Fig.10 is to maintain the bauxite feed rate, control the total flow of liquor to result in the required molar ratio of the digester blow-off, and to keep a suitable solid to liquid ratio in the ball mill.

The bauxite feed rate is set manually by a mechanical gate feeder.

The resulting weight rate is measured by the belt scale WRS.

A weir assembly as shown was used in earlier applications to control the total liquor flow rate by FRSC. A magnetic flow meter would give higher accuracy

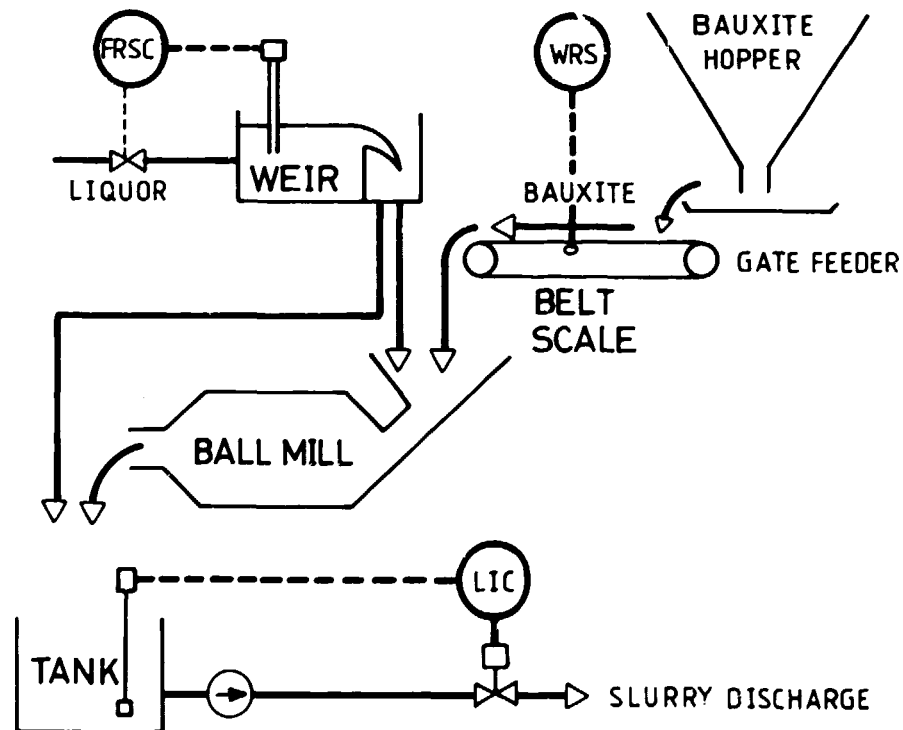


Fig.10 Instrumentation and control of wet grinding

3.2 Instrumentation and control of digestion

The process instrumentation and control diagram as shown in Fig.11 has the following features:

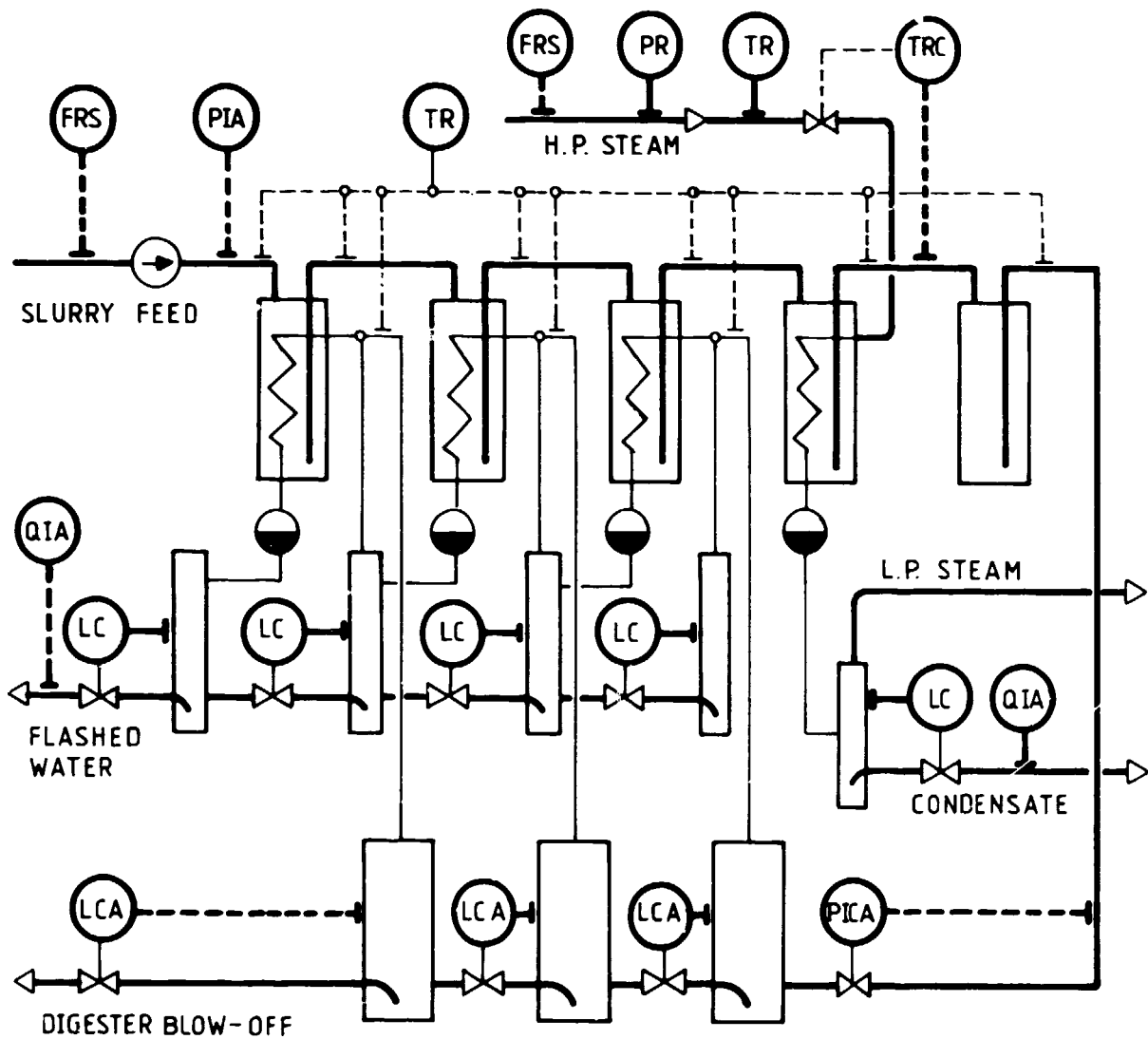


Fig.11 Process instrumentation and control of digestion

Digestion temperature is automatically controlled with TRC by adjusting steam flow. The pressure control PICA at digester discharge prevents boiling in digesters. Special control loops LCA and LC control levels in flash tanks. Caustic soda concentration of flashed water and condensate are checked by QIA conductivity measurements. Steam consumption and slurry feed rate should be metered by FRS loops for supervision of performance.

3.3 Instrumentation and control of clarification

Characteristics of the instrumentation and control scheme as shown in Fig.12 are as follows:

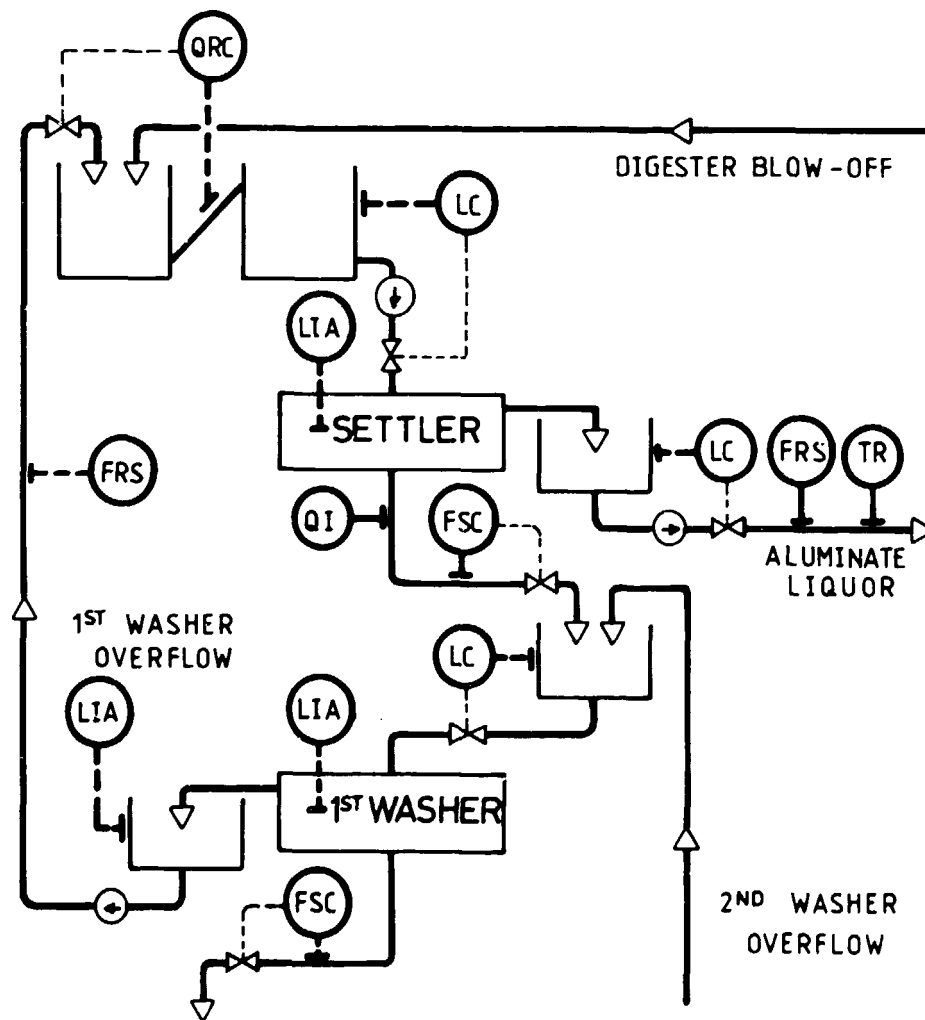


Fig.12 Process instrumentations and control of clarification

Wash water has to be added proportionally to changing digester blow-off loads, to get a suitable dilution for effective separation of red mud from alumina liquor. The density controller QRC is applied to this purpose.

The underflow of each thickener is controlled by FSC to maintain the steady countercurrent flow of solids. (Constant head control with a weir flow meter would compensate varying solids content in underflow). To check solids content in settler underflow a density gauge QI may be used.

The interface level between thickened slurry and liquid overflow has to be monitored by LIA meters, to check inventory of solids in each settler. When interface levels are found as beyond a certain range, the flow rates of related underflow controllers FSC must be reset appropriately.

Wash water flow rate to be fed into the countercurrent decantation at the last washer unit can be set to keep level in first washer overflow storage tank within certain limits. The level gauge LIA indicates wash water inventory.

Flow measurements of aluminate liquor and diluting water (FRS) are for process supervision:

3.4 Instrumentation and control of evaporation

The evaporation rate will change within periodic wash cycles due to deteriorating heat transfer caused by scaling and fouling. To keep the strong liquor concentration constant, the load has to be adjusted.

As shown in Fig.13, the liquor feed and the steam consumption rates (FRS) are set appropriately and recorded together with output quality (QR), inferred from measurements of either boiling point elevation or density.

Automatic controls LC keep optimal liquor levels in evaporators for best performance.

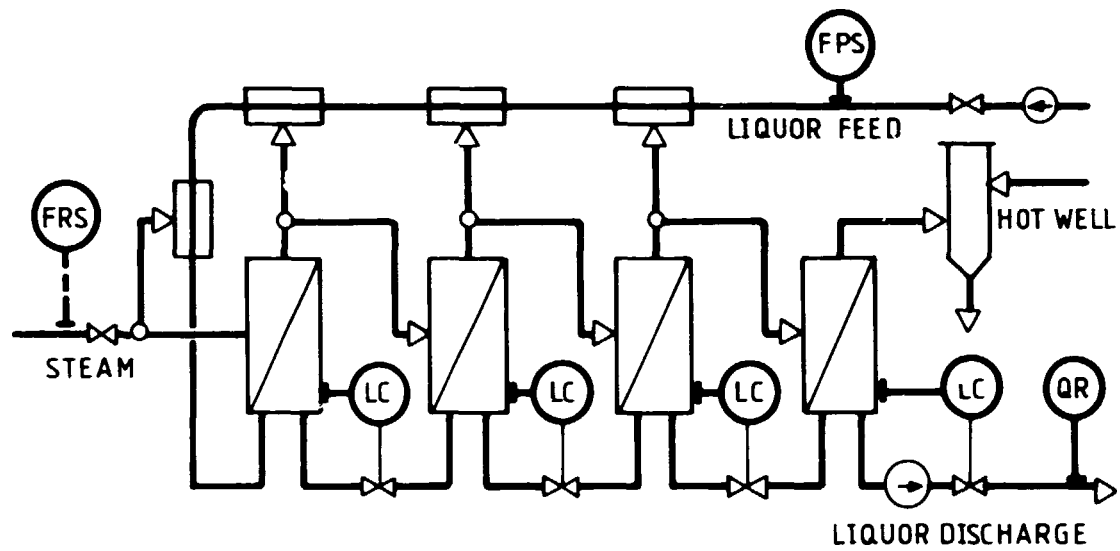


Fig.13 Instrumentation and control of evaporation

3.5 Process instrumentation of calciner kilns

The instrumentation concept is shown in Fig.14.

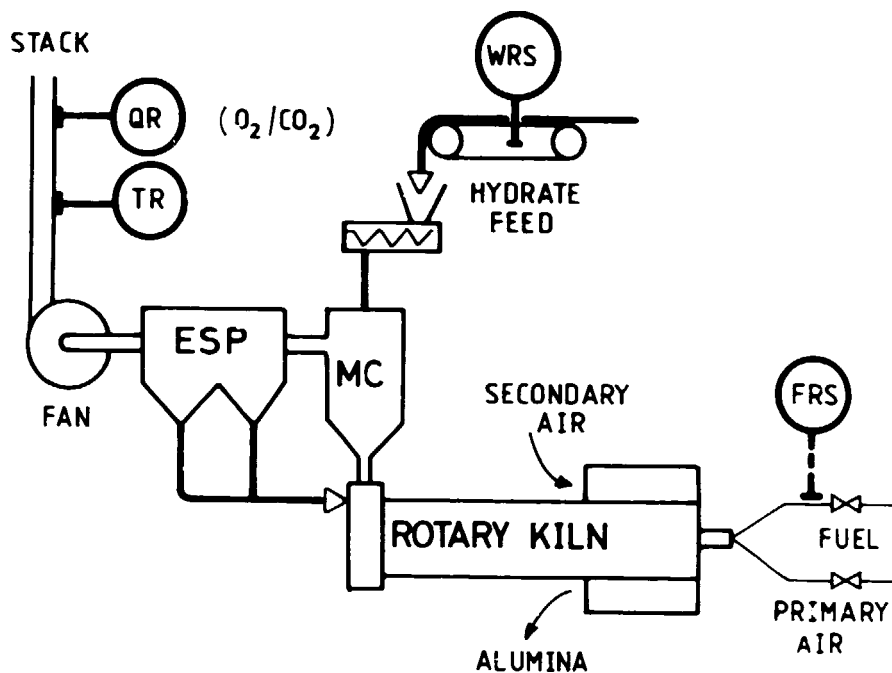


Fig.14 Process instrumentation of a calciner kiln

The fuel heat input required per ton of alumina calcined for any kiln will be fairly constant. The heat balance will be stable if constant feed rates of both hydrate and fuel are maintained.

Kilns are best operating at stable loads, wherefor the hydrate feeder is manually set and the resulting feed rate is metered by WRS. The fuel input metered by FRS is also adjusted to keep the specific fuel to hydrate ratio.

Stack losses are monitored by TR temperature- and QR composition readings. Acceptable excess air ratios in flue gas must be kept by combustion air flow adjustments. Many burners feature an automatic fuel to air ratio control.

4. ADVANCED CONTROL CONCEPTS

The flow diagram in Fig.15 shows the process technology, as practised in many alumina plants. The economy of the process depends mainly upon the product output and its quality, and upon the costs of production. Major portions of production costs are linked to the consumption of bauxite ore, caustic soda and energy. Energy is required to heat the process and to drive the circulating process fluids.

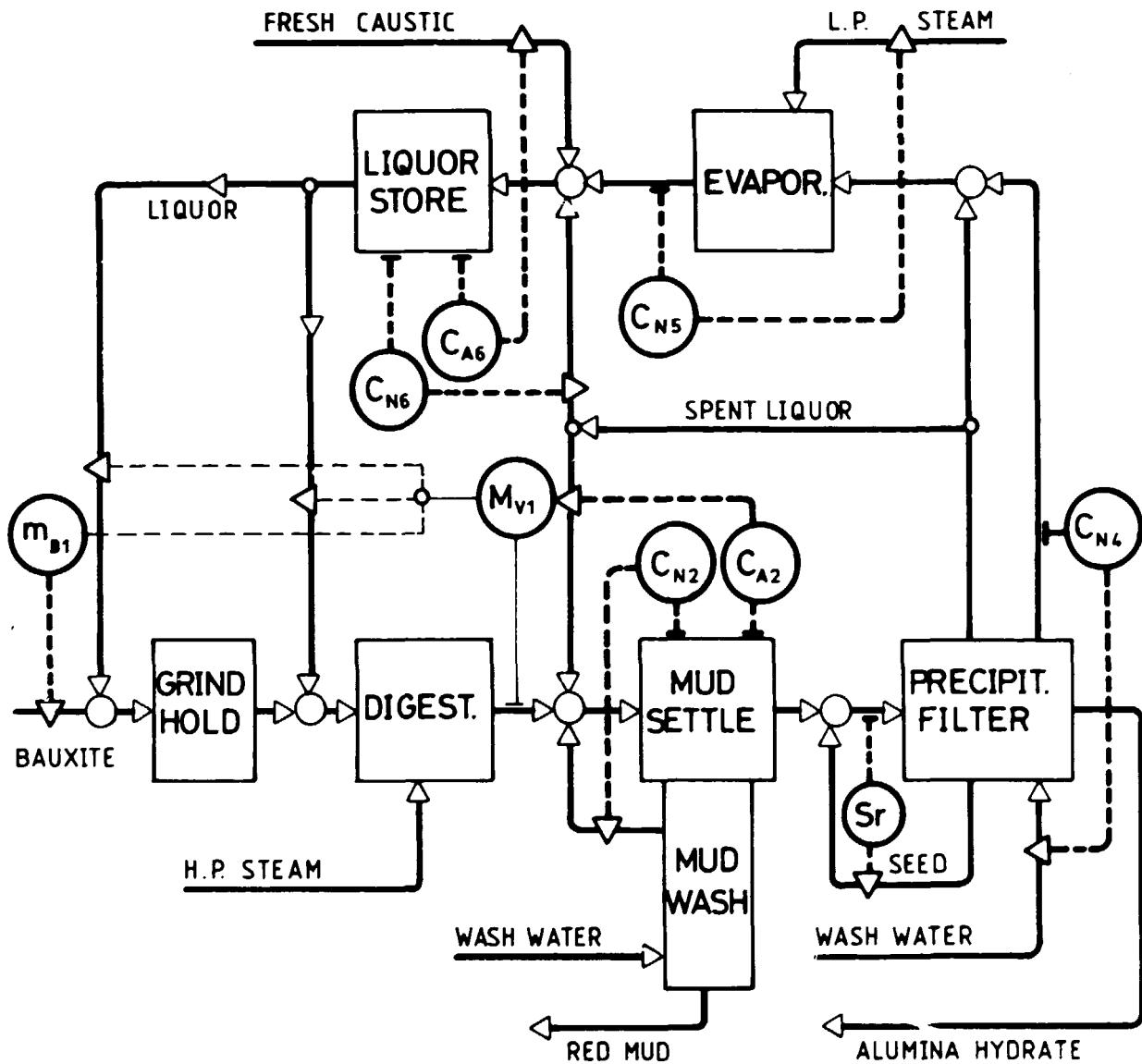


Fig.15 Concentration controls in the Bayer-process

Product quality depends upon particle size, size distribution and purity of alumina produced. Quality and production rate as well as the specific consumption of bauxite, caustic soda and energy will be determined to a certain extent by the existing process equipment and the composition of the available ore. On the other hand, the way the plant is actually run could have a strong influence upon process economy.

A better plant use could be achieved by holding a number of important process variables at predetermined values. The following concentrations are considered to have a decisive influence upon production economy of the Bayer-cycle, and should be controlled automatically within narrow limits:

- Molar ratio Na_2O to Al_2O_3 in digester blow-off (M_{v1})
- Na_2O - and Al_2O_3 concentrations in aluminate liquor (C_{N2} , C_{A2})
- Seed-ratio (Al_2O_3 in solids to liquid) for precipitation (S_{r3})
- Na_2O -concentration at hydrate wash filtration (C_{N4})
- Na_2O -concentration after liquor evaporation (C_{N5})
- Na_2O -concentration of digestion liquor (C_{N6})

A fundamental knowledge of the process is required, because the functions of advanced control systems are derived from the material and energy balances of the alumina processing operations.

The general concept is to compute the optimal settings for above important process variables, for any new situation of equipment availability, production rate demand, raw material and energy costs and selling prize of the product.

Once the optimal settings are found, the particular control loops will stabilize them at the required levels.

4.1 Molar ratio control in digestion

(Process control diagram in Fig.16)

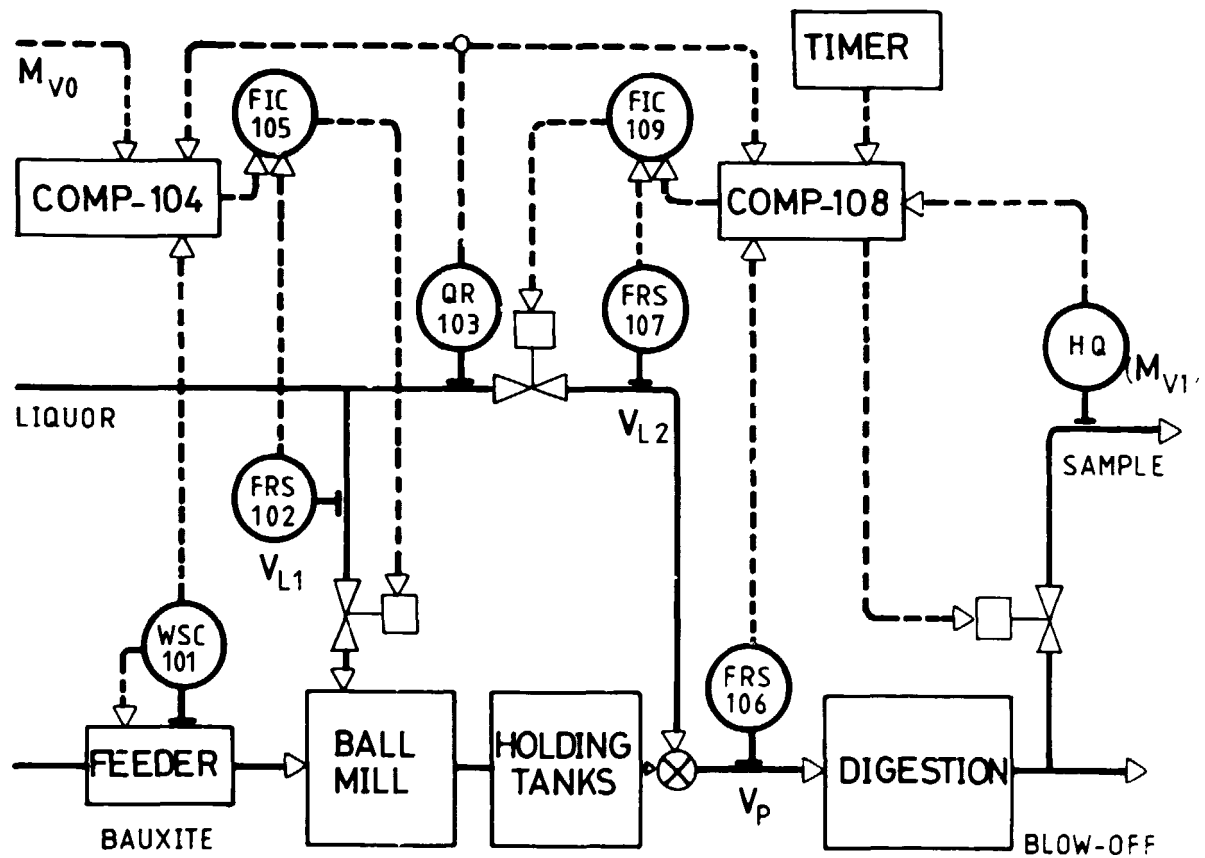


Fig.16 Molar ratio control in digestion.

The molar ratio (M_{V1}) at the discharge end of the digestion has a strong influence upon plant economy. Liquor shortage would decrease M_{V1} , resulting in loss of extractable alumina and consequently higher bauxite consumption. Liquor overcharge indicated by M_{V1} higher than optimal, would by the increased volume rate of the digester feed require more heating steam. Additionally the extraction output of the digester string would be less. The optimal setting for M_{V1} is close to the alumina saturation point for the soda concentration and temperature of digestion, but by a safety margin of a few percents above it.

Even small deviations from the optimal M_{V1} effect the process economy significantly. High control accuracy is required therefore, which can only be achieved by special control methods because of the severe process dynamics.

Main features of a practicable control strategy as shown in Fig.16 are:

- 1.) Bauxite weight rate, liquor volume rate, soda and alumina concentrations of the liquor are measured at the feed end of the process unit. An average level of the extractable alumina content of the bauxite is assumed.
- 2.) Changing bauxite composition is the most important disturbance. The changes are rapid and of stochastic nature, but the mixed tanks of the desilication unit will smooth out the disturbances, so only slow changes may reach the digester feed.
- 3.) The changes at the digester blow-off are compensated by feed-back control. Periodic checks of the molar ratio in the digester blow-off are quite adequate to set the correctiv liquor flow rate (\dot{V}_{L2}) in a mode of sampled-data control.
- 4.) Because of the different holding times of the two subunits desilication and digestion the delays in control response would also differ. Mostly wet grinding is not in full time continuous operation, while the digestion strings are. Enough reason to divide the total M_{V1} -control into two steps.
- 5.) The production load is set by the feed-rate control WSRC-101. The bauxite feed-rate measurement is used to govern the first ratio controller FIC-105. The set point for controlling the liquor flow rate V_{L1} is derived from the computation marked COMP-104:

$$\dot{V}_{L1S} = \dot{m}_{BX} \left[\frac{X_{BX}}{C_{NL} - \frac{M_{VO}}{1,645} \cdot C_{AL}} \right] \cdot \frac{M_{VO}}{1,645}$$

where \dot{V}_{L1S} is the set point for the liquor flow rate
 \dot{m}_{BX} is the weight rate of the bauxite feed
 X_{BX} is an average of extractable alumina content
in the bauxite
 C_{NL} is the caustic soda concentration of the
liquor
 C_{AL} is the alumina concentration of the liquor
 M_{VO} is the set point for the first step in molar
ratio control (which is always less the final
molar ratio M_{V1})

The variables C_{NL} and C_{AL} are measurements of the
analyser unit QR-103, while \dot{m}_{BX} comes from the belt
scale of the control loop WSRC-101.

6.) The final stage is the second ratio controller FIC-109,
its set point being obtained from the unit designated
COMP-108:

$$\dot{V}_{L2S} = \dot{V}_P \frac{(1-b_1)C_{A1}}{C_{NL} - \frac{M_{VO}}{1,645} \cdot C_{AL}} \cdot \frac{\Delta M_{V1}}{1,645}$$

where \dot{V}_{L2S} is the set point for the corrective liquor flow
controller FIC-109
 \dot{V}_P is the slurry feed rate to the digestion,
metered by FRS-106
 C_{NL} is the caustic soda concentration of the
liquor
 C_{AL} is the alumina concentration of the liquor
(both measurements coming from the analyser
QR-103)

- C_{A1} is the alumina concentration at the digester discharge
- b_1 is the volumetric solids concentration at the digester discharge
- M_{V0} is the imaginary molar ratio without corrective liquor addition
- ΔM_{V1} originally is $M_{V1} - M_{V0}$, i.e. the difference between the molar ratio with and without additional V_{L2} feed. ΔM_{V1} remains unchanged as long as the required value expressed by the set-point M_{V1S} is equal to the actual sampled-data measurement M_{V1} gained from periodic checks indicated as HQ in the flow diagram. When, however, it is found that $M_{V1S} \neq M_{V1}$, the earlier ΔM_{V1} must be altered by the difference $M_{V1S} - M_{V1}$ of the last cyclic check.

Theoretically the sampled-data control made is best suited for applications with delayed and slow response which is characteristic for the digester string. The time interval between periodic checks of M_{V1} followed by modification of flow ratio may be about 8 hours. The advantage is that conventional analysis can be made.

Normally the ratio of V_{L2} to V_p may be around 0,2. Whenever V_{L2} tends to become too low, M_{V0} may be changed to reduce the flow rate of V_{L1} adequately. In the course of events the feed-back would trim the flow rate of V_{L2} back to normal.

The concept of molar ratio control of the digester blow-off could be realized in a very sophisticated way, by using microcomputers and automatic analysers, which would be able to hold the molar ratio within a narrow

margin over the saturation level. The same concept may also be followed by applying two simple ratio controllers and modify the set-point of the second loop manually according the periodic analyses carried out in the conventional way. M_{V1} would float within a somewhat broader range but could be acceptable for running the process.

4.2 Concentration control of aluminate liquor

(Process control diagram in Fig.17)

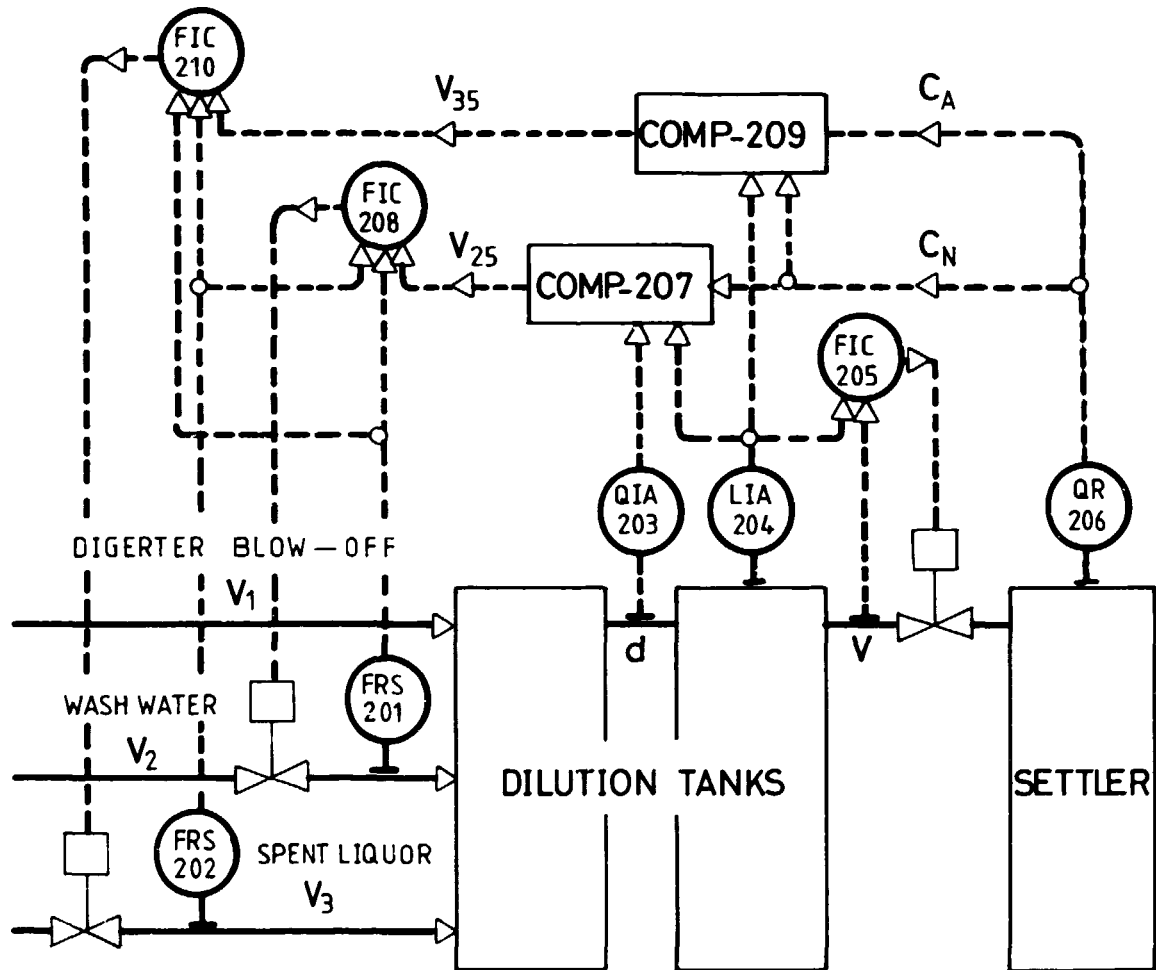


Fig.17 Concentration control of aluminate liquor

Dilution of the digester discharge as shown in Fig.17 means mixing three process streams V_1 , V_2 and V_3 , as practiced in many alumina plants.

The resulting caustic soda and alumina concentrations play a key role in the process cycle along with the concentrations of the digestion liquor. Locally they can curb alumina hydrolysing losses in the settling step, and the performance of precipitation is also strongly influenced by them.

The digester blow-off V_1 as marked in Fig.17 is the uncontrolled or wild variable, which may freely change depending on the different process loads the digester plant is run. Diluting water (V_2) and spent liquor (V_3) should be added in a controlled manner to hold the concentrations C_N and C_A of the discharged alumina liquor V at preset values.

Referring to Fig.17 the following characteristics may be mentioned:

- 1.) The measurements QR-206 of the alumina liquor concentrations are made in the overflow of the settler by either an automatic analyser or by conventional plant routine.
- 2.) The adapted concept shown is a feed-forward control with feed-back trim to meet high accuracy requirements. From measurements of variables at the feed end a forward information is derived, namely how they would effect the process output. Balances of mass rate, alumina and caustic soda content are ap-

plied to derive usable feed-forward expressions for the required set points of V_{2S} and V_{3S} . The total discharged flow rate V and its density d are employed, because V_1 cannot be measured directly, the digester blow-off stream being a mixture of slurry and flashed vapour.

The required flow rates of V_2 and V_3 to hold C_N and C_A stable can be expressed as under a, and b, :

$$a) \quad V_{2S} = (K_{1N} \cdot d - K_{2N} \cdot C_N - k_{4N}) \cdot V - K_{5N} \cdot V_3;$$

The first part of the expression is considered to be calculated in the unit marked COMP-207, which receives signals from measurement QIA-203 for d (density), FIC-205 for V (total flow rate), and QR-206 for C_N (caustic soda concentration).

The second part is a value proportional to V_3 , which is taken from the measurement FRS-202 and is fed to the flow controller FIC-208 as a so-called decoupling loop.

$$b) \quad V_{3S} = (K_{1A} \cdot C_N - K_{2A} \cdot C_A) \cdot V - K_{3A} \cdot V_2$$

Again the calculation of the first part is thought to be performed in the unit marked COMP-209, which accepts signals from FIC-205 for V , QR-206 for C_N and C_A .

The second part is a decoupling line between measurement FRS-201 (V_2) and the controller FIC-210 (V_3).

The feed-back trim is actuated in both control loops by the caustic soda (C_N) and alumina (C_A) concentrations, both showing a delayed and slow response to disturbances due to the smoothing effect of the large settler volume. A periodic check of the concentra-

tions (for instance at every 4 hours) would be acceptable.

Density (d) and the flow rates V_2 , V_3 and V may change quickly, so their continuous measurement is needed.

- 3.) Wash-water flow rate (V_2) has been assigned as manipulated variable to control the caustic soda concentration (C_N). Similarly the spent liquor flow rate (V_3) to compensate for changes of the alumina concentration (C_A). The two control loops working simultaneously show a strong interaction, that means a change of any one manipulated variable would not only compensate its assigned controlled variable but influence also the other one. To cancel out this disturbing cross action the decoupling loops as explained above have been provided.
- 4.) The flow rate control FIC-205 receives its set point from the level measurement LIA-204, that means the flow-rate V will be strictly proportional to the level of the second tank of the dilution section. The changes in the flow-rate will therefor be smoothed, which is beneficial for both the concentration control described and for the settler operation.

The constant appearing in the equations a, and b, under point 2 can be expressed as functions of various process variables which are not monitored continuously but simply taken as constant.

Just for additional information the expressions are shown below:

$$K_{1N} = \frac{(1-b_1) \cdot C_{N1}}{(1-b_1) \cdot C_{N1} \cdot d_2 - C_{N2} \cdot d_1}$$

$$K_{2N} = \frac{(1-b_1) d_1}{(1-b_1) \cdot C_{N1} \cdot d_2 - C_{N2} \cdot d_1}$$

$$K_{4N} = \frac{(1-b_1)(C_{NO} \cdot d_1 + C_{N1} d_o) - (1-b) \cdot C_{NO} \cdot d_1}{(1-b_1) \cdot C_{N1} \cdot d_2 - C_{N2} \cdot d_1}$$

$$K_{5N} = \frac{(1-b_1) C_{N1} \cdot d_3 - C_{N3} \cdot d_1}{(1-b_1) \cdot C_{N1} \cdot d_2 - C_{N2} \cdot d_1}$$

$$K_{1A} = \frac{(1-b) \cdot C_{A1}}{C_{A3} \cdot C_{N1} - C_{A1} \cdot C_{N3}}$$

$$K_{2A} = \frac{(1-b) C_{N1}}{C_{A3} \cdot C_{N1} - C_{A1} \cdot C_{N3}}$$

$$K_{3A} = \frac{C_{A2} \cdot C_{N1} - C_{A1} \cdot C_{N2}}{C_{A3} \cdot C_{N1} - C_{A1} \cdot C_{N3}}$$

where b stands for volumetric solids content, C_N and C_A for caustic soda and alumina concentrations, d for density.

The numbers in subscript refer to the process fluids streams V_1 , V_2 and V_3 . The additional suffix 0

denotes the value at the zero point (if any) of the measuring range, which has a practical importance with density meters.

A lot of calculation would be required to determine the k-constants for various technological modifications. Fortunately the constants may be set for once and all, because the feed-back trim would compensate any deviation from the required values.

The computing units may be realized by micro-processors as well as by analog computer modules. Both should be able to hold C_N and C_A within narrow limits.

But a simplified version of this concept has been proved to be very useful in practical plant service: The density measured with QIA-203 is directly controlled by changing V_2 . The flow rate of the wash water is measured as with FRS-201, and its output is applied to govern a ratio controller to adjust the flow rate of the spent liquor proportionately as if by FIC-210. By checking the concentrations of alumina liquor at intervals, the set point of the density controller and the ratio for spent liquor control can be readjusted manually. With some operating experience C_N and C_A should remain within acceptable margins.

4.3 Seed ratio control for precipitation

(Process control diagram in Fig.18)

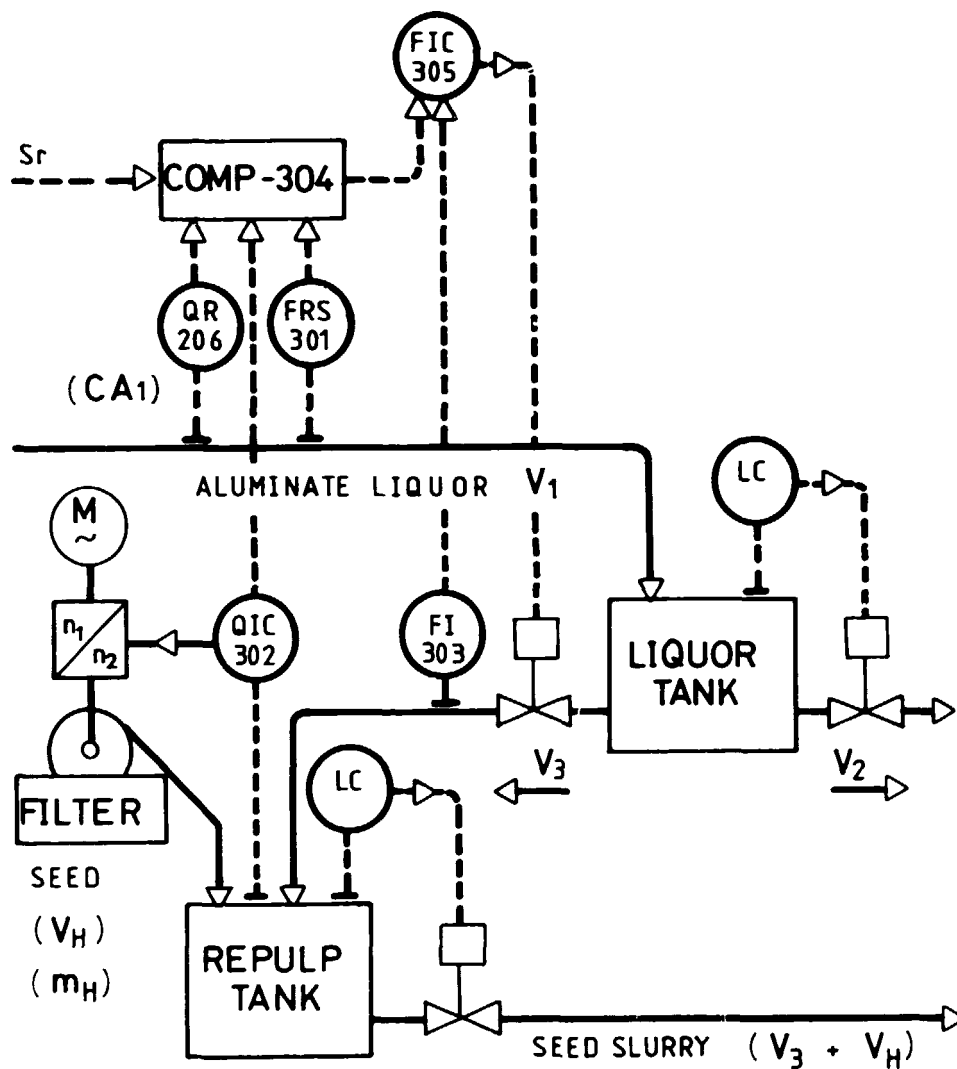


Fig.18 Seed ratio control for precipitation

Main factors influencing the efficiency of the precipitation plant unit are the caustic soda and alumina concentrations of the aluminate (or pregnant) liquor, temperature conditions in the course of the precipitation process and last but not least the seed ratio, S_r .

Seed ratio is by definition the mass ratio of alumina in the seeding alumina hydrate to alumina in the pregnant liquor feed:

$$S_r = \frac{\dot{m}_H \cdot 0,645}{V_1 \cdot C_{A1}} = \frac{\dot{V}_H \cdot d_H \cdot 0,645}{V_1 \cdot C_{A1}}$$

where \dot{m}_H , V_H and d_H are the mass rate, the volume rate and the density of alumina hydrate, V_1 is the flow rate of pregnant liquor feed to the precipitation plant and C_{A1} its alumina concentration.

The total pregnant liquor feed V_1 is divided into two streams as shown in the flow diagram. One part (V_2) is directly fed into the precipitator, the other one (V_3) is used to repulp seed hydrate (\dot{m}_H), which arrives as filter cake in the mixer tank.

The final aim of the control system is to hold the seed ratio as defined above at a preset value.

The alumina hydrate input (\dot{m}_H) mostly cannot be measured directly because the plant layout makes it difficult to apply the scales required. By using a special density measurement this difficulty may be overcome. The density meter calibrated to a measuring range of $(d_H - d_L)$,

i.e. the difference between the densities of alumina hydrate and pregnant liquor, would be quite helpful. With a density d in the repulping tank the output signal of such a meter would be

$$X_d = \frac{d-d_L}{d_H-d_L}$$

Considering the mixing flow rates in the repulp tank the equation

$$\frac{V_H}{V_H+V_3} = \frac{d-d_L}{d_H-d_L} = X_d \text{ can be derived, which}$$

renders the useful expression

$$V_H = V_3 \cdot \frac{X_d}{1+X_d}, \text{ showing that the volume rate}$$

of the seed hydrate (V_H) can be determined from readings of the repulping liquor stream (V_3) by FI-303 and the slurry density (X_d) by QIC-302 measurements.

The flow rate of the total pregnant liquor feed V is monitored with the meter FRS-301, its alumina concentration with the analyser QR-206.

The expression for seed ratio may now be rewritten as

$$S_r = \frac{V_3 \cdot \frac{X_d}{1+X_d} \cdot d_H \cdot 0,645}{V_1 \cdot C_{A1}}, \text{ from which the required}$$

repulping liquor stream (V_{3S}) to arrive at any preset seed ratio S_{rs} can be derived as

$$V_{3S} = S_{rs} \cdot \frac{1+X_d}{X_d} \cdot \frac{V_1}{d_H \cdot 0,645} \cdot C_{A1}$$

(The constant 0,645 is the ratio of the molar weights Al_2O_3 to $Al_2O_3 \cdot 3H_2O$. The density of alumina hydrate can also be taken as a constant $d_H = 3,0$ for practical purposes.)

The unit marked COMP-304 may calculate the set point for the flow rate controller FIC-305 according to above equation. Measurements of V_1 , C_{A1} , X_d and the manual set point input S_{rs} are fed into the computer modul to render V_{3S} .

At different process loads indicated by changing pregnant liquor feed rates (V_1) the seed hydrate input (V_H) must also be adjusted. The density control loop QIC-302 automatically takes care of that. For changing loads the filter drive is manipulated to modify filter cake output. A high- and low alarm announcement would remind the filter operator to shut down or start up another drum filter unit so that the filter speed adjustment could hold the repulp density within acceptable limits.

An advanced control concept has been derived from fundamental considerations and ways to practical realization shown. The immediate result is a highly sophisticated scheme as appearing in the process flow diagram Fig.18.

It should be noted, however, that the whole system is based upon two straight-forward control loops. One

is a simple flow-ratio controller making V_3 proportional to V which is taking care of load changes. The other one is a density controller of the repulping process. The later has its main difficulty in the execution of the controlled adjustment of seed hydrate filtration output. Once this has been solved to satisfaction the required seed ratio can be maintained at a permanent level.

4.4 Control of hydrate wash filtration

(Process control diagram in Fig.19)

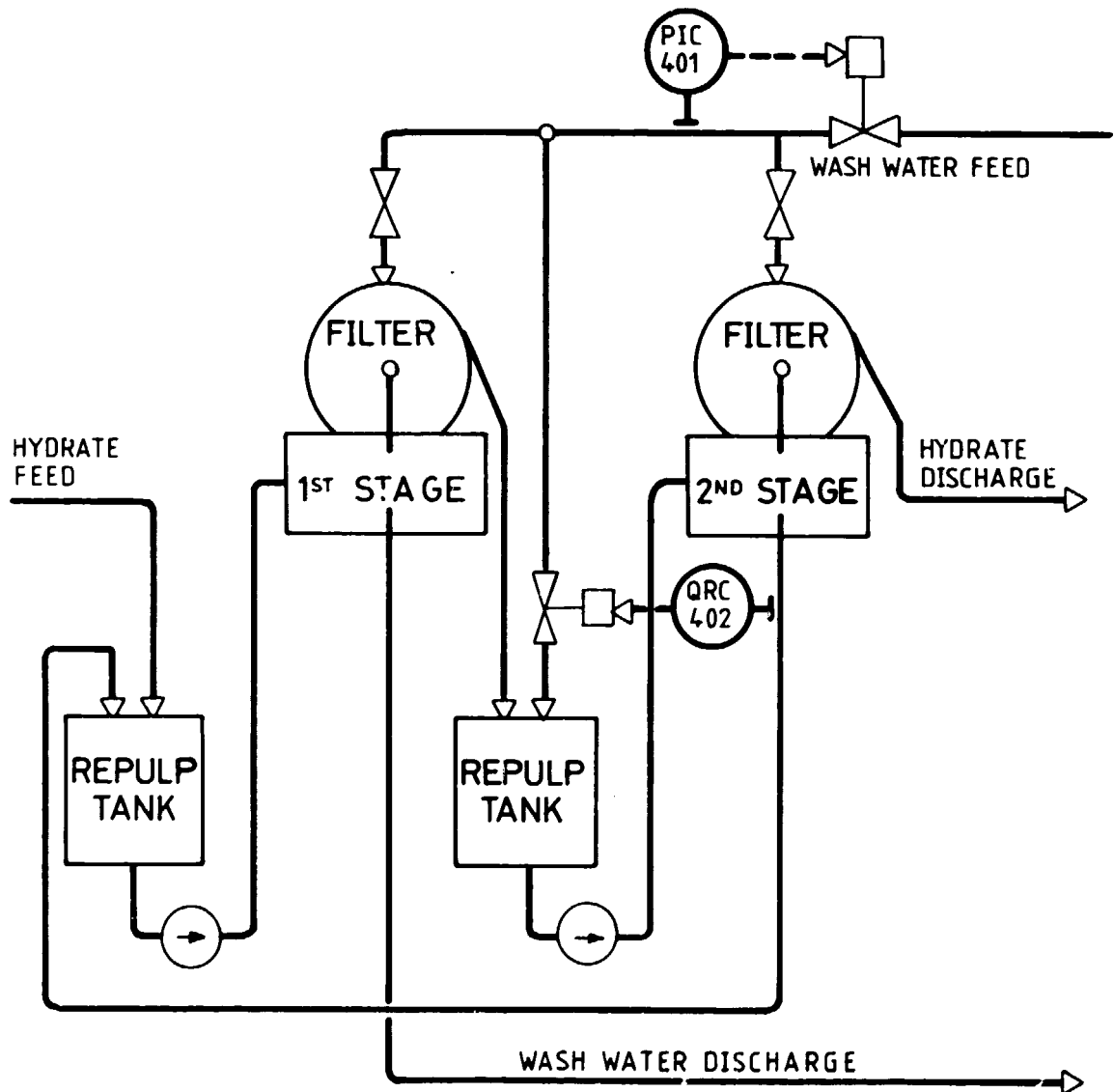


Fig.19 Control of hydrate wash filtration

The alumina hydrate wash filtration process has to fulfill two somewhat contrary requirements: high purity of product hydrate and reasonably low wash water consumption.

The economy is linked to the selling price depending on product purity and to steam costs of evaporation increasing with higher wash water consumption.

The proposed control scheme aimed at optimal economy has been conceived from detailed investigation of the process, the main characteristics of which are dealt with as follows:

- 1.) Given a two stage filtration as shown in the flow diagram, it is adviseable to have a steady amount of approximately 0,32 cubicmeter wash water per ton alumina plant output, divided between the two filter stages into the optimal portions of 0,08 m³/t to the first stage and 0,24 m³/t into the second one.

A simple solution is to apply sprinklers of adequate numbers and nozzle areas to get 25 percent of the total cross section into the first stage and 75 percent into the second. By engaging the pressure controller (PIC-401) for the wash water supply, a steady flow rate of sprinkling water for each stage would be maintained in optimal portions. Filter load changes must not be taken into consideration because of the flat optimum.

- 2.) Product purity can be related to the caustic soda concentration of the adhesive moisture. Plant investigations have revealed a relation between alumina

hydrate impurity and the caustic soda concentration of the second stage filtrate liquid, so that the equilibrium constant could be determined. Consequently the plain control loop QRC-402 could be applied to keep product purity in check.

- 3.) Within the concentration range in question there is a linear relation between electrical conductivity and caustic soda concentration. By controlling this related variable, product purity may be kept within acceptable limits using wash water rates to best economy.

4.5 Concentration control of strong liquor

(Process control diagram in Fig.20)

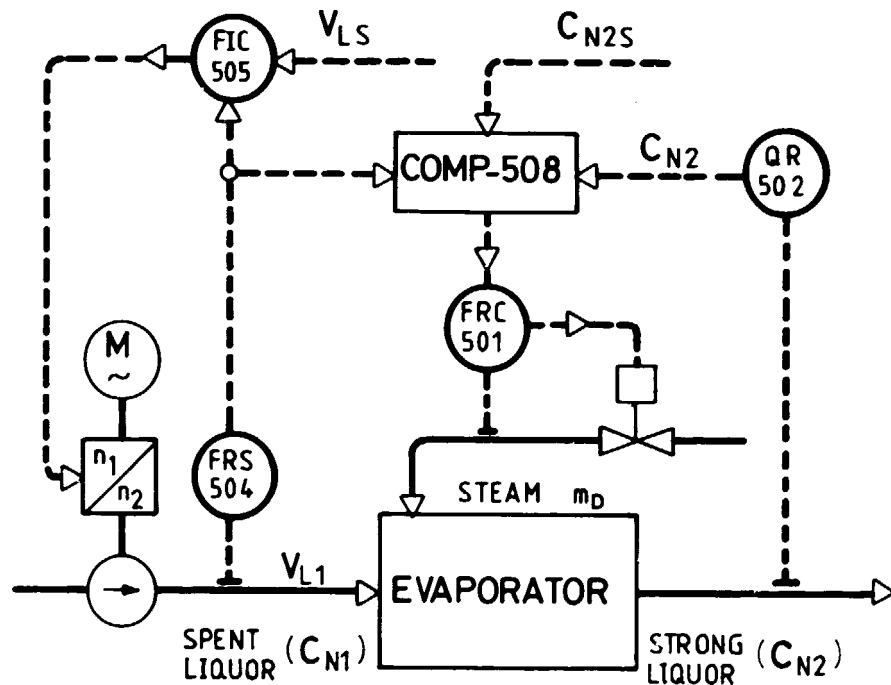


Fig.20 Concentration control of strong liquor

In the course of the process cycle water is added to and taken out from the circulating liquor at different processing stages. The water balance is finally restored by the evaporator which extracts exactly as much water as is needed to hold the strong liquor concentration at a constant level. Aim of the control scheme to be introduced is the automatic stabilization of the concentration in the evaporator discharge. A steady concentration of the strong liquor is of advantage for salt removal and for digestion liquor preparation.

The fundamental considerations which lead to the control scheme adopted should first be dealt with briefly. Any increase in liquor concentration is the result of some water extraction. A balance of caustic soda expressed for a water extracting process yields the basic equation:

$$C_{N2} = C_{N1} \frac{V_{L1}}{V_{L1} - W}$$

where C_{N1} and C_{N2} are the caustic soda concentrations at the feed- and the discharge end, V_{L1} in the liquor feed rate and W the amount of water extracted.

For a conventional multistage evaporator W might be expressed as a function of the heating steam flow rate m_D :

$$W = k \cdot n \cdot \frac{r_D}{r_B} \cdot m_D$$

where r_D and r_B are the latent heats of the heating steam and the vapour extracted, n is the number of evaporator stages and k is a factor less than unity depending upon the process heat losses.

From above equations a relation between liquor feed rate V_{L1} and heating steam rate m_D can be derived:

$$m_D = V_{L1} \cdot \left(1 - \frac{C_{N1}}{C_{N2}}\right) \cdot \frac{r_B}{k \cdot n \cdot r_D}$$

For any given evaporator, either of the counter-current or cocurrent type, the ratio $r_B/k \cdot n \cdot r_D$ can be taken as constant. The feed concentration C_{N1} would possibly change within a certain range, but will not be monitored, and regarded stable for purposes of control strategy.

Referring to the flow diagram the advisable control system for the evaporation is shown as below:

- 1.) The flow rate of the liquor feed can be set by the control loop FIC-505, to meet the need of the process cycle, indicated by the relation of stored volumes of spent liquor and strong liquor. The signal of the evaporation feed rate (V_{L1}) is feed to the computer module COMP-508, which would determine the steam rate required for this load.
- 2.) Essentially the same relation as given above is calculated by COMP-508, with a slight modification to accommodate the feed-back correction (ΔC_{N2}):

$$m_D = K_E \cdot V_{L1} \cdot \left(1 - \frac{C_{N1}}{C_{N2S} - \Delta C_{N2}}\right)$$

where $K_E = \frac{r_B}{k \cdot n \cdot r_D}$ and

$$\Delta C_{N2} = (C_{N2} - C_{N2S})$$

When $C_{N2S} = C_{N2}$ and $\Delta C_{N2} = 0$, the ratio between V_{L1} and m_D remains unchanged.

With $C_{N2S} > C_{N2}$ and $\Delta C_{N2} > 0$ (in case of too low concentration), the ratio would increase to call for a higher steam rate.

If, however, $C_{N2S} < C_{N2}$ and $\Delta C_{N2} > 0$, that means the concentration is higher than required, the ratio would decrease to require less steam for the same liquor feed rate.

- 3.) The feed-back signal C_{N2} is furnished from QR-502. The measurement of the liquor boiling point elevation is a practicable substitute for a direct analyses of the caustic soda concentration. Within the required measuring range there is a linear relationship between the elevation of the boiling point and the caustic soda concentration of the liquor.

As a simplified version of evaporation control a conventional ratio controller with the proper ratio set out from practical plant experience could also do the job acceptably.

4.6 Concentration control of digestion liquor

(Process control diagram in Fig.21)

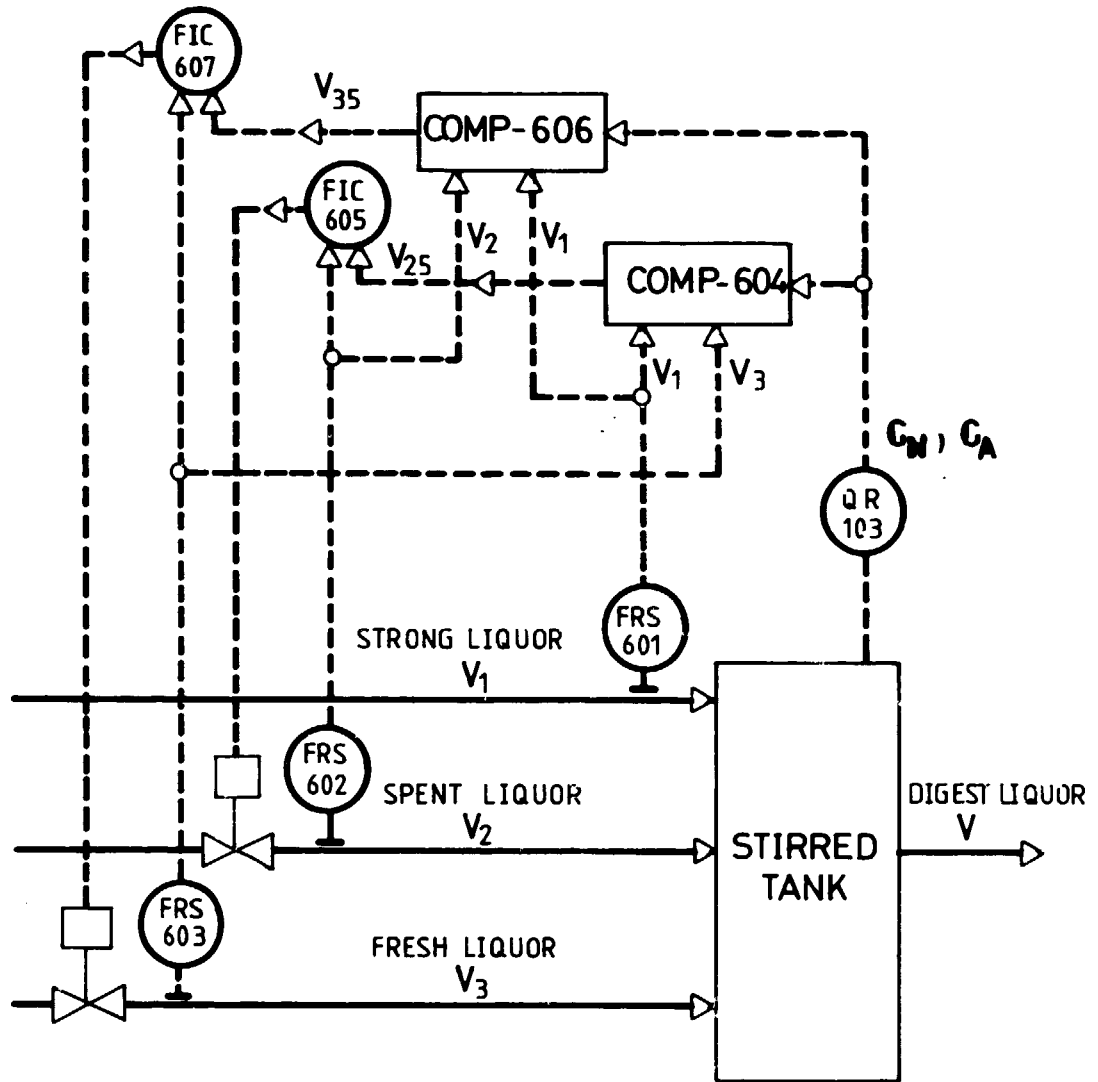


Fig.21 Concentration control of digestion liquor

Digestion liquor is a mixture of three liquor streams. Strong liquor V_1 flows into the mixed tank freely as the uncontrolled "wild" variable. Spent liquor V_2 and make-up liquor V_3 are to be added proportionately to arrive at a uniform and desired composition of the digestion liquor.

The caustic soda - and alumina concentrations of the outgoing digestion liquor have a decisive local influence on the specific steam consumption and the production output of the digestion plant.

Further on the economy of the whole Bayer cycle will also depend upon the relation between the liquor concentrations in the digestion and precipitation plant units.

The control concept is similar to that shown for aluminate liquor concentration control. The controlled variables (C_N and C_A) and the manipulated variables (V_2 and V_3) must first be matched properly one to another. The primary function of V_2 is considered to control C_{N1} while V_3 is to influence C_{A1} (or rather the ratio C_N to C_A) first of all.

The set point V_{2S} for FIC-605 may be calculated by COMP-604 according to the equation as under, to hold required caustic concentration C_{NS} :

$$V_{2S} = \frac{C_{N10} - C_{NS}}{C_{NS} - C_{N20}} \cdot V_1 + \frac{C_{N30} - C_{NS}}{C_{N30} - C_{N20}} \cdot V_3$$

V_3 is to be controlled by FIC-607, for which COMP-606 furnishes the set point to shift the alumina concentration towards C_{AS} :

$$V_{3S} = \frac{C_{A10} - C_{AS}}{C_{AS} - C_{A30}} \cdot V_1 + \frac{C_{A20} - C_{AS}}{C_{AS} - C_{A30}} \cdot V_2$$

when (as it might be the case) $C_{A30} = 0$, above equation becomes

$$V_{3S} = \left(\frac{C_{A10}}{C_{AS}} - 1\right)V_1 + \left(\frac{C_{A20}}{C_{AS}} - 1\right)V_2$$

where C_{NiO} and C_{AiO} are the average concentrations for the related process streams V_i . C_{NS} and C_{AS} are the set points for both caustic soda and alumina concentrations.

If feed-back trim is applied, as indicated in the process control diagram, $C_{NS} - (C_{NS} - C_N) = C_{NS} - \Delta C_N$ is to be substituted for the simple C_{NS} used in the feed-forward equation. The result is a shifting C_{NS}' , which while on the increase would reduce the flow of V_2 and vice versa.

The same applies for C_A -control. The feed-back expression $C_{AS} - (C_{AS} - C_A)$ indicates, that the actual value C_A becoming higher than the set point C_{AS} , would finally increase the flow of V_3 .

The simplified version of above concept would be to apply two ratio controllers. The strong liquor flow rate V_1 would be the wild variable for both loops. The flow rates of spent liquor V_2 and fresh liquor V_3 would be controlled at manually set ratios proportionally to strong liquor flow rate V_1 .

