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UNITED NATIONS INDUSTRIAL DEVELOPMENT ORGANIZATION
Vienna International Centre, P.O. Box 300, 1400 Vienna, Austria
Tel: (+43-1) 26026-0 · www.unido.org · unido@unido.org

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Techno-Economic Study for Upgrading the Technology of the Korba Alumina Plant in India

Contract No. 87/144

Final Report

Aluterv-FKI
Budapest/Hungary
January 1989

CONTENTS

0. Executive summary; conclusions and recommendations

In pursuance of UNIDO Contract No.87/144 dated 3/29 Feb.'88 four Hungarian experts spent about two weeks each in the Korba Alumina Plant of Bharat Aluminium Company, India to study the state and capabilities of the said Alumina Plant, the reasons for not achiving its rated capacity upto the present time and the possibilities for increasing the plant's output and improving its technical parameters. They have found that in spite of the serious efforts made by the plant's personnel to improve the operation and the addition of some equipment since the last visit of a Hungarian expert team some problems still make impossible to attain the rated capacity of the plant. These reasons are the following:

a) the repeated failures of the steam and power supply systems,

b) the high levels of some contaminants in the plant liquor (partly caused by the othervise very successful test of adding some lime to the digestion slurry),

c) the high amounts of non-technological ("unauthorized") water entering the process circuit,

d) the inadequate precipitation efficiency caused partly by the high level of contaminants mentioned above, partly by the long-term after-effects of some power interruptions (the problems connected with the difficulty of restarting some precipitartors after such failures),

e) the poor performance of the old seed filters and also of the newly installed disc filter and

f) the poor state and obsoleteness of the plant instrumentation.

By the time the third centrifuge and the second disc filter will be commissioned (about the end of 1988) the basic equipment required for attaining the rated capacity will be essentially in place, except for the grinding capacity, which - until the implementation of the secondary crushers and the closed circuit grinding - will cuse the loss of about 5,000 tonnes of production per year during the maintenance period of one of the ball mills.

However, in order to fully utilize the capacity of the equipment, some further conditions will have to be fulfilled:

a) the frequent steam and power failures will have to be eliminated or their effect be minimized by improving the operation of the steam boilers and by installing the Diesel compressors, already decided and the high pressure one suggested by the Hungarian team during their visit to protect the precipitators from the effects of the power failures,

b) the level of the contaminants will have to be drawn down to acceptable levels by making continuous efforts to utilize the capacity of the three centrifuges in an optimum way,

c) continuous efforts will have to be made to reduce the amount of non-technological water entering the process circuit at least by half,

d) efforts will have to be made to maintain a constant high precipitation efficiency by reducing the downtimes of the precipitators and by increasing the seeding ratio,

e) the operation of the hydrate filters, first of all that of the disc filters will have to be closely monitored and efforts will have to be made to improve their productivity and

f) extraordinary efforts will have to be made by both the operating and the laboratory personnel to compensate for the deficiencies of the instrumentation system.

Some of the suggestions incorporated in the present report serve also the better utilization of the plants's capacity. These are:

a) the use of up-to-date tube-in-tube type heat exchangers for replacing the present HP and LP digestion preheaters requiring frequent and time consuming maintenance,

b) the evaporating of most or all of the hydrate wash water for improving the precipitation of the contaminants,

c) some improvements suggested for the operation of the precipitator tanks,

d) the improvement of the precipitation by adding an extra precipitator to the line through relieving it from its present function of storing caustic lye,

e) the updating of the wourn-out aluminate liquor cooling system (PHEs), by replacing them with new ones,

f) a modification in the air supply of the disc filters,

g) some updating suggested for the interstage cooling system of the precipitators.

Other suggestions included in the present report serve the reduction of the steam consumption (by bypassing the flash steam heated digesters with tube-in-tube type preheaters for improving the utilization of the heat content of the digested slurry), the reduction of the bauxite consumption (by increasing the extraction efficiency of alumina through lime addition after making sure that the adverse effects of this otherwise very efficient method experienced during 1987 are eliminated), the reduction of the caustic soda consumption (by improving the mud washing and causticizing efficiency), the reduction of the fuel consumption of the calciners (by using an additive for reducing the adhesive moisture content of the product hydrate) and the improvement of the grain size distribution of the product hydrate (by using a recently developed hydroseparator for hydrate classifying).

A complete replacement and updating of the instrumentation system is also suggested in the present Report because they are very much run down and need excessive maintenance even to maintain their poor performance and because the whole system (pneumatic control) applied during the construction of the Korba Alumina Plant has become totally obssolete. The suggested electronic system is much more accurate, needs much less maintenance and can be integrated into a computer controlled system at a later date.

The capital costs of every suggested modification the returns expected of them and the pay-back times are shown in Chapter 6 at the end of the Report. The total capital costs of all of the suggested modifications amount to Rs according to our estimate; the expected total returns amount to Rs per year (at a 200,000 tpy production level) and the average pay-back time would be years. Over and above these returns it has also to be taken into consideration that the majority of the suggested modifications also serve the attaining of the rated capacity of the plant and would also give lots of indirect benefits.

1. Preliminaries

The designed capacity of the Korba Alumina Plant of Bharat Aluminium Co. in India is 200,000 tpy. The Korba Alumina Plant was established in the second half of the 1960 s based on the know-how and basic engineering design by Aluterv-FKI. As the plant has not attained the rated capacity and is producing about 170,000 tpy of alumina at present due to the shortage of electric power and modernization reguirements of the process and since then considerable R D work has been done in updating the technology throughout the world the plant requested UNIDO assistance in updating the production process. As the result of this request a Contract has been entered into between the UNITED NATIONS INDUSTRIAL DEVELOPMENT ORGANIZATION /"UNIDO"/ and Aluterv-FKI. The aim of this Contract is to assist Bharat Aluminium Co., India in selecting the most suitable technology which would enable them to significantly reduce energy consumption and increase quality and productivity in the production of alumina at the Korba Alumina Plant and to provide assistance to the Government in carrying out the project entitled "Preparation of a Techno-economic Study for Upgrading the Technology of the Korba Alumina Plant".

According to the UNIDO Contract a team of Hungarian experts from Aluterv-FKI visited Korba Alumina Plant from the 10th February to the 6th March, 1988 in order to collect necessary background information about the actual state of the plant and an Indian team visited Aluterv-FKI and the Hungarian alumina plants in September, 1988 to appraise themselves with the development work carried ont in the recent past in Hungary.

This Report contains the findings and recommendations of Aluterv-FKI for the improvement of the technological parameters based on the bauxite supplied from the Amarkantak and Phutka-Pahar resources, as we do not have sufficient data about further possible bauxite supplies.

2. Processing of the technical data supplied by BALCO

The examination of the technological state of the alumina plant was started by the analysis of the technological data recorded by the alumina plant. Aluterv-FKI has a mathematical model since 1964 for the purpose of calculating the material and heat balances of any Bayer alumina plant. Basically the model may be utilized in two ways:

(1) Designing a new alumina plant

According to the technological system chosen, the main technological data may be given, and the computer will determine the material balance for the various technological operations by means of the mathematical model. In such a way it is suitable for the fast calculation and also for the comparison of the material and heat balances of the different technological variants.

(2) Calculating the material and heat balances of operating alumina plants

At this time the material balance is based on the technological data recorded at the alumina plant. The accuracy of the material balance depends on the quantity of data recorded and (as operational data are concerned) on the reliability of the different variables measured, such as material flow-rates, analysis data, etc. Conclusions may be derived from the actual data regarding technological failures (non-technological water getting into the process circuit, extra heat losses at various points, the suitability of the treatment of alkaline condensate, etc).

The data of three financial years were collected by Aluterv-FKI's experts:

April, 1985	- March,	1986
April, 1986	- March,	1987
April, 1987	- January,	1988

The alumina plant utilized lime addition in the Digestion Unit for ten months. The different character of this period motivated, that data regarding this period was handled separately. Thus material and heat balances for four periods were determined by the help of our mathematical model:

April,	1985	-	March,	1986
April,	1986	-	January,	1987
February,	1987	-	November,	1987
December,	1987	-	January,	1988

During the processing of the data we have made the following observations:

- The alumina production was determined by utilizing the following data: the composition of bauxite, the Al_2O_3 yield based on the mud analysis, and the quantity of bauxite brought into the alumina plant. It was different - expecially from the year of 1987 on - from the data given in the tables for hydrate production.

- Significantly more water gets into the technological circuit at various technological operations, than required by the technology. This water is disadvantageous, because it either increases the quantity of water to be evaporated, or if the capacity of evaporation is limited, it decreases the quantity of water utilizable for red mud washing, thus causing excess caustic and heat losses.

- The process liquor concentrations may be calculated based on the well known chemical reactions and the technological water balance. The losses may be estimated according to the analysis data. Nevertheless, in operating alumina plants at some technological points usually somewhat lower values are measured for the process liquor concentrations, than it could be calculated from the previously mentioned balances. The dilution is due to the feeding of non-technological water, namely floor washing water utilized at different points, high pressure gland water and unsatisfactory flashing. The model considers non-technological water at the following points:

  - wet grinding, slurry adjustment, desilication,
  - digestion, flash tanks, dilution, settling,
  - red mud washing, red mud causticization, red mud filtration,
  - precipitation, hydrate classification, hydrate filtration,
  - evaporation, salt removal.

  The calculation for determining the quantity of non-technological water will be discussed at the appropriate place.

- The specific bauxite consumption is very high in all of the four periods, the digestion yield is low according to the mud analysis data, and the hydrolysis losses are high.

- The Na_2O content of the last washer mud is higher than that of the digester slurry's mud phase, which fact is in opposition to what we know about the linkage of caustic to the mud. During the digestion sodium aluminium silicates and sodium titanates are formed, then during red mud washing a part of the Na content bound to the Na-titanates is dissolved. That is why the Na_2O content of the last washer mud may only be lower than that of the digester mud. According to the data of the Korba Alumina Plant this tendency is reversed, which indicates a probably systematic error in the analysis.

- The carbonate level and organic content of the process liquors demonstrates a rising tendency.

- The material balance based on the data of the fourth period indicates in some cases extremely high or extremely low (sometimes even negative) values for heat losses. It may be due to several reasons:

    - the selected period is shorter than required to reach a steady state operation, necessary for the calculation of the material balance,

    - during the period the plant operation was characterized by technological disturbances,

    - the data recorded at different points were unreliable.

Becacuse of these, the detailed material balance for this period is not given, only the characteristic technological data are included in Table 2-4.

The following conclusions were drawn from the balances of the different plant units and from the analysis data given:

Wet grinding of bauxite

The model uses the solids content of the ground slurry as one of the basic data, however, there were no such data available, so the calculations were made with the 500 gpl design data. The greatest problem in the wet grinding operation is, that the ratio of particles bigger than 300 microns in diameter is around 14-16 %, and that of those bigger than 700 microns is 1.5-3 %. Introducing of the closed circuit (recirculating) wet grinding process is considered in the alumina plant to enable it to maintain a size distribution better than maintained by the existing method.

The bauxite to liquor ratio control is not working reliably, considerably contributing to the low digestion efficiency. The uniformity of the lime addition could not be seen from the data. Its control is necessary.

Slurry adjustment, desilication

The temperature of the desilication did not reach the desired temperature (above 100 °C) value at any of the periods. The temperature of the flashed slurry is high (125-138 °C) indicating that the heat transfer coefficients of the desilication heat-exchangers is low due to scaling. A significant amount of non-technological water gets into the process (approximately 0.5 m^3/t) in the wet grinding, slurry adjustment and desilication units.

Digestion

Some technological faults were revealed during the evaluation of the digestion parameters. First of all the low digestion yield is evident, on the other hand the poor caloric operating state of the digestion lines. The low yield is due to the boehmite content of the processed bauxite, the digestion of which is difficult. In order to reduce the production of alumina, it was tried to reduce the molar ratio of the digestion, but this led to the sharp decrease of the digestion yield and produced a secondary unfavourable technological effect, namely the undigested boehmite helps to decompose the oversaturated aluminate liquor (this is the so-called hydrolysis) in the settling and washing process of red mud, thus beside the decrease of the digestion efficiency the overall plant yield decreases even further.

The caloric faults of the digestion have different origins:

- Because of the faults of the power plant due to different reasons the end temperature of the digestion does not reach the designed 240 °C. Due to the steam fall-outs the final temperature of digestion is fluctuating from time to tome, resulting in the fact that a part of the slurry flow leaves undigested for the red mud separation.

- The final temperature of the recuperation section never reaches the designed 170 °C, it fluctuates between 162 and 167 °C. The heat exchangers (the so called HP heat exchangers) heated by flash steam (2 stages) are raising the slurry temperature by 40 °C, while the digesters (3 stages) are raising it further by only 30 °C. The pipes of the heat exchangers are frequently clogged because of the unfavourably ground slurry and they are demanding repeated maintenance. In the recuperating digesters the temperature increment produced by one stage is low, mainly due to insufficient agitation.

- The final temperature of the flashing slurry is high, it is between 125 and 138 °C. The reason is similar to that discussed earlier, and furthermore, the heat transfer coefficient of the heat exchangers before desilication (the so called L.D. heat exchangers) is relatively low as well. This is also the reason of the temperature lag mentioned at the desilication.

- Significant amounts of non-technological water get into the digestion. It was determined by the help of our model in the following way:

  The molar ratio and the concentration of the digesting liquor is known from the data supplied. The amount of dissolved Al_2O_3 is known, the chemical Na_2O loss may be calculated using the data of mud analysis. The amount of flashed water may be calculated from the digestion temperature and the final temperature of the flashing slurry. The amount of water brought into the process by the bauxite in the form of moisture is known as well. Based on all of these data the expected Na_2O and Al_2O_3 concentrations of the aluminate liquor may be calculated at the end of the flash line. If the actual concentrations are lower than the calculated ones, the differences may only be due to non-technological water getting into the system. This water may get into the system by the following ways:

- with the H.P. gland water used at the agitators of the digesters and at the pumps,

- with the water used to wash-up floors at different places (through the sump tanks).

- Apart from the above mentioned facts the concentration of the blow-off liquor will also decrease, if not only slurry but a mixture of slurry and vapour leaves the individual flash stages. This phenomenon is more or less similar at all kinds of digesting systems, and it may be featured by an efficiency-like parameter. Ultimately less vapour leaves the slurry, which is resulting in a diluting effect. This phenomenon may be reduced by level control in the flash tanks. Unsatisfactory heat insulation may also reduce the flashing efficiency.

- The quantity of non-technological water originating from the above mentioned sources is accounted by the model as a single item. At the elaboration of the detailed material balance it was shared between the desilication and the digestion in a 50-50 % ratio.

Dilution, settling and washing

Detailed data have been provided regarding the settling-washing unit. The solids content of the underflows was appropriate, except for the year of 1986. The hydrolysis loss was too high, fluctuating in the range of 59 to 100 kg/t.

Beyond the reasons detailed at the digestion the low temperature of the washing line (the temperature of the first washer overflow fluctuates between 72 and 76 °C) also helps decompose the aluminate liquor, especially in the first washer, where the concentration is most advantageous for this process. A considerable amount of non-technological water gets into the red mud washing line, probably in the form of wash-up water. The reason for the low temperature of the washing line is the high heat loss of the red mud filtration, and beyond that a considerable amount of alkaline condensate produced in the plant (in the digestion and evaporation sections) gets lost; furthermore, the return water (having a much lower temperature) is used for red mud washing instead of it.

Causticization

The causticization efficiency is low, it is fluctuating between 11 and 20 %. The main problem in this technological operation is the insufficiency of control, that is why the proper handling cannot be assured. A decisive role is assigned to the temperature, the residence time and to the proper lime addition.

Control filtration, aluminate liquor cooling

No problems were observed in the operation of the control filtration, as the iron content of the product is in accordance with the requirements.

We had no data available regarding aluminate liquor cooling, the temperatures concerned are not recorded systematically in the alumina plant.

Precipitation

The precipitation efficiency was low (appr. 49 %) in the period examined. The impurities of the process liquor, the low value of the seed ratio (between 0.8 and 1.4) and the insufficient precipitation time (70-80 hours) are considered to be the reasons of it. After the frequent compressed air fall-outs, the plant is not able to restart the air agitation in the precipitator tanks, where the hydrate particles settle and compact, that is the reason why out of the 14 precipitator tanks only 11-12 pieces are in operation at any given time. According to the data a considerable amount of non-technological water gets into the process liquor during this operation.

In theory the liquor should be concentrated during the precipitation due to the water content of the crystals precipitated. On the basis of this the expected increase of the liquor concentration may be calculated. The lack of concentration increase or even the dilution of the process liquor from time to time is the consequence of non-technological water getting into the precipitators or related equipment.

According to the information gathered in Korba there was no observable difference in the particle size distribution of hydrate between the overflow and underflow of the hydroseparator, that is why the reduction of the grain size resulting from the liquor impurities cannot be counterbalanced by the classification.

Hydrate filtration

There were no data available on the moisture content of the seed hydrate and the pre-filtered product hydrate, since these data are usually not recorded.

Aluterv-FKI's experts experienced during their mission that the product hydrate is not removed from the surface of the disc filter by the compressed air. Due to this the productivity of the seed filters was reduced, they are not able to filter the sufficient amount of seed hydrate.

This may also be the reason of the rather low seed ratio, providing one reason for the low precipitation efficiency. The moisture content of the product hydrate fluctuates between 9 and 11 %, it is rather high.

We could only estimate the quantity of hydrate wash water, because, besides the logged Na_2O concentration, the moisture content of the pre-filtered product hydrate should also have been known for an accurate calculation. In our estimation it is in the range of 0.8-1.4 m^3/t. With the washing technology applied, a specific value of 0.8 m^3/t is acceptable for the wash water. Most of the non-technological water may get into the process with the wash water used to wash the filters.

The floating hydrate content of the spent liquor was between 1.5 and 2.9 gpl, calculated as Al_2O_3. This value is calculated on the basis of the molar ratio of the digesting liquor. The molar ratio of the digesting liquor may be calculated definitely

- from the data of the decomposed, filtered spent liquor, part of which goes for evaporation,
- from the data of the make-up caustic added for compensating the Na_2O losses,
- from the estimated value of the scalings in the precipitators removed either by hot spent liquor or by fresh caustic,
- from the estimated values of caustic and alumina included in the moisture content of the removed salts and from some mechanical losses.

Should the actual molar ratio of the digestion liquor be lower than that of its calculated value, this indicates either that some hydrate has been left in the spent liquor at the filtration, or some aluminate liquor has got into the spent liquor at the heat exchangers.

Evaporation, salt removal

According to the recorded data, the evaporation capacity was not utilized more than about 77-88 % even at this low process efficiency level, so a significant reserve capacity is available for increasing the production. The specific steam consumption of the evaporation equipment is between 0.27-0.30 t/t of evaporated water, similar to other comparable types of equipment. At this technological point, a lot of non-technological water goes also into the process partly from the washing operations and partly during the salt separation by washing the equipment of the salt filtering and separation (0.6-1.1 m^3/t).

The salt separating system was suitable for the separation of both V_2O_5 and soda salts in its original form. Nowadays the V_2O_5 salt is separated by liquor cooling in the evaporation plant unit. This method is well proven for years, and sufficient amounts of sodium carbonate and organics have left the process liquor together with the separated salt to keep the impurities at a reasonable level. (The sodium carbonate content was under 10 % expressed as Na_2O_c/Na_2O_k, the organics level was 16-18 gpl calculated as oxalate, based on a 140 gpl concentration process liquor.)

In January - February of 1987 lime addition was introduced at the digestion, as an experiment. Following this, the sodium carbonate and organics level of the process liquor increased. According to our assumption the reason was the following: during the digestion the added lime formed insoluble calcium-phosphate with the P_2O_5 dissolved from the bauxite, and by this way after a certain time the P_2O_5 level of the process liquor was reduced, and its amount was insufficient to form the common vanadium-phosphorus-fluorine ternary salt, the so-called vanadium (or V_2O_5-) salt. With the vanishing of a separable vanadium salt, the possibility of the separation of other impurities came to an end, resulting in the rapid increase of the level of impurities.

Digesting liquor preparation

The method of digesting liquor preparation was discussed in the section concerning hydrate filtration. During the data processing we found the concentration and molar ratio lower than the acceptable value, which resulted from the unused evaporation capacity, the high quantity of non-technological water in the evaporation, and the high floating hydrate content of the filtered liquor.

The material and heat balances for the different periods are given in Tables 2-1; 2-2; 2-3 and the characteristic technological data are summarized in Table 2-4.

Table 2-1/1

TECHNOLOGICAL MATERIAL AND HEAT BALANCE
April 1985. - March 1986.
Alumina production 20 tph

MATERIAL STREAM	Na2O	Al2O3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	Al2O3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
BAUXITE GRINDING														
In: Bauxite: solid phase	-	1309.72	2.710	54.2	0.903	18.1	3.000	-	-	-	27	0.837	1.2	-
moisture	-	-	0.123	2.5	0.123	2.5	1.000	-	-	-	27	4.187	0.3	-
Digestion liquor	752.87	369.93	5.598	112.0	4.394	87.8	1.274	171.3	84.2	3.35	70	3.639	28.6	-
Out: Mill slurry	752.87	1679.65	8.431	168.6	5.420	108.4	1.552	166.7	81.9	3.35	56	2.746	25.7	500
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	4.4	
SLURRY ADJUSTMENT														
In: Mill slurry	752.87	1679.65	8.431	168.6	5.420	108.4	1.552	166.7	81.9	3.35	56	2.746	25.7	500
Digestion liquor	1269.91	624.04	9.438	188.8	7.412	148.2	1.274	171.3	84.2	3.35	70	3.639	48.1	-
Out: Adjusted slurry	2022.78	2303.69	17.869	357.4	12.832	256.6	1.393	169.6	83.3	3.35	63	3.218	71.9	211
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	1.9	-
SLURRY PREHEATING AND PREDESILICATION														
In: Adjusted slurry	2022.78	2303.69	17.869	357.4	12.832	256.6	1.393	169.6	83.3	3.35	63	3.218	71.9	211
Flash steam	-	-	0.850	17.0	-	-	-	-	-	-	(2378)		46.5	-
Alkaline condensate	-	-	1.675	33.5	1.675	33.5	1.000	-	-	-	145	4.210	20.5	-
Non-technological water	-	-	0.305	6.1	0.305	6.1	1.000	-	-	-	35	4.187	5.4	-
Out: Predesilicated slurry:														
liquid phase	1982.47	1910.77	16.831	336.6	12.785	255.7	1.316	155.1	149.5	1.71	84	3.552	100.4	101
solid phase	35.31	392.92	1.343	26.9	0.448	9.0	3.000	-	-	-	84	0.837	1.9	-
Alkaline condensate	-	-	2.527	50.5	2.527	50.5	1.000	-	-	-	128	4.200	27.2	-
Heat loss + heat of reaction	-	-	-	-	-	-	-	-	-	-	-	-	14.8	-
DIGESTION														
In: Predesilicated slurry	2017.78	2303.69	18.174	363.5	13.233	264.7	1.373	155.1	149.5	1.71	84	3.351	102.3	101
High-pressure steam	-	-	2.650	53.0	-	-	-	-	-	-	(3172.3)		168.1	-
Out: Flash effluent: liquid phase	1950.59	2056.78	14.468	289.4	10.321	206.4	1.402	189.0	199.3	1.56	137.5	3.426	136.3	110
solid phase	67.18	246.90	1.179	23.6	0.393	7.9	3.000	-	-	-	137.5	0.837	19.8	-
Flash steam for slurry preheating	-	-	0.850	17.0	-	-	-	-	-	-	(2738)		46.5	-
Pure condensate	-	-	2.650	53.0	2.650	53.0	1.000	-	-	-	178	4.238	40.0	-
Alkaline condensate	-	-	1.675	33.5	1.675	33.5	1.000	-	-	-	145	4.210	20.5	-
Heat loss + heat of reaction	-	-	-	-	-	-	-	-	-	-	-	-	7.3	-

* Figures in brackets are heat contents of steam phases in kJ/kg

Table 2-1/2 (cont.)

MATERIAL STREAM	Na20	A1203	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na20	A1203	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
FLASHING OF PURE CONDESATE														
In: High-pressure pure Condesate from digestion	-	-	2.650	53.0	2.650	53.0	1.000	-	-	-	178	4.238	40.0	-
Out: Low-pressure steam	-	-	0.159	3.2	-	-	-	-	-	-	(2732)		8.7	-
Pure condesate	-	-	2.491	49.8	2.491	49.8	1.000	-	-	-	140	4.205	29.3	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.0	-
DILUTION														
In: Flash efflvenct: liquid phase	1950.59	2056.78	14.468	289.4	10.321	206.4	1.402	189.0	199.3	1.56	137.5	3.426	136.3	110
solid phase	67.18	246.90	1.179	23.6	0.393	7.9	3.000	-	-	-	137.5	0.837	19.8	-
1st washer overflow	403.90	306.40	6.896	137.9	6.149	123.0	1.121	65.7	49.8	2.17	76	3.904	40.9	-
Spent liquor	140.91	71.72	1.221	24.4	0.994	19.9	1.064	141.8	72.2	3.23	79	3.711	7.2	-
Out: Diluted slurry	2562.58	2681.80	23.764	475.3	17.857	357.2	1.331	142.9	139.4	1.69	103	3.451	169.0	66
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	28.0	-
RED MUD SETTLING														
In: Diluted slurry: liquid phase	2495.40	2434.90	22.585	451.7	17.464	349.3	1.293	142.9	139.4	1.69	103	3.587	167.0	66
solid phase	67.18	246.90	1.179	23.6	0.393	7.9	3.000	-	-	-	103	0.837	2.0	-
Out: Settler overflow	2112.97	2068.95	19.234	384.7	14.891	297.8	1.292	141.9	138.9	1.68	96	3.590	132.6	-
Settler underflow:														
liquid phase	382.43	356.86	3.338	66.8	2.568	51.4	1.300	148.9	138.9	1.76	96	3.578	22.9	402
solid phase	67.18	256.00	1.193	23.9	0.399	8.0	2.992	-	-	-	96	0.837	1.9	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	11.6	-
RED MUD WASHING														
In: Settler underflow:														
liquid phase	382.43	356.86	3.338	66.8	2.568	51.4	1.300	148.9	138.9	1.76	96	3.578	22.9	402
solid phase	67.18	256.00	1.193	23.9	0.399	8.0	2.992	-	-	-	96	0.837	1.9	-
Filtrate	60.55	0.01	5.146	102.9	5.077	101.5	1.014	11.9	0.0	-	66	4.155	28.2	-
Alkaline hydrate wash water	23.78	11.89	0.631	12.6	0.593	11.9	1.064	-	-	-	60	4.033	3.0	-
Non-technological water	-	-	0.588	11.8	0.588	11.8	1.000	-	-	-	35	4.187	1.7	-
Out: 1st washer overflow	403.90	306.40	6.896	137.9	6.149	123.0	1.121	65.7	49.8	2.17	76	3.904	40.9	-
4th washer underflow:														
liquid phase	56.96	10.74	2.695	53.9	2.623	52.5	1.014	21.7	4.1	8.71	66	4.122	14.7	429
solid phase	73.10	307.62	1.306	26.1	0.422	8.4	2.954	-	-	-	66	0.837	1.4	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	0.7	-

Table 2-1/3 (cont.)

MATERIAL STREAM	Na2O	A1203	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	A1203	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
RED MUD CAUSTICIZATION AND FILTRATION														
In: 4th washer underflow:														
liquid phase	56.96	10.74	2.695	53.9	2.623	52.5	1.014	21.7	4.1	8.71	66	4.122	14.7	429
solid phase	73.10	307.62	1.306	26.1	0.422	8.4	2.954	-	-	-	66	0.837	1.4	-
Lime milk	-	-	0.608	12.2	0.547	10.9	1.111	-	-	-	75	3.589	3.3	150
Wash water	3.09	-	2.725	54.4	2.725	54.4	1.000	1.1	-	-	70	4.187	15.9	-
Low-pressure steam	-	-	0.180	3.6	-	-	-	-	-	-	(2792.6)		10.1	-
Out: Filtrate	60.55	0.01	5.146	102.9	5.077	101.5	1.014	11.9	0.0	-	66	4.155	28.2	-
Red mud to the pond:														
liquid phase	12.55	0.00	0.956	19.1	0.942	18.8	1.015	13.3	0.0	-	66	4.152	5.2	-
solid phase	60.05	318.36	1.412	28.2	0.478	9.6	2.954	-	-	-	66	0.837	1.6	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	10.4	-
MUD WASH WATER PREPARATION														
In: Digestion alkaline condesate	-	-	2.527	50.5	2.527	50.5	1.000	-	-	-	128	4.200	27.2	-
Evaporation alkaline condensate	-	-	1.123	22.5	1.123	22.5	1.000	-	-	-	75	4.187	7.0	-
Return water	3.09	-	0.618	12.4	0.618	12.4	1.000	5.0	-	-	35	4.187	1.8	-
Out: Red mud wash water	3.09	-	2.725	54.4	2.725	54.4	1.000	1.1	-	-	70	4.187	15.9	-
Heat loss + condesate loss	-	-	1.543	-	1.543	-	-	-	-	-	-	-	20.1	-
CONTROL FILTRATION														
In: Settler overflow	2112.97	2068.95	19.234	384.7	14.891	297.8	1.292	141.9	138.9	1.68	96	3.590	132.6	-
Out: Filtered aluminate liquor	2112.97	2068.95	19.234	384.7	14.891	297.8	1.292	141.9	138.9	1.68	94	3.590	129.8	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.8	-
HEAT INTERCHANGE														
In: Filtered aluminate liquor	2112.97	2068.95	19.234	384.7	14.891	297.8	1.292	141.9	138.9	1.68	94	3.590	129.8	-
Spent liquor	2064.13	1050.60	17.888	357.8	14.558	291.2	1.229	141.8	72.2	3.23	48	3.711	63.7	-
Out: Cooled aluminate liquor	2112.97	2068.95	19.234	384.7	14.891	297.8	1.292	141.9	138.9	1.68	63	3.590	87.0	-
Spent liquor	2064.13	1050.60	17.888	357.8	14.558	291.2	1.229	141.8	72.2	3.23	79	3.711	105.4	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	1.1	-

Table 2-1/4 (cont.)

MATERIAL STREAM	Na20	A1203	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na20	A1203	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
PRECIPITATION														
In: Cooled aluminate liquor	2112.97	2068.95	19.234	384.7	14.891	297.8	1.292	141.9	138.9	1.68	63	3.590	87.0	-
Seed hydrate: liquid phase	165.43	82.72	1.384	27.7	1.119	22.4	1.237	147.8	73.9	3.29	48	3.698	4.9	232
solid phase	-	2689.64	4.115	82.3	1.700	34.0	2.420	-	-	-	48	0.837	3.3	-
Cooling water	-	-	11.225	224.5	11.225	224.5	1.000	-	-	-	35	4.187	32.9	-
Out: Precipitated slurry	2278.40	4831.53	24.718	494.4	17.745	354.9	1.393	147.8	73.9	3.29	50	3.047	75.3	317
Precipitator's scales	-	9.78	0.015	0.3	0.006	0.1	2.420	-	-	-	50	0.837	-	
Cooling water	-	-	11.225	224.5	11.225	224.5	1.000	-	-	-	45	4.187	42.3	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	10.5	-
HYDRATE FILTRATION														
In: Precipitated slurry:														
liquid phase	2278.40	1157.74	19.097	382.0	15.423	308.5	1.237	147.8	73.9	3.29	50	3.698	70.6	317
solid phase	-	3673.79	5.621	112.4	2.322	46.4	2.420	-	-	-	50	0.837	4.7	-
Non-technological water	-	-	0.584	11.7	0.584	11.7	1.000	-	-	-	35	4.187	1.7	-
Out: Seed hydrate: liquid phase	165.43	82.72	1.384	27.7	1.119	22.4	1.237	147.8	73.9	3.29	48	3.698	4.9	-
solid phase	-	2689.64	4.115	82.3	1.700	34.0	2.420	-	-	-	48	0.837	3.3	-
Product hydrate:														
liquid phase	48.84	24.42	0.409	8.2	0.330	6.6	1.237	147.8	73.9	3.29	48	3.698	1.5	-
solid phase	-	984.15	1.506	30.1	0.622	12.4	2.420	-	-	-	48	0.837	1.2	-
Spent liquor	2064.13	1050.60	17.888	357.8	14.558	291.2	1.229	141.8	72.2	3.23	48	3.711	63.7	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.4	-
PRODUCT HYDRATE WASHING														
In: Product hydrate: liquid phase	48.84	24.42	0.409	8.2	0.330	6.6	1.237	147.8	73.9	3.29	48	3.698	1.5	-
solid phase	-	984.15	1.506	30.1	0.622	12.4	2.420	-	-	-	48	0.837	1.2	-
Wash water	-	-	0.908	18.2	0.908	18.2	1.000	-	-	-	75	4.187	5.7	-
Low-pressure steam	-	-	0.035	0.7	-	-	-	-	-	-	(2792.6)		2.0	-
Out: Product hydrate	-	984.15	1.506	30.1	0.622	12.4	2.420	-	-	-	55	0.837	1.4	-
Free water of product hydrate	4.90	2.45	0.185	3.7	0.177	3.5	1.044	27.7	13.8	3.29	55	4.074	0.8	-
Alkaline hydrate wash water	43.94	21.97	1.166	23.3	1.096	21.9	1.064	40.1	20.1	3.29	60	4.033	5.6	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.6	-

Table 2-1/5 (cont.)

MATERIAL STREAM	Na2O	Al2O3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	Al2O3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
EVAPORATION														
In: Spent liquor	847.54	431.37	7.347	146.9	5.978	119.6	1.229	141.8	72.2	3.23	79	3.711	43.3	-
Alkaline hydrate wash water	20.16	10.08	0.535	10.7	0.503	10.1	1.064	40.1	20.1	3.29	60	4.033	2.6	-
Cooling water	-	-	37.017	740.3	37.017	740.3	1.000	-	-	-	35	4.187	108.5	-
Non-technological water	-	-	0.373	7.5	0.373	7.5	1.000	-	-	-	35	4.187	1.1	-
Low-pressure steam	-	-	0.861	17.2	-	-	-	-	-	-	(2792.6)		48.0	-
Out: Strong liquor	867.70	441.45	5.065	101.3	3.664	73.3	1.382	236.8	120.5	3.23	70	3.481	24.7	-
Pure condensate	-	-	0.861	17.2	0.861	17.2	1.000	-	-	-	100	4.187	7.2	-
Alkaline condensate	-	-	2.552	51.0	2.552	51.0	1.000	-	-	-	75	4.187	16.0	-
Cooling water to tower	-	-	37.655	753.1	37.655	753.1	1.000	-	-	-	45	4.187	141.9	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	13.7	-
SALT REMOVAL														
In: Strong liquor	220.62	112.25	1.288	25.8	0.932	18.6	1.382	236.8	120.5	3.23	70	3.481	6.3	45
Make-up caustic	39.55	-	0.133	2.7	0.088	1.8	1.511	450.0	-	-	27	3.383	0.3	-
Low-pressure steam	-	-	0.119	2.4	-	-	-	-	-	-	(2792.6)		6.7	-
Cooling water	-	-	3.539	70.8	3.539	70.8	1.000	-	-	-	35	4.187	10.4	-
Non-technological water	-	-	0.373	7.5	0.373	7.5	1.000	-	-	-	35	4.187	1.1	-
Out: Wet salt	1.83	0.75	0.021	0.4	0.013	0.3	1.615	-	-	-	30	0.837	-	-
Salt free strong liquor to the digeston liquor preparation	255.10	107.49	1.773	35.5	1.380	27.6	1.285	184.9	77.9	3.90	65	3.629	8.4	-
Warmed water	-	-	3.539	70.8	3.539	70.8	1.000	-	-	-	45	4.187	13.3	-
Pure condensate	-	-	0.119	2.4	0.119	2.4	1.000	-	-	-	140	4.205	1.4	-
Mechanical losses	3.24	4.01	-	-	-	-	-	-	-	-	-	-	-	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	1.7	-

Table 2-1/6 (cont.)

MATERIAL STREAM	Na2O	Al2O3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	Al2O3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
DIGESTION LIQUOR ADJUSTMENT														
In: Strong liquor from evaporation	647.08	329.20	3.777	75.5	2.732	54.7	1.382	236.8	120.5	3.23	70	3.481	18.4	-
Strong liquor from salt removal	255.10	107.49	1.773	35.5	1.380	27.6	1.285	184.9	77.9	3.90	65	3.629	8.4	-
Spent liquor	1075.68	547.51	9.320	186.5	7.586	151.7	1.229	141.8	72.2	3.23	79	3.711	54.9	-
Make-up caustic + prec.scale	44.92	9.78	0.166	3.3	0.106	2.1	1.566	423.8	92.3	7.56	50	3.291	0.5	-
Out: Digestion liquor	2022.78	993.98	15.036	300.7	11.806	236.1	1.274	171.3	84.2	3.35	70	3.369	76.7	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	5.5	-
LIME MILK PREPARATION														
In: Burnt lime (61.9 % active CaO)	-	-	0.132	2.6	0.039	0.8	3.350	-	-	-	27	0.837	0.1	-
Alkaline condensate	-	-	0.521	10.4	0.521	10.4	1.000	-	-	-	75	4.187	3.3	-
Low-pressure steam	-	-	0.005	0.1	-	-	-	-	-	-	(2792.6)		0.3	-
Heat of reaction	-	-		-	-	-	-	-	-	-	-	-	2.2	-
Out: Lime milk to the causticization	-	-	0.608	12.2	0.547	10.9	1.111	-	-	-	75	3.589	3.3	150
Reject	-	-	0.050	1.0	0.015	0.3	3.350	-	-	-	75	0.837	0.1	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.5	-

Table 2-2/1

TECHNOLOGICAL MATERIAL AND HEAT BALANCE
April 1986. - January 1987.
Alumina production 18.08 tph

MATERIAL STREAM	Na20	A1203	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na20	A1203	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
BAUXITE GRINDING														
In: Bauxite: solid phase	-	1291.48	2.690	48.6	0.897	16.2	3.000	-	-	-	27	0.837	1.1	-
moisture	-	-	0.147	2.7	0.147	2.7	1.000	-	-	-	27	4.187	0.3	-
Digestion liquor	742.70	372.03	5.528	99.9	4.336	78.4	1.275	171.3	85.8	3.28	73	3.637	26.5	
Out: Mill slurry	742.70	1663.51	8.365	151.2	5.380	97.3	1.555	165.7	83.0	3.28	58	2.746	23.9	500
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	4.0	-
SLURRY ADJUSTMENT														
In: Mill slurry	742.70	1663.51	8.365	151.2	5.380	97.3	1.555	165.7	83.0	3.28	58	2.746	23.9	500
Digestion liquor	1278.72	640.49	9.519	172.1	7.465	135.0	1.275	171.3	85.8	3.28	73	3.637	45.9	-
Out: Adjusted slurry	2021.42	2304.00	17.884	323.3	12.845	232.3	1.392	169.2	84.7	3.28	65	3.220	68.1	209
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	1.7	-
SLURRY PREHEATING AND PREDESILICATION														
In: Adjusted slurry	2021.42	2304.00	17.884	323.3	12.845	232.3	1.392	169.2	84.7	3.28	65	3.220	68.1	209
Flash steam	-	-	0.864	15.6	-	-	-	-	-	-	(2721)		42.5	
Alkaline condensate	-	-	1.959	35.4	1.959	35.4	1.000	-	-	-	135	4.202	20.1	
Non-technological water	-	-	0.485	8.8	0.485	8.8	1.000	-	-	-	27	4.187	1.0	-
Out: Predesilicated slurry:														
liquid phase	1976.54	1916.56	17.027	307.8	12.977	234.6	1.312	152.3	147.7	1.70	87	3.559	95.3	100
solid phase	39.88	387.44	1.342	24.3	0.447	8.1	3.000	-	-	-	87	0.837	1.8	
Alkaline condensate	-	-	2.823	51.0	2.823	51.0	1.000	-	-	-	116	4.193	24.7	
Soda losses	5.00													
Heat loss + heat of reaction	-	-	-	-	-	-	-	-	-	-	-	-	9.9	-
DIGESTION														
In: Predesilicated slurry	2016.42	2304.00	18.370	332.1	13.425	242.7	1.368	152.3	147.7	1.70	87	3.360	97.1	100
High pressure steam	-	-	2.64	47.7	-	-	-	-	-	-	(3172.3))		151.3	-
Out: Flash effluent:														
liquid phase	1946.78	2067.30	14.371	259.8	10.219	184.8	1.406	190.5	202.3	1.55	125	3.419	111.0	111
solid phase	69.64	236.71	1.176	21.3	0.392	7.1	3.000	-	-	-	125	0.837	2.2	
Flash steam for slurry preheating	-	-	0.864	15.6	-	-	-	-	-	-	(2721)		42.5	-
Pure condensate	-	-	2.640	47.7	2.640	47.7	1.000	-	-	-	285	4.431	60.2	-
Alkaline condensate	-	-	1.959	35.4	1.959	35.4	1.000	-	-	-	135	4.202	20.1	-
Heat loss + heat of raction	-	-	-	-	-	-	-	-	-	-	-	-	12.4	-

Table 2-2/2 (cont.)

MATERIAL STREAM	Na2O	Al2O3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	Al2O3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
FLASHING OF PURE CONDENSATE														
In: High-pressure pure condensate from digestion	-	-	2.640	47.7	2.640	47.7	1.00	-	-	-	285	4.431	60.2	-
Out: Low-pressure steam	-	-	0.751	13.6	-	-	-	-	-	-	(2732)		37.2	-
Pure condensate	-	-	1.889	34.1	1.889	34.1	1.000	-	-	-	140	4.205	20.1	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.9	-
DILUTION														
In: Flash effluent:														
liquid phase	1946.78	2067.30	14.371	259.8	10.219	184.8	1.406	190.5	202.3	1.55	125	3.419	111.0	111
solid phase	69.64	236.71	1.176	21.3	0.392	7.1	3.000	-	-	-	125	0.837	2.2	-
1st washer overflow	480.91	353.07	7.805	141.1	6.927	125.2	1.127	69.4	51.0	2.24	73	3.894	40.1	-
Spent liquor	98.04	50.48	0.852	15.4	0.693	12.5	1.229	141.4	72.8	3.20	85	3.710	4.9	-
Out: Diluted slurry	2595.37	2707.65	24.204	437.6	18.231	329.6	1.328	141.6	138.5	1.68	101	3.457	152.8	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	5.4	-
RED MUD SETTLING														
In: Diluted slurry:														
liquid phase	2525.73	2470.85	23.028	416.3	17.839	322.5	1.291	141.6	138.5	1.68	101	3.591	151.0	65
solid phase	69.64	236.71	1.176	21.3	0.392	7.1	3.000	-	-	-	101	0.837	1.8	
Out: Settler overflow	2108.94	2077.37	19.378	350.4	15.032	271.8	1.289	140.3	138.2	1.67	98	3.594	123.4	-
Settler underflow:														
liquid phase	411.63	387.83	3.638	65.8	2.806	50.7	1.296	146.7	138.2	1.75	98	3.584	23.1	371
solid phase	74.80	242.35	1.188	21.5	0.395	7.1	3.004	-	-	-	98	0.837	1.8	
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	4.5	-
RED MUD WASHING														
In: Settler underflow:														
liquid phase	411.63	387.83	3.638	65.8	2.806	50.7	1.296	146.7	138.2	1.75	98	3.584	23.1	371
solid phase	74.80	242.35	1.188	21.5	0.395	7.1	3.004	-	-	-	98	0.837	1.8	-
Filtrate	81.50	6.75	6.303	114.0	6.204	112.2	1.016	13.1	1.1	19.87	64	4.149	30.7	-
Spent liquor	33.18	17.09	0.289	5.2	0.235	4.2	1.229	141.4	72.8	3.20	85	3.710	1.6	-
Alkaline hydrate wash water	32.86	16.28	0.753	13.6	0.700	12.7	1.075	46.9	23.2	3.32	60	4.009	3.3	-
Non-technological water	-	-	0.588	10.6	0.588	10.6	1.000	-	-	-	25	4.187	1.6	-
Out: 1st washer overflow	480.91	353.07	7.805	141.1	6.927	125.2	1.127	69.4	51.0	2.24	73	3.894	40.1	-
4th washer underflow:														
liquid phase	78.26	14.90	3.656	66.1	3.553	64.2	1.029	22.0	4.2	8.64	62	4.118	17.0	325
solid phase	74.80	302.33	1.297	23.4	0.439	7.9	2.958	-	-	-	62	0.837	1.2	
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	3.8	-

Table 2-2/3 (cont.)

MATERIAL STREAM	Na_2O	Al_2O_3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na_2O	Al_2O_3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
RED MUD CAUSTICIZATION AND FILTRATION														
In: 4th washer underflow:														
liquid phase	78.26	14.90	3.656	66.1	3.553	64.2	1.029	22.0	4.2	8.64	62	4.118	17.0	325
solid phase	74.89	302.33	1.297	23.4	0.439	7.9	2.958	-	-	-	62	0.837	1.2	
Lime milk	-	-	0.585	10.6	0.527	9.5	1.111	-	-	-	75	3.589	2.3	150
Wash water	2.1	-	2.873	51.9	2.873	51.9	1.000	0.7	-	-	68	4.187	14.8	-
Low-pressure steam (4.5 bar)	-	-	0.238	4.3	-	-	-	-	-	-	(2792.6)		12.0	-
Out: Filtrate	81.50	6.75	6.303	114.0	6.204	112.2	1.016	13.1	1.1	19.87	64	4.149	30.7	-
Red mud to the pond:														
liquid phase	13.73	0.16	0.947	17.1	0.933	16.9	1.014	2.3	2.2	128.90	68	4.151	4.6	-
solid phase	59.94	310.63	1.399	25.3	0.473	8.6	2.958	-	-	-	68	0.837	1.4	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	11.0	-
MUD WASH WATER PREPARATION														
In: Digestion alkaline condensate	-	-	2.823	51.0	2.823	51.0	1.000	-	-	-	116	4.193	24.7	-
Evaporation alkaline condensate	-	-	1.162	21.0	1.162	21.0	1.000	-	-	-	75	4.187	6.6	-
Return water	2.1	-	0.420	7.6	0.420	7.6	1.000	5.0	-	-	35	4.187	1.1	-
Out: Red mud wash water	2.1	-	2.873	51.9	2.873	51.9	1.000	0.7	-	-	68	4.187	14.8	-
Heat loss + condensate	-	-	1.532	27.7	1.532	27.7	1.000	-	-	-	-	-	17.6	-
CONTROL FILTRATION														
In: Settler overflow	2108.94	2077.37	19.378	350.4	15.032	271.8	1.289	140.3	138.2	1.67	98	3.594	123.4	-
Out: Filtered aluminate liquor	2108.94	2077.37	19.378	350.4	15.032	271.8	1.289	140.3	138.2	1.67	96	3.594	120.9	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.5	-
HEAT INTERCHANGE														
In: Filtered aluminate liquor	2108.94	2077.37	19.378	350.4	15.032	271.8	1.289	140.3	138.2	1.67	96	3.594	120.9	-
Spent liquor	2049.82	1055.44	17.811	322.0	14.493	262.0	1.229	141.4	72.8	3.20	47	3.710	56.0	-
Out: Cooled aluminate liquor	2108.94	2077.37	19.378	350.4	15.032	271.8	1.289	140.3	138.2	1.57	59	3.594	74.3	-
Spent liquor	2049.82	1055.44	17.811	322.0	14.493	262.0	1.229	141.4	72.8	3.20	85	3.710	101.2	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	1.4	-

Table 2-2/4 (cont.)

MATERIAL STREAM	Na20	A1203	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na20	A1203	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
PRECIPITATION														
In: Cooled aluminate liquor	2108.94	2077.37	19.378	350.4	15.032	271.8	1.289	140.3	138.2	1.67	59	3.594	74.3	-
Seed hydrate: liquor phase	155.75	77.16	1.314	23.8	1.065	19.3	1.234	146.2	72.4	3.32	47	3.702	4.1	224
solid phase	-	2596.71	3.973	71.8	1.642	29.7	2.420	-	-	-	47	0.837	2.8	-
Cooling water	-	-	5.409	97.8	5.409	97.8	1.000	-	-	-	47	4.187	14.3	-
Out: Precipitated slurry	2264.69	4741.92	24.651	445.7	17.775	321.4	1.387	146.2	72.4	3.32	49	3.063	66.8	308
Precipitator's scales	-	9.32	0.014	0.3	0.006	0.1	2.420	-	-	-	49	0.837	-	-
Cooling water	-	-	5.409	97.8	5.409	97.8	1.000	-	-	-	45	4.187	18.4	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	10.3	-
HYDRATE FILTRATION														
In: Precipitated slurry:														
liquid phase	2264.69	1161.89	19.174	346.7	15.511	280.5	1.234	146.2	72.4	3.32	49	3.702	14.3	308
solid phase	-	3580.03	5.477	99.0	2.264	40.9	2.420	-	-	-	49	0.837	4.1	-
Non-technological water	-	-	0.451	8.2	0.451	8.2	1.000	-	-	-	35	4.187	1.2	-
Out: Seed hydrate: liquid phase	155.75	77.16	1.314	23.8	1.065	19.3	1.234	146.2	72.4	3.32	47	3.702	4.1	-
solid phase	-	2596.71	3.973	71.8	1.642	29.7	2.420	-	-	-	47	0.837	2.8	-
Product hydrate: liquid phase	59.12	29.29	0.499	9.0	0.404	7.3	1.234	146.2	72.4	3.32	47	3.702	1.6	-
solid phase	-	983.32	1.504	27.2	0.622	11.2	2.420	-	-	-	47	0.837	1.1	-
Spent liquor	2049.82	1055.44	17.811	322.0	14.493	262.0	1.229	141.4	72.8	3.20	47	3.710	56.0	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.4	-
PRODUCT HYDRATE WASHING														
In: Product hydrate: liquid phase	59.12	29.29	0.499	9.0	0.404	7.3	1.234	146.2	72.4	3.32	47	3.702	1.6	-
solid phase	-	983.32	1.504	27.2	0.622	11.2	2.420	-	-	-	47	0.837	1.1	-
Wash water	-	-	0.897	16.2	0.897	16.2	1.000	-	-	-	75	4.187	5.1	-
Low-pressure steam	-	-	0.038	0.7	-	-	-	-	-	-	(2792.6)		1.9	-
Out: Product hydrate	-	983.32	1.504	27.2	0.622	11.2	2.420	-	-	-	55	0.837	1.3	-
Free water of product hydrate	4.80	2.38	0.189	3.4	0.181	3.3	1.042	26.5	13.1	3.32	55	4.076	0.8	-
Alkaline hydrate wash water	54.32	26.91	1.245	22.5	1.158	20.9	1.075	46.9	23.2	3.32	60	4.008	5.4	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.2	-

Table 2-2/5 (cont.)

MATERIAL STREAM	Na20	A1203	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na20	A1203	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
EVAPORATION														
In: Spent liquor	884.73	455.52	7.750	140.1	6.306	114.0	1.229	140.3	72.8	3.20	85	3.710	44.0	-
Alkaline hydrate wash water	21.46	10.63	0.492	8.9	0.458	8.3	1.075	46.9	23.2	3.32	60	4.008	2.1	-
Cooling water	-	-	37.133	671.4	37.133	671.4	1.000	-	-	-	35	4.187	98.4	-
Non-technological water	-	-	0.400	7.2	0.400	7.2	1.000	-	-	-	35	4.187	1.1	-
Low-pressure steam	-	-	0.842	15.2	-	-	-	-	-	-	(2792.6)		42.4	-
Out: Strong liquor	906.19	466.15	5.442	98.4	3.964	71.7	1.373	228.6	117.6	3.20	70	3.498	24.1	-
Pure condensate	-	-	0.842	15.2	0.842	15.2	1.000	-	-	-	100	4.187	6.4	-
Alkaline condensate	-	-	2.560	46.3	2.560	46.3	1.000	-	-	-	75	4.187	14.5	-
Cooling water to tower	-	-	37.773	682.9	37.773	682.9	1.000	-	-	-	45	4.187	128.7	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	14.3	-
SALT REMOVAL														
In: Strong liquor	230.40	118.54	1.384	25.0	1.008	18.2	1.373	228.6	117.6	3.20	70	3.498	6.1	34
Make-up caustic	41.20	-	0.139	2.5	0.092	1.7	1.511	450.0	-	-	27	3.383	0.2	-
Low-pressure steam	-	-	0.127	2.3	-	-	-	-	-	-	(2792.6)		6.4	-
Cooling water	-	-	2.022	36.6	2.022	36.6	1.000	-	-	-	35	4.187	5.4	-
Non-technological water	-	-	0.400	7.2	0.400	7.2	1.000	-	-	-	35	4.187	1.1	-
Out: Wet salt	1.69	0.84	0.017	0.3	0.011	0.2	1.660	-	-	-	30	0.837	-	-
Salt-free strong liquor to the digestion liquor preparation	267.61	123.24	1.906	34.5	1.489	26.9	1.280	179.7	82.8	3.57	65	3.627	8.1	-
Warmed water	-	-	2.022	36.6	2.022	36.6	1.000	-	-	-	45	4.187	6.9	-
Pure condensate	-	-	0.127	2.3	0.127	2.3	1.000	-	-	-	140	4.205	1.4	-
Mechanical losses	2.40	-5.54	-	-	-	-	-	-	-	-	-	-	-	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.8	-
DIGESTION LIQUOR ADJUSTMENT														
In: Strong liquor from evaporation	675.79	347.61	4.058	73.4	2.956	53.5	1.373	228.6	117.6	3.20	70	3.498	18.0	-
Strong liquor from salt removal	267.61	123.24	1.906	34.5	1.489	26.9	1.280	179.7	82.8	3.57	65	3.627	8.1	-
Spent liquor	1033.87	532.35	8.920	161.3	7.259	131.2	1.229	140.3	72.8	3.20	85	3.710	50.7	-
Make-up caustic + prec.scale	44.15	9.32	0.162	2.9	0.104	1.9	1.558	424.5	89.6	7.79	50	3.290	0.5	-
Out: Digestion liquor	2021.42	1012.52	15.047	272.0	11.801	213.4	1.275	171.3	85.8	3.28	73	3.637	72.4	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	4.9	-

Table 2-2/6 (cont.)

MATERIAL STREAM	Na_2O	Al_2O_3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na_2O	Al_2O_3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
LIME MILK PREPARATION														
In: Burnt lime (63.3 % active CaO)	-	-	0.125	2.3	0.037	0.7	3.35	-	-	-	27	0.837	-	-
Alk. condensate	-	-	0.501	9.0	0.501	9.0	1.000	-	-	-	75	4.387	2.0	-
Low-pressure steam	-	-	0.005	0.1	-	-	-	-	-	-	(2792.6)		0.3	-
Heat of reaction	-	-	-	-	-	-	-	-	-	-	-	-	2.0	-
Out: Lime milk to the causticization	-	-	0.585	10.6	0.527	9.5	1.111	-	-	-	75	3.589	2.9	150
Reject	-	-	0.046	0.8	0.014	0.2	3.35	-	-	-	75	0.837	-	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.2	-

Table 2-3/1

TECHNOLOGICAL MATERIAL AND HEAT BALANCE
February - November 1987.
Alumina production 18.25 tph

MATERIAL STREAM	Na20	A1203	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na20	A1203	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
BAUXITE GRINDING														
In: Bauxite: solid phase	-	1272.70	2.651	48.4	0.884	16.1	3.000	-	-	-	27	0.837	1.1	-
moisture	-	-	0.108	2.0	0.108	2.0	1.000	-	-	-	27	4.187	0.2	-
Digestion liquor	712.58	358.05	5.456	99.6	4.310	78.7	1.266	165.3	83.1	3.27	74	3.651	26.8	-
Out: Mill slurry	712.58	1630.95	8.215	150.0	5.302	96.8	1.549	161.3	81.2	3.27	59	2.753	24.2	500
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	3.9	-
SLURRY ADJUSTMENT														
In: Mill slurry	712.58	1630.95	8.215	150.0	5.302	96.8	1.549	161.3	81.2	3.27	59	2.753	24.2	500
Digestion liquor	1258.72	632.75	9.636	175.9	7.613	138.9	1.266	165.3	83.1	3.27	74	3.651	47.3	-
Lime milk	-	-	0.280	5.1	0.252	4.6	1.111	-	-	-	75	3.569	1.4	150
Out: Adjusted slurry	1971.30	2263.70	18.131	330.9	13.167	240.3	1.377	160.6	80.8	3.27	66	3.243	71.1	204
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	1.8	-
SLURRY PREHEATING AND PREDESILICATION														
In: Adjusted slurry	1971.30	2263.70	18.131	330.9	13.167	240.3	1.377	160.6	80.8	3.27	66	3.243	71.1	204
Flash steam	-	-	0.974	17.8	-	-	-	-	-	-	(2721)		[illegible]	-
Alkaline condensate	-	-	1.837	33.5	1.837	33.5	1.000	-	-	-	135	4.202		-
Non-technological water	-	-	0.296	5.4	0.296	5.4	1.000	-	-	-	35	4.187	[illegible]	-
Out: Predesilicated slurry:														
liquid phase	1927.06	1881.89	17.061	311.4	13.109	239.2	1.301	147.0	143.5	1.69	90	3.574	100.2	100
solid phase	39.24	381.81	1.367	24.9	0.456	8.4	3.000	-	-	-	90	0.837	1.9	-
Alkaline condensate	-	-	2.811	51.3	2.811	51.3	1.000	-	-	-	116	4.187	25.0	-
Heat loss + heat of reaction	-	-	-	-	-	-	-	-	-	-	-	-	12.2	-
DIGESTION														
In: Predesilicated slurry	1966.30	2263.70	18.428	336.3	13.565	247.6	1.359	147.0	143.5	1.69	90	3.371	102.1	100
High-pressure steam	-	-	2.640	48.2	-	-	-	-	-	-	(3172.3)		152.9	-
Out: Flash effluent: liquid phase	1899.05	2047.98	14.430	263.3	10.350	188.9	1.394	183.5	197.9	1.525	126.5	3.435	114.7	110
solid phase	68.65	215.72	1.186	21.6	0.395	7.2	3.000	-	-	-	126.5	0.837	2.3	-
Flash steam for slurry preheating	-	-	0.974	17.8	-	-	-	-	-	-	(2721)		48.4	-
Pure condensate	-	-	2.640	48.2	2.640	48.2	1.000	-	-	-	285	4.431	60.9	-
Alkaline condensate	-	-	1.837	33.5	1.837	33.5	1.000	-	-	-	135	4.202	19.0	-
Heat loss + heat of reaction	-	-	-	-	-	-	-	-	-	-	-	-	9.7	-

Table 2-3/2 (cont.)

MATERIAL STREAM	Na2O	A12O3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	A12O3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
FLASHING OF PURE CONDENSATE														
In: High-pressure pure condensate from digestion	-	-	2.640	48.2	2.640	48.2	1.000	-	-	-	285	4.431	60.9	-
Out: Low-pressure steam	-	-	0.753	13.8	-	-	-	-	-	-	(2732)		37.7	-
Pure condensate	-	-	1.887	34.4	1.887	34.4	1.000	-	-	-	140	4.205	20.3	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.9	-
DILUTION														
In: Flash effluent: liquor phase	1899.05	2047.98	14.430	263.3	10.350	188.9	1.394	183.5	197.9	1.525	126.5	3.435	114.7	110
solid phase	68.65	215.72	1.186	21.6	0.395	7.2	3.000	-	-	-	126.5	0.837	2.3	-
1st washer overflow	402.46	302.64	6.795	118.5	6.053	110.5	1.122	66.5	50.0	2.19	72	3.902	33.3	-
Spent liquor	118.45	61.40	1.034	18.9	0.842	15.4	1.229	140.7	72.9	3.17	86	3.712	6.0	-
Out: Diluted slurry	2488.61	2627.74	23.445	427.8	17.640	322.0	1.329	140.3	139.9	1.65	101	3.496	150.9	67
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	5.4	-
RED MUD SETTLING														
In: Diluted slurry: liquid phase	2419.96	2412.02	22.259	406.2	17.245	314.8	1.291	140.3	139.9	1.65	101	3.638	149.1	67
solid phase	68.65	215.72	1.186	21.6	0.395	7.2	3.000	-	-	-	101	0.837	1.8	-
Out: Setller overflow	2047.94	2054.19	18.989	346.5	14.793	268.9	1.289	139.0	139.4	1.64	98	3.593	122.0	-
Settler underflow: liquid phase	366.93	349.88	3.255	59.4	2.503	45.8	1.297	146.2	139.4	1.73	98	3.582	20.9	413
solid phase	73.74	223.66	1.202	21.9	0.400	7.3	3.002	-	-	-	98	0.837	1.8	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	6.2	-
RED MUD WASHING														
In: Settler underflow: liquid phase	366.93	349.88	3.255	59.4	2.509	45.8	1.297	146.2	139.4	1.73	98	3.582	20.9	413
solid phase	73.74	223.66	1.202	21.9	0.400	7.3	3.002	-	-	-	98	0.837	1.8	-
Filtrate	65.20	0.00	5.268	96.1	5.184	94.6	1.014	12.5	-	-	66	4.153	26.3	-
Alkaline hydrate wash water	32.42	16.52	0.671	12.2	0.619	11.3	1.085	52.4	26.7	3.23	60	3.986	2.9	-
Non-technological water	-	-	0.588	10.7	0.588	10.7	1.000	-	-	-	35	4.187	1.6	-
Out: 1st washer overflow	402.46	302.64	6.795	118.5	6.053	110.5	1.122	66.5	50.0	2.19	72	3.902	33.3	-
4th washer underflow:														
liquid phase	62.09	7.96	2.884	52.6	2.806	51.2	1.028	22.1	2.8	12.83	63	4.131	13.7	402
solid phase	73.74	279.46	1.305	23.8	0.441	8.0	2.960	-	-	-	63	0.837	1.3	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	5.2	-

Table 2-3/3 (cont.)

MATERIAL STREAM	Na2O	Al2O3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	Al2O3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
RED MUD CAUSTICIZATION AND FILTRATION														
In: 4th washer undeflow:														
liquid phase	62.09	7.96	2.884	52.6	2.806	51.2	1.028	22.1	2.8	12.83	63	4.131	13.7	402
solid phase	73.74	279.46	1.305	23.8	0.441	8.0	2.960	-	-	-	63	0.837	1.3	-
Lime milk	-	-	0.488	8.9	0.440	8.0	1.111	-	-	-	75	3.589	2.4	150
Wash water	2.62	-	2.716	49.6	2.716	49.6	1.000	1.0	-	-	70	4.187	14.5	-
Low-pressure steam	-	-	0.209	3.8	-	-	-	-	-	-	(2792.6)		10.7	-
Out: Filtrate	65.20	0.00	5.268	96.1	5.184	94.6	1.014	12.5	-	-	66	4.153	26.3	-
Red mud to the pond:														
liquid phase	8.97	0.00	0.938	17.1	0.931	17.0	1.008	9.6	-	-	70	4.169	5.0	-
solid phase	64.29	287.42	1.396	25.5	0.472	8.6	2.960	-	-	-	70	0.837	1.5	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	9.8	-
MUD WASH WATER PREPARATION														
In: Digestion alkaline condensate	-	-	2.811	51.3	2.811	51.3	1.000	-	-	-	116	4.187	25.0	-
Evaporation alkaline condensate	-	-	0.982	17.9	0.982	17.9	1.000	-	-	-	75	4.187	5.6	-
Return water	2.62	-	0.524	9.6	0.524	9.6	1.000	5.0	-	-	35	4.187	1.4	-
Out: Red mud wash water	2.62	-	2.716	49.6	2.716	49.6	1.000	1.0	-	-	70	4.187	14.5	-
Heat loss + condensate loss	-	-	1.601	29.2	1.601	29.2	1.000	-	-	-	-	-	17.5	-
CONTROL FILTRATION														
In: Settler overflow	2047.94	2054.19	18.989	346.5	14.733	268.9	1.289	139.0	139.4	1.64	98	3.593	122.0	-
Out: Filtered aluminate liquor	2047.94	2054.19	18.989	346.5	14.733	268.9	1.289	139.0	139.4	1.64	96	3.593	119.5	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.5	-
HEAT INTERCHANGE														
In: Filtered aluminate liquor	2047.94	2054.19	18.989	346.5	14.733	268.9	1.289	139.0	139.4	1.64	96	3.593	119.5	-
Spent liquor	1987.89	1030.33	17.358	316.7	14.133	257.9	1.228	140.6	72.9	3.16	50	3.712	58.8	-
Cut: Cooled aluminate liquor	2047.94	2054.19	18.989	346.5	14.733	268.9	1.289	139.0	139.4	1.64	61	3.593	75.9	-
Spent liquor	1987.89	1030.33	17.358	316.7	14.133	257.9	1.228	140.6	72.9	3.18	86	3.712	101.1	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	1.3	-

Table 2-3/4 (cont.)

MATERIAL STREAM	Na_2O	Al_2O_3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na_2O	Al_2O_3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
PRECIPITATION														
In: Cooled aluminate liquor	2047.94	2054.19	18.989	346.5	14.733	268.9	1.289	139.0	139.4	1.64	61	3.593	75.9	-
Seed hydrate: liquid phase	173.80	88.53	1.480	27.0	1.200	21.9	1.234	144.9	73.8	3.23	50	3.702	5.0	246
solid phase	-	2855.32	4.369	79.7	1.805	32.9	2.420	-	-	-	50	0.837	3.3	-
Cooling water	-	-	5.436	99.2	5.436	99.2	1.000	-	-	-	35	4.187	14.5	-
Out: Precipitated slurry	2221.74	4988.98	24.823	453.0	17.775	324.4	1.397	144.9	73.8	3.23	52	3.023	71.2	330
Precipitator's scales	-	9.07	0.014	0.3	0.006	0.1	2.420	-	-	-	52	0.837	-	-
Cooling water	-	-	5.436	99.2	5.436	99.2	1.000	-	-	-	40	4.187	16.6	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	10.9	-
HYDRATE FILTRATION														
In: Precipitated slurry:														
liquid phase	2221.74	1149.45	18.948	345.8	15.348	280.1	1.235	144.9	73.8	3.23	52	3.702	66.5	330
solid phase	-	3839.53	5.875	107.2	2.427	44.3	2.420	-	-	-	52	0.837	4.7	-
Non-technological water	-	-	0.400	7.3	0.400	7.3	1.000	-	-	-	35	4.187	1.1	-
Out: Seed hydrate: liquid phase	173.80	88.53	1.480	27.0	1.200	21.9	1.234	144.9	73.8	3.23	50	3.702	5.0	-
solid phase	-	2855.32	4.369	79.7	1.805	32.9	2.420	-	-	-	50	0.837	3.3	-
Product hydrate: liquid phase	60.05	30.59	0.511	9.3	0.415	7.6	1.234	144.9	73.8	3.23	50	3.703	1.7	-
solid phase	-	984.21	1.506	27.5	0.622	11.4	2.420	-	-	-	50	0.837	1.2	-
Spent liquor	1987.89	1030.33	17.358	316.7	14.133	257.9	1.228	140.6	72.9	3.18	50	3.712	58.8	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.3	-
PRODUCT HYDRATE WASHING														
In: Product hydrate: liquid phase	60.05	30.59	0.511	9.3	0.415	7.6	1.234	144.9	73.8	3.23	50	3.702	1.7	-
solid phase	-	984.21	1.506	27.5	0.622	11.4	2.420	-	-	-	50	0.837	1.2	-
Wash water	-	-	0.798	14.6	0.798	14.6	1.000	-	-	-	75	4.187	4.6	-
Low-pressure steam	-	-	0.035	0.6	-	-	-	-	-	-	(2792.6)		1.7	-
Out: Product hydrate	-	984.21	1.506	27.5	0.622	11.4	2.420	-	-	-	55	0.837	1.3	-
Free water of product hydrate	4.70	2.39	0.198	3.6	0.191	3.5	1.040	24.6	12.6	3.23	55	4.102	0.8	-
Alkaline hydrate wash water	55.35	28.20	1.146	20.9	1.056	19.3	1.085	52.4	26.7	3.23	60	3.986	5.0	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.1	-

Table 2-3/5 (cont.)

MATERIAL STREAM		Na_2O	Al_2O_3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na_2O	Al_2O_3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
		kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
EVAPORATION															
In:	Spent liquor	845.83	438.25	7.386	134.8	6.015	109.8	1.228	140.6	72.9	3.18	86	3.712	43.0	-
	Alkaline hydrate wash water	22.93	11.68	0.475	8.7	0.438	8.0	1.085	52.4	26.7	3.23	60	3.986	2.1	-
	Cooling water	-	-	33.855	617.9	33.855	617.9	1.000	-	-	-	35	4.187	90.6	-
	Non-technological water	-	-	0.540	9.9	0.540	9.9	1.000	-	-	-	35	4.187	1.5	-
	Low-pressure steam	-	-	0.927	16.9	-	-	-	-	-	-	(2792.6)		47.2	-
Out:	Strong liquor	868.76	449.93	5.341	97.5	3.933	71.8	1.358	220.9	114.4	3.18	70	3.512	23.8	-
	Pure condensate	-	-	0.927	16.9	0.927	16.9	1.000	-	-	-	100	4.187	7.1	-
	Alkaline condensate	-	-	2.448	44.7	2.448	44.7	1.000	-	-	-	75	4.187	13.8	-
	Cooling water to tower	-	-	34.467	629.0	34.467	629.0	1.000	-	-	-	45	4.187	118.5	-
	Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	21.2	-
SALT REMOVAL															
In:	Strong liquor	220.89	114.40	1.358	24.8	1.000	18.3	1.358	220.9	114.4	3.18	70	3.512	6.1	32
	Make up caustic	39.60	-	0.133	2.4	0.088	1.6	1.510	450.0	-	-	27	3.383	0.2	-
	Low-pressure steam	-	-	0.125	2.3	-	-	-	-	-	-	(2792.6)		6.4	-
	Cooling water	-	-	2.137	39.0	2.137	39.0	1.000	-	-	-	35	4.187	5.7	-
	Non-technological water	-	-	0.540	9.9	0.540	9.9	1.000	-	-	-	35	4.187	1.5	-
Out:	Wet salt	1.21	0.6	0.016	0.3	0.010	0.2	1.660	-	-	-	30	0.837	-	-
	Salt-free strong liquor to the digestion liquor preparation	257.26	115.73	2.015	36.8	1.618	29.5	1.245	159.0	71.5	3.66	65	3.682	8.8	-
	Warmed water	-	-	2.137	39.0	2.137	39.0	1.000	-	-	-	45	4.187	7.3	-
	Pure condensate	-	-	0.125	2.3	0.125	2.3	1.000	-	-	-	140	4.205	1.4	-
	Mechanical losses	2.02	-1.93	-	-	-	-	-	-	-	-	-	-	-	-
	Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.4	-
DIGESTION LIQUOR ADJUSTMENT															
In:	Strong liquor from evaporation	647.87	335.53	3.983	72.7	2.933	53.5	1.358	220.9	114.4	3.18	70	3.512	17.7	-
	Strong liquor from salt removal	257.26	115.73	2.015	36.8	1.618	29.5	1.245	159.0	71.5	3.66	65	3.682	8.8	-
	Spent liquor	1023.61	530.68	8.938	163.1	7.276	132.8	1.229	140.7	72.9	3.17	86	3.712	52.1	-
	Make-up caustic + prec. scale	42.57	9.07	0.157	2.9	0.101	1.8	1.554	421.5	89.8	7.72	50	3.293	0.5	-
Out:	Digestion liquor	1971.30	991.00	15.092	275.4	11.923	217.6	1.266	165.3	83.1	3.27	74	3.651	74.1	-
	Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	5.0	-

Table 2-3/6 (cont.)

MATERIAL STREAM	Na_2O	Al_2O_3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na_2O	Al_2O_3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m^3/t	m^3/h	t/m^3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
LIME MILK PREPARATION														
In: Burnt lime (64.2 active CaO)	-	-	0.141	2.6	0.042	0.8	3.35	-	-	-	27	0.837	0.1	-
Alkaline condensate	-	-	0.668	12.1	0.668	12.2	1.000	-	-	-	75	4.187	3.8	-
Low-pressure steam	-	-	0.005	0.1	-	-	-	-	-	-	(2792.6)		0.3	-
Heat of reaction	-	-	-	-	-	-	-	-	-	-	-	-	2.3	-
Out: Lime milk to the causticization	-	-	0.488	8.9	0.440	8.0	1.111	-	-	-	75	3.589	2.4	150
Lime milk to the slurry adjustment	-	-	0.280	5.1	0.252	4.6	1.111	-	-	-	75	3.589	1.4	150
Reject	-	-	0.042	0.8	0.013	0.2	3.35	-	-	-	75	0.837	0.1	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.6	-

Table 2-[illegible]

CHARACTERISTIC TECHNOLOGICAL PARAMETERS

	Units	April 1985- March 1986.	April 1986- January 1987.	February 1987- November 1987.	December 1987- January 1988.	Remarks
Alumina production	tpy	175,960	158,400	159,900	162,700	calculated
Alumina production	tph	20.08	18.10	18.25	18.57	
Bauxite quality:						
Al_2O_3	%	48.33	48.01	48.01	47.75	
SiO_2	%	3.55	3.52	3.53	4.05	
Fe_2O_3	%	16.05	16.50	16.55	17.30	
TiO_2	%	8.24	8.21	8.24	8.26	
L.o.I.	%	23.09	22.94	2[illegible].53	22.36	
Free water	%	4.33	4.92	3.90	4.00	
Digestion liquor						
temperature	oC	70	73	74	75	estimated
caustic Na_2O	gp	171.6	171.3	165.3	169.9	
molar ratio	-	3.35	3.28	3.27	2.88	
Lime addition	%	-	-	1.2	-	
Specific bauxite consumption	t/t	2.71	2.69	2.65	3.11	calculated
Predesilication						
temperature in the first tank	oC	92.3	96.5	99	97.3	
temperature in the last tank	oC	84	87	90	88	estimated
Digestion temperature	oC	237.5	238	238	238.7	
Final temperature of heat recuperation	oC	162.1	167.1	167.0	166.5	
Temperature in the last flash tank	oC	137.5	125	126.5	131.5	
Holding time at the digestion temperature	min	45	50	49	45	
Final digestion caustic molar ratio		1.56	1.55	1.525	1.50	
Digestion efficiency	%	81.1	81.3	82.9	75.1	
Non-technological water from grinding to flashing	m3/t	0.305	0.485	0.296	0.445	calculated
Flashed water	m3/t	2.527	2.821	2.811	3.088	
Underflow solids concentrations						
settlers	gpl	402	371	413	345	
washers	gpl	455	360	440	466	average
Settler overflow						
temperature	oC	96	98	98	96	
caustic Na_2O	gpl	141.9	140.3	139.0	140.2	
caustic molar ratio		1.68	1.67	1.64	1.64	
Na_2O/SiO_2 ratio in the red mud		0.758	0.790	0.777	0.817	

Table 2-4/2 (cont.)

	Units	April 1985- March 1987.	April 1986- January 1987.	February 1987- November 1987.	December 1987- January 1987.	Remarks
Non-technological water in the mud washing	m3/t	0.588	0.588	0.588	0.185	calculated
Auto-precipitation	kg/t	60.72	65.63	63.75	112.2	
Causticization efficiency	%	17.8	19.8	12.4	11.2	
temperature	oC	89	89	91	91	
Lime addition	kg/t	82	79	66	109	
Burnt lime, active CaO	%	61.9	63.3	64.2	66.0	
Lime milk concentration	gpl	150	150	150	150	estimated
Wash water temperature	oC	70	68	70	70	
Filtrate to the 4th washer	oC	66	64	66	66	
Aluminate liquor tempeature						
before cooling	oC	94	96	96	94	
after cooling	oC	62.6	59	61	60.5	
Spent liquor temperature						
before heat interchange	oC	47.9	46.9	50.1	52.0	
after heat interchange	oC	79.4	84.4	86.1	86.1	
Carbonate level as Na2O in carbonate/Na2O caustic	%	9.53	16.41	22.80	24.42	
Average holding time	hrs	no data	81.6	79.0	71.0	
Seed ratio		1.3	1.25	1.39	0.8	
Mid-stage cooling	oC	5	3	3	3	estimated
Non-technological water from liquor cooling to hydrate filtration	m3/t	0.584	0.451	0.400	0.473	calculated
Temperature in the first precipitator tank	oC	60.9	57	59.2	58.5	
Temperature in the last precipitator tank	oC	49.9	48.9	52.1	53	
Precipitation efficiency	%	48.6	49.7	49.2	42.1	
Free water content of wet seed	%	18	18	18	18	estimated
Free water content of washed hydrate	%	10.3	10.5	10.96	11.7	
Hydrate wash water	m3/t	0.908	0.897	0.798	1.438	
Alkaline hydrate wash water						
caustic Na2O	gpl	40.1	46.9	52.4	34.1	
for evaporation	m3/t	0.503	0.458	0.438	1.236	
Hydrate in spent liquor	gpl	1.41	2.85	1.29	2.69	calculated
Precipitator's scale	kg/t	9.6	10.0	9.1	8.16	

Table 2-4/3 (cont.)

	Units	April 1985- March 1986.	April 1986- January 1987.	February 1987- November 1987.	December 1987- January 1988.	Remarks
Evaporated water	t/t	3.19	3.20	3.06	3.61	
Non-technological water from evaporation to digestion liquor adjustment	m3/t	0.746	0.800	1.08	0.435	calculated
Strong liquor						
caustic Na20	gpl	236.8	228.6	220.9	232.2	
temperature	oC	75	75	75	75	
Removed salt	t/t	0.021	0.0175	0.0154	0.0188	

3. Suggestions for improving the plant operation

Based on the experiences gathered during the site mission and on the conclusions drawn from the technological data, suggestions are made to improve the plant operation. Though parts of our suggestions concern such smaller modifications or methods, which were already handed over to the experts of the Korba Alumina Plant, nevertheless they will be explained here in this description in details.

3.1 Suggestions concerning the technology

The bauxite supply of the Korba Alumina Plant is in a critical situation. This time the plant processes bauxite originating from the Phutka-Pahar and Amarkantak sites. The bauxite resources are running out at the mentioned sites. Earlier, in 1979 a study was prepared and handed over by Aluterv-FKI on the possibility of the processing of the Ghandamardan bauxite. Another possibility which emerged lately is the supply of the alumina plant with the bauxite produced by Nalco. A series of thorough technological tests with the representative samples of this bauxite have to precede the possible change-over to the Nalco bauxite, with the following main elements:

- the chemical and phase analysis of the bauxite,
- the determination of the optimum digestion temperature,
- the determination of the optimum digestion molar ratio (possibly the utilization of the so called sweetening digestion technology may come into question, the essence of which is that after a digestion done at a somewhat higher

digestion molar ratio, by adding gibbsitic bauxite at an appropriate temperature during the flashing process, the molar ratio is reduced significantly),

- the consideration of the changes at the other points of the process required by the changes in the conditions of the digestion (evaporation, removal of impurities, etc.).

The processing of the existing bauxite was considered when giving our suggestions for the improvement of the technology. Our suggestions include the improvement of the digestion, the purification of the process liquor (together with this the improvement of the precipitation efficiency) and several smaller solutions resulting in caustic soda and energy savings. The instrumentation level of the alumina plant was reviewed and suggestions were made for the instrumentation of some operations which have not yet been automated.

The material and heat balances for the modified process are given in Table 3.1-1. The characteristic technological data are summarized in Table 3.1-2.

3.1.1 Updating the Digestion

The most important task in the digestion plant unit is the increase of the digestion yield. From the analysis of the bauxite and mud samples gathered during the February site mission it can be seen that a high amount of undigested boehmite is present in the mud (see Tables 3.1.1-1; 3.1.1-2).

Table 3.1-1/1

TECHNOLOGICAL MATERIAL AND HEAT BALANCE
Projected for 1989
Alumina production 25 tph

MATERIAL STREAM	Na2O	Al2O3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	Al2O3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
BAUXITE GRINDING														
In: Bauxite: solid phase	-	1197.64	2.495	62.4	0.832	20.8	3.000	-	-	-	27	0.837	1.4	-
moisture	-	-	0.101	2.5	0.101	2.5	1.000	-	-	-	27	4.187	0.3	-
Digestion liquor	726.43	345.36	5.205	130.1	4.057	101.4	1.283	179.1	85.1	3.46	72	3.625	34.0	-
Out: Mill slurry	726.43	1543.00	7.801	195.0	4.990	124.8	1.563	174.7	83.1	3.46	57	2.741	30.7	500
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	5.0	-
SLURRY ADJUSTMENT														
In: Mill slurry	726.43	1543.00	7.801	195.0	4.990	124.8	1.563	174.7	83.1	3.46	57	2.741	30.7	500
Lime milk	-	-	0.172	4.3	0.150	3.8	1.148	-	-	-	75	3.738	1.2	200
Digestion liquor	1148.48	546.03	8.232	205.8	6.414	160.4	1.283	179.1	85.1	3.46	72	3.625	53.8	-
Out: Adjusted slurry	1874.91	2089.03	16.205	405.1	11.554	288.9	1.403	175.0	83.2	3.46	65	3.201	84.0	219
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	1.7	-
SLURRY PREHEATING AND PREDESILICATION														
In: Adjusted slurry	1874.91	2089.03	16.205	405.1	11.554	288.9	1.403	175.0	83.2	3.46	65	3.201	84.0	219
Flash steam	-	-	0.697	17.4	-	-	-	-	-	-	(2763)		48.1	-
Alkaline condensate	-	-	2.145	53.6	2.145	53.6	1.000	-	-	-	145	4.212	32.8	-
Non-technological water	-	-	0.121	3.0	0.121	3.0	1.000	-	-	-	35	4.187	0.4	-
Out: Predesilicated slurry:														
liquid phase	1828.38	1729.74	15.042	376.1	11.341	283.5	1.326	161.2	152.5	1.74	90	3.537	119.7	109
solid phase	46.53	359.29	1.284	32.1	0.428	10.7	3.000	-	-	-	90	0.837	2.4	-
Alkaline condensate	-	-	2.842	71.1	2.842	71.1	1.000	-	-	-	108	4.191	32.2	-
Heat loss + heat of reaction	-	-	-	-	-	-	-	-	-	-	-	-	11.0	-
DIGESTION														
In: Predesilicated slurry	1874.91	2089.03	16.326	408.2	11.769	294.2	1.387	161.2	152.5	1.74	90	3.325	122.1	109
High-pressure steam	-	-	1.707	42.7	-	-	-	-	-	-	(3172.3)		135.5	-
Out: Flash effluent: liquid phase	1799.11	1909.39	12.368	309.2	8.531	213.3	1.450	210.9	223.8	1.55	118	3.363	122.7	125
solid phase	72.20	179.65	1.116	27.9	0.372	9.3	3.000	-	-	-	118	0.837	2.8	-
Flash steam for slurry preheating	-	-	0.697	17.4	-	-	-	-	-	-	(2763)		48.1	-
Pure condensate	-	-	1.707	42.7	1.707	42.7	1.000	-	-	-	228	4.293	41.8	-
Alkaline condensate	-	-	2.145	53.6	2.145	53.6	1.000	-	-	-	145	4.212	32.8	-
Heat loss + heat of reaction	-	-	-	-	-	-	-	-	-	-	-	-	9.4	-

Table 3.1-1/2

MATERIAL STREAM	Na2O	Al2O3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	Al2O3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
FLASHING OF PURE CONDENSATE														
In: Hihg-pressure pure condensate from digestion	-	-	1.707	42.7	1.707	42.7	1.000	-	-	-	228	4.293	41.8	-
Out: Low pressure steam	-	-	0.272	6.8	-	-	-	-	-	-	(2732)		18.6	-
Pure condensate	-	-	1.435	35.9	1.435	35.9	1.000	-	-	-	140	4.205	21.1	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.1	-
DILUTION														
In: Flash effluent: liquid phase	1799.11	1909.39	12.368	309.2	8.531	213.3	1.450	210.9	223.8	1.55	118	3.363	122.7	125
solid phase	72.20	179.65	1.116	27.9	0.372	9.3	3.000	-	-	-	118	0.837	2.8	-
1st washer overflow	382.42	316.62	7.756	193.9	7.025	175.6	1.104	54.4	45.1	1.99	80	3.940	61.1	-
Spent liquor	113.13	55.23	0.972	24.3	0.792	19.8	1.228	142.9	69.8	3.37	83	3.714	7.5	-
Out: Diluted slurry: liquid phase	2294.66	2281.24	21.096	527.4	16.348	409.6	1.290	140.4	139.5	1.66	99	3.591	187.9	67
solid phase	72.2	179.65	1.116	27.9	0.372	9.3	3.000	-	-	-	99	0.837	2.8	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	3.4	-
RED MUD SETTLING														
In: Diluted slurry	2366.86	2460.89	22.212	555.3	16.720	418.9	1.328	140.4	139.5	1.66	99	3.453	190.7	67
Out: Settler overflow	1946.39	1952.39	18.047	451.2	14.003	350.1	1.289	139.0	139.4	1.64	98	3.594	158.9	-
Settler underflow: liquid phase	342.93	327.00	3.042	76.1	2.345	58.6	1.297	146.2	139.4	1.73	98	3.583	26.7	413
solid phase	77.55	185.56	1.123	28.1	0.373	9.3	2.998	-	-	-	98	0.837	2.3	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.8	-
RED MUD WASHING														
In: Settler underflow:														
liquid phase	342.93	327.00	3.042	76.1	2.345	58.6	1.297	146.2	139.4	1.73	98	3.583	26.7	413
solid phase	77.55	185.56	1.123	28.1	0.373	9.3	2.998	-	-	-	98	0.837	2.3	-
Filtrate	56.68	-	4.214	105.3	4.150	103.7	1.015	13.7	-	-	75	4.151	32.8	-
Red mud wash water	-	-	2.352	58.8	2.352	58.8	1.000	-	-	-	90	4.187	22.2	-
Alkaline hydrate wash water	21.85	10.47	0.452	11.3	0.417	10.4	1.083	52.4	25.1	3.43	60	3.993	2.7	-
Non-technological water	-	-	0.300	7.5	0.300	7.5	1.000	-	-	-	35	4.187	1.1	-
Out: 1st washer overflow	382.42	316.62	7.756	193.9	7.025	175.6	1.104	54.4	45.1	1.99	80	3.940	61.1	-
4st washer underflow:														
liquid phase	39.04	3.06	2.560	64.0	2.513	62.8	1.019	15.5	1.2	20.98	78	4.143	20.7	402
solid phase	77.55	199.35	1.167	29.2	0.389	9.7	2.996	-	-	-	78	0.837	1.9	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	4.1	-

Table 3.1-1/3

MATERIAL STREAM	Na_2O	Al_2O_3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na_2O	Al_2O_3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
RED MUD CAUSTICIZATION AND FILTRATION														
In: 4th washer underflow:														
liquid phase	39.04	3.06	2.560	64.0	2.513	62.8	1.019	15.5	1.2	20.98	78	4.143	20.7	402
solid phase	77.55	199.35	1.167	29.2	0.389	9.7	2.996	-	-	-	78	0.837	1.9	-
Lime milk	-	-	0.821	20.5	0.715	17.9	1.148	-	-	-	75	3.738	5.7	200
Wash water	-	-	1.804	45.1	1.804	45.1	1.000	-	-	-	70	4.187	13.2	-
Low-pressure steam	-	-	0.087	2.2	-	-	-	-	-	-	(2792.6)		6.1	-
Out: Filtrate	56.68	-	4.214	105.3	4.150	103.7	1.015	13.7	-	-	75	4.151	32.8	-
Red mud to the pond:														
liquid phase	7.14	-	0.895	22.4	0.887	22.2	1.009	0.7	-	-	70	4.163	6.5	-
solid phase	52.76	202.41	1.330	33.3	0.444	11.1	2.996	-	-	-	70	0.837	2.0	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	6.3	-
MUD WASH WATER PREPARATION														
In: Digestion alkaline condensate	-	-	2.842	71.1	2.842	71.1	1.000	-	-	-	108	4.191	32.2	-
Evaporation alkaline condensate	-	-	1.620	40.5	1.620	40.5	1.000	-	-	-	75	4.187	12.7	-
Return water	-	-	0.246	6.2	0.246	6.2	1.000	-	-	-	35	4.187	0.9	-
Out: Red mud wash water	-	-	2.352	58.8	2.352	58.8	1.000	-	-	-	90	4.187	22.2	-
Red mud wash water to the filters	-	-	1.804	45.1	1.804	45.1	1.000	-	-	-	70	4.187	13.2	-
Heat loss + condensate loss	-	-	0.552	13.8	0.552	13.8	1.000	-	-	-	-	-	10.4	-
CONTROL FILTRATION														
In: Settler overflow	1946.39	1952.33	18.047	451.2	14.003	350.1	1.289	139.0	139.4	1.64	98	3.594	158.9	-
Out: Filtered aluminate liquor	1946.39	1952.33	18.047	451.2	14.003	350.1	1.289	139.0	139.4	1.64	96	3.594	155.7	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	3.2	-
HEAT INTERCHANGE														
In: Filtered aluminate liquor	1946.39	1952.33	18.047	451.2	14.003	350.1	1.289	139.0	139.4	1.64	96	3.594	155.7	-
Spent liquor	1888.40	921.94	16.222	405.6	13.212	330.3	1.228	142.9	69.8	3.37	46	3.714	69.3	-
Out: Cooled aluminate liquor	1946.39	1952.33	18.047	451.2	14.003	350.1	1.289	139.0	139.4	1.64	60	3.594	97.3	-
Spent liquor	1888.40	921.94	16.222	405.6	13.212	330.3	1.228	142.9	69.8	3.37	83	3.714	124.3	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	3.4	-

Table 3.1-1/4

MATERIAL STREAM	Na2O	Al2O3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	Al2O3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
PRECIPITATION														
In: Cooled aluminate liquor	1946.39	1952.33	18.047	451.2	14.003	350.1	1.289	139.0	139.4	1.64	60	3.594	97.3	-
Seed hydrate: liquid phase	227.21	108.90	1.925	48.1	1.564	39.1	1.230	145.2	69.6	3.43	46	3.708	8.2	331
solid phase	-	3904.65	5.974	149.4	2.469	61.7	2.420	-	-	-	46	0.837	5.8	-
Cooling water	-	-	14.716	367.9	14.716	367.9	1.000	-	-	-	35	4.187	53.9	-
Out: Precipitated slurry:														
liquid phase	2173.60	1058.64	18.438	461.0	14.975	374.4	1.230	145.2	69.6	3.43	48	3.708	82.1	414
solid phase	-	4898.95	7.495	187.4	3.098	77.5	2.420	-	-	-	48	0.837	7.5	-
Precipitator's scales	-	8.29	0.013	0.3	0.005	0.1	2.420	-	-	-	48	0.837	-	-
Cooling water	-	-	14.716	367.9	14.716	367.9	1.000	-	-	-	45	4.187	69.3	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	6.3	-
HYDRATE FILTRATION														
In: Precipitated slurry:														
liquid phase	2173.60	1058.64	18.438	461.0	14.975	374.4	1.230	145.2	69.6	3.43	48	3.708	82.1	414
solid phase	-	4898.95	7.495	187.4	3.098	77.5	2.420	-	-	-	48	0.837	7.5	-
Non-technological water	-	-	0.200	5.1	0.200	5.1	1.000	-	-	-	35	4.178	0.7	-
Out: Seed hydrate: liquid phase	227.21	108.90	1.925	48.1	1.564	39.1	1.230	145.2	69.6	3.43	46	3.708	8.2	-
solid phase	-	3904.65	5.974	149.4	1.469	61.7	2.420	-	-	-	46	0.837	5.8	-
Product hydrate: liquid phase	57.99	27.79	0.491	12.3	0.399	10.0	1.230	145.2	69.6	3.43	46	3.708	2.1	-
solid phase	-	994.30	1.521	38.0	0.629	15.7	2.420	-	-	-	46	0.837	1.5	-
Spent liquor	1888.40	921.94	16.222	405.6	13.212	330.3	1.228	142.9	69.8	3.37	46	3.714	69.3	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	3.4	-
PRODUCT HYDRATE WASHING														
In: Product hydrate: liquid phase	57.99	27.79	0.491	12.3	0.399	10.0	1.230	145.2	69.6	3.43	46	3.708	2.1	-
solid phase	-	994.30	1.521	38.0	0.629	15.7	2.420	-	-	-	46	0.837	1.5	-
Wash water	-	-	0.780	19.5	0.780	19.5	1.000	-	-	-	75	4.187	6.1	-
Low pressure steam	-	-	0.029	0.7	-	-	-	-	-	-	(2792.6)		2.0	-
Out: Product hydrate	-	994.30	1.521	38.0	0.629	15.7	2.420	-	-	-	55	0.837	1.7	-
Free water of product hydrate	4.70	2.25	0.199	5.0	0.191	4.8	1.050	24.5	11.8	3.43	55	4.082	1.1	-
Alkaline hydrate wash water	53.29	25.54	1.101	27.5	1.017	25.4	1.083	52.4	25.1	3.43	60	3.993	6.6	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.3	-

Table 3.1-1/5

MATERIAL STREAM	Na2O	Al2O3	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	Al2O3	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
SPENT LIQUOR EVAPORATION														
In: Spent liquor	754.39	368.35	6.481	162.0	5.278	132.0	1.228	142.9	69.8	3.37	83	3.714	49.6	-
Cooling water	-	-	33.890	847.3	33.890	847.3	1.000	-	-	-	35	4.187	124.2	-
Non-technological water	-	-	0.182	4.6	0.182	4.6	1.000	-	-	-	35	4.187	0.7	-
Low pressure steam	-	-	0.703	17.6	-	-	-	-	-	-	(2792.6)		49.1	-
Out: Strong liquor from spent liquor	754.39	368.35	4.343	108.6	3.140	78.5	1.383	240.2	117.3	3.37	70	3.480	26.5	-
Pure condensate	-	-	0.703	17.6	0.703	17.6	1.000	-	-	-	100	4.187	7.4	-
Alkaline condensate	-	-	1.740	43.5	1.740	43.5	1.000	-	-	-	75	4.187	13.7	-
Cooling water to tower	-	-	34.470	861.8	34.47	861.8	1.000	-	-	-	45	4.187	162.4	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	13.6	-
ALKALINE HYDRATE WASH WATER EVAPORATION														
In: Alkaline hydrate wash water	31.44	15.07	0.650	16.3	0.600	15.0	1.083	52.4	25.1	3.43	60	3.993	3.9	-
Spent liquor	112.63	54.99	0.968	24.2	0.788	19.7	1.228	142.9	69.8	3.37	83	3.714	7.5	-
Cooling water	-	-	12.800	320.0	12.800	320.0	1.000	-	-	-	35	4.187	46.9	-
Non-technological water	-	-	0.068	1.7	0.068	1.7	1.000	-	-	-	35	4.187	0.3	-
Low pressure steam	-	-	0.267	6.7	-	-	-	-	-	-	(2792.6)		18.7	-
Out: Strong from alk. hydrate wash water	144.07	70.06	0.806	20.2	0.576	14.4	1.399	250.0	121.6	3.38	70	3.460	4.9	-
Pure condensate	-	-	0.267	6.7	0.267	6.7	1.000	-	-	-	100	4.187	2.8	-
Alkaline condensate	-	-	0.660	16.5	0.660	16.5	1.000	-	-	-	75	4.187	5.2	-
Cooling water to tower	-	-	13.020	325.5	13.020	325.5	1.000	-	-	-	45	4.187	61.3	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	3.1	-
SALT REMOVAL FROM EVAPORATED HYDRATE WASH WATER														
In: Strong liquor	144.07	70.06	0.806	20.2	0.576	14.4	1.399	250.0	121.6	3.38	70	3.460	4.9	17
Low pressure steam	-	-	0.047	1.2	-	-	-	-	-	-	(2792.6)		3.4	-
Cooling water	-	-	2.204	55.1	2.204	55.1	1.000	-	-	-	33	4.187	7.6	-
Out: Wet salt	1.21	0.61	0.015	0.4	0.009	0.2	1.656	-	-	-	30	0.837	-	-
Salt free strong liquor to the digestion liquor preparation	140.84	71.38	0.791	19.8	0.567	14.2	1.395	248.4	125.9	3.25	70	3.460	4.8	-
Warmed cooling water	-	-	2.204	55.1	2.204	55.1	1.000	-	-	-	43	4.187	9.9	-
Pure condensate	-	-	0.047	1.2	0.047	1.2	1.000	-	-	-	140	4.187	0.7	-
Mechanical losses	2.02	-1.93	-	-	-	-	-	-	-	-	-	-	-	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	0.5	-

Table 3.1-1/6

MATERIAL STREAM	Na2O	A1203	MASS		VOLUME		DENS. AT 25 oC	CAUSTIC Na2O	A1203	MOLAR RATIO	TEMP.	SPEC. HEAT	HEAT CONT.	SOLIDS
	kg/t	kg/t	t/t	t/h	m3/t	m3/h	t/m3	gpl	gpl	-	oC	kJ/kgoC	GJ/h	gpl
DIGESTION LIQUOR ADJUSTMENT														
In: Strong liquor from evaporation	754.39	368.35	4.343	108.6	3.140	78.5	1.383	240.2	117.3	3.37	70	3.480	26.5	-
Strong liquor from salt removal	140.84	71.38	0.791	19.8	0.567	14.2	1.395	248.4	125.9	3.25	70	3.460	4.8	-
Spent liquor	908.25	443.37	7.801	195.0	6.354	158.8	1.228	142.9	69.8	3.37	83	3.714	59.7	-
Non-technological water	-	-	0.250	6.3	0.250	6.3	1.000	-	-	-	27	4.187	59.7	-
Make-up caustic + scale	71.42	8.29	0.253	6.3	0.164	4.1	1.543	435.5	50.3	14.24	50	3.333	1.0	-
Out: Digestion liquor	1874.91	891.39	13.437	335.9	10.471	261.8	1.283	179.1	85.1	3.46	72	3.625	87.8	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	5.1	-
LIME MILK PREPARATION														
In: Burnt lime (80 % active CaO)	-	-	0.216	5.4	0.065	1.6	3.350	-	-	-	27	0.837	0.1	-
Return water	-	-	0.792	19.8	0.792	19.8	1.000	-	-	-	35	4.187	2.9	-
Low-pressure steam	-	-	0.006	0.1	-	-	-	-	-	-	(2792.6)		0.4	-
Heat of reaction	-	-	-	-	-	-	-	-	-	-	-	-	5.9	-
Out: Lime milk to the causticization	-	-	0.821	20.5	0.715	17.9	1.148	-	-	-	75	3.738	5.7	200
Lime milk to the digestion	-	-	0.172	4.3	0.150	3.8	1.148	-	-	-	75	3.738	1.2	200
Reject	-	-	0.021	0.5	0.006	0.2	3.350	-	-	-	75	0.837	-	-
Heat loss	-	-	-	-	-	-	-	-	-	-	-	-	2.4	-

Table 3.1-2/1 (cont.)

CHARACTERISTIC TECHNOLOGICAL PARAMETER
Projected for 1989

	Units	
Alumina production	tpy	200,000
Alumina production	tph	25
Design operating factor	%	91.3
Bauxite quality:		
Al_2O_3	%	48.00
SiO_2	%	4.0
Fe_2O_3	%	16.2
TiO_2	%	8.2
L.o.I.	%	22.8
Free water	%	3.9
Digestion liquor		
temperature	oC	72
caustic Na_2O	gpl	179.1
molar ratio		3.46
Lime addition	%	1.2
Specific bauxite consumption (dry)	t/t	2.495
Predesilication		
temperature in the first tank	oC	100
temperature in the last tank	oC	90
Digestion temperature	oC	240
Final temperature of heat recuperation	oC	184
Temperature in the last flash tank	oC	118

Table 3.1-2/2 (cont.)

	Units	
Holding time at the digestion temperature	min	80
Final digestion caustic molar ratio		1.55
Digestion efficiency	%	85.0
Non-technological water from grinding to flashing	m3/t	0.121
Flashed water	m3/t	2.842
Underflow solids concentrations		
settlers	gpl	412
washers	gpl	402
Settler overflow		
temperature	oC	98
caustic Na2O	gpl	139.0
caustic molar ratio		1.64
Na2O/SiO2 ratio in the red mud		0.777
Non-technological water in the red mud washing	m3/t	0.3
Auto-precipitation	kg/t	19.70
Causticization efficiency	%	35.4
temperature	oC	95
min. holding time	hrs	4
Lime addition	kg/t	143
Burnt lime, active CaO	%	80
Lime milk concentration	gpl	200
Wash water temperature	oC	90
Filtrate to the 4th washer	oC	75
Aluminate liquor temperature		
before cooling	oC	96
after cooling	oC	60

Table 3.1-2/3 (cont.)

	Units	
Spent liquor temperature		
before heat interchange	oC	46
after heat interchange	oC	83
Carbonate level as Na_2O in carbonate/Na_2O caustic	%	11.1
Average holding time	hrs	73
Seed ratio		2
Mid-stage cooling	oC	8
Non-technological water from liquor cooling to hydrate filtration	m3/t	0.2
Temperature in the first precipitator tank	oC	58
in the last precipitator tank	oC	48
Precipitation efficiency	%	52.2
Free water content of wet seed	%	18
Free water content of washed hydrate	%	10.96
Hydrate wash water	m3/t	0.780
Alkaline hydrate wash water		
caustic Na_2O	gpl	52.4
for evaporation	m3/t	0.600
Hydrate in spent liquor	gpl	1.29
Precipitator's scale	kg/t	8.29
Non-technological water from evaporation to digestion liquor adjustment	m3/t	0.5
Strong liquor		
caustic Na_2O	gpl	241.7
temperature	oC	70
Removed salt	t/t	0.016 (max.0.040)

Table 3.1.1-1

CHEMICAL COMPOSITION

		Amarkantak Bauxite yearly Composite "1987"	Phutka-Pahar Bauxite yearly Composite "1987"	Last Washer 5.Under - flow Mud Composite "jan.1988"	Filtered Caus- ticized Mud Comp. "jan.1988"
L.O.I.	%	22.2	23.9	11.1	11.9
Al_2O_3	%	48.4	49.6	29.1	27.7
SiO_2	%	4.0	2.7	7.4	7.4
Fe_2O_3	%	16.4	15.0	29.1	26.8
TiO_2	%	8.7	7.6	13.9	13.2
CaO	%	-	-	1.7	7.2
MgO	%	-	-	0.1	0.3
Na_2O	%	-	-	6.1	5.1

analysed by ALUTERV-FKI

Table 3.1.1-2

MINERALOGICAL COMPOSITION

	Amarkantak Bauxite yearly Composite "1987"	Phutka-Pahar Bauxite yearly Composite "1987"	Last Washer 5.Under - flow Mud Composite "jan.1988"	Filtered Caus- ticized Mud Comp. "jan.1988"
Al_2O_3 %				
in gibbsite	30.2	35.6	4.2	2.8
boehmite	11.5	8.6	12.9	12.5
diaspore	2.1	2.1	3.7	2.8
kaolinite	3.4	2.3	-	-
goethite	1.0	1.0	1.2	1.0
hematite	0.2	0.2	0.4	0.4
sodalite	-	-	6.3	5.5
calcium aluminium silicate	-	-	0.4	2.7
total	48.4	49.7	29.1	27.7
SiO_2 %				
in kaolinite	4.0	2.7	-	-
sodalite	-	-	7.4	6.5
calcium aluminium silicate	-	-	-	0,9
total	4.0	2.7	7.4	7.4

Table 3.1.1-2 (cont.)

	Amarkantak Bauxite yearly Composite "1987"	Phutka-Pahar Bauxite yearly Composite "1987"	Last Washer 5.Under - flow Mud Composite "jan.1988"	Filtered Caus- ticized Mud Comp. "jan.1988"
Fe_2O_3				
in goethite	5.6	5.0	8.5	7.2
hematite	10.8	10.0	20.6	19.6
total	16.4	15.0	29.1	26.8
TiO_2 %				
in anatase	7.7	7.0	8.9	8.5
rutil	1.0	0.6	2.3	2.1
sodium titanates	-	-	1.3	-
calcium titanate	-	-	1.4	2.6
total	8.7	7.6	13.9	13.2
CaO %				
in calcium titanate			1.0	1.8
calcium aluminium silicate			0.7	4.4
calcite			-	1,0
total			1.7	7.2

3.1.1.1 Improvements in the setting of the molar ratio and reduction of the hydrolysis losses

For increasing the digestion yield the following technological measures are suggested:

- The molar ratio at the end of digestion must be set to 1.55.
- The lime addition in the digestion should be continued. It is necessary to decompose and transform the sodium titanates formed during the digestion. The sodium titanates, when precipitated on the surface of the most difficult-to-digest boehmite fraction, separate the surface of the latter from the digesting liquor. If lime is added, Ca-titanate will be formed having a looser structure, which does not hinder the digestion of the boehmite.
- The bauxite to digesting liquor and lime milk ratio controls have to be made more reliable.
- The final temperature of the digestion must be 240 ^{o}C. First of all the steam supply must be reliably assured. (The reasons of the frequent power plant failures must be determined and eliminated). The safe operation of the digestion pressure and temperature controls is also important.
- The digestion process mus be continuously checked by means of frequent chemical and mineralogical analyses of the liquor, the bauxite and the digester mud. The digesting

liquor and the liquid phase of the digested slurry must be analysed every hour. (The total and caustic Na_2O contents, the Na_2CO_3 content, the Al_2O_3 content may be measured reliably by the liquor analysing method developed by Aluterv-FKI, which method will be handed over to the Indian Party in the frame of this work. For the determination of the mineralogical composition of the bauxite and the mud the procurement of an X-ray diffractometer suitable for powder diffraction phase analysis is suggested, possibly in co-operation with UNIDO. A Philips made equipment is in operation in the Renukoot Plant of HINDALCO, its price is approximately 250,000 NLG. Aluterv-FKI's experts may cooperate in purchasing the suitable equipment and in training its operating personnel. The digestion of the bauxite may be checked by observing the quantity of undigested boehmite in the digested red mud.) In this case the digestion yield may be increased either by increasing the digestion molar ratio or by increasing the lime quantity added.

- The gibbsite appearing in the red mud of the first washer must also be analysed by the X-ray diffractometer. Gibbsite must not remain in the red mud when digested at 240 °C, its appearance in the settling-washing unit indicates a decomposition of the aluminate liquor, the so-called hydrolysis loss. The analysis is to be carried out at least once a day. Should the hydrolysis exceed a certain value (for example 1 % of the Al_2O_3 content of the processed bauxite; this value is presently 4 % in the Korba Alumina Plant), the molar ratio of the first washer should be increased (suitably by spent liquor addition).

By all of these measures, the digestion yield should be increased from the present 78-81 % to at least 85 %.

3.1.1.2 Partial bypassing of the digesters heated by flash steam with tube-in-tube type heat exchangers

During the investigation performed by Aluterv-FKI's experts in May, 1988 suggestions were made to replace the digestion H.P. heat exchangers with tube-in-tube type heat exchangers because of the high maintenance demand of the former ones.

The digestion unit was examined concerning the reduction of the high pressure steam consumption and the improvement of the conditions of the maintenance.

It is suggested that the recuperation heat exchangers and those digesters with recuperation function be replaced with tube-in-tube type heat exchangers.

The digesters bypassed in this way may be utilized for increasing the heating surface used for live steam and for increasing the holding time.

According to our calculations, if 50 m^2 heating surface is installed per recuperation stage, a 60 bar steam consumption of 21 t/h may be attained, per digestion line. This figure may slightly vary as a function of the heat transfer cofficient.

The installation of the new recuperation stages may be carried out gradually. To begin with, the originally suggested replacement of the H.P. heat exchangers with tube-in-tube type ones might be done. If the results are favourable, the second step may follow with the replacement of the three recuperation digesters as well. The replacement of the L.P. heat exhangers may be realized in the third step, the equipment needed for this is included in Variant B of our suggestion. In the flowsheet prepared (Figure 3.1.1.2-1) the whole implementation is shown.

At the live steam heated system the de-aeration of the digesters is expected to improve the heat transfer coefficient significantly.

The above mentioned solutions may be realized by the installation of the following equipment:

Variant A

	Number Tot.	 Operating	Price in RS
1. Heat exchangers 3x0 101.6x5.7/0324x9 L = 15.4 m F = 12.8 m^2	32	32	5,760,000
2. Pipes and valves			200,000
3. Steel structures and pipe supports			520,000

Variant B

		Number		Price
		Tot.	Operating	in RS
1.	Heat exchangers 3x∅ 101.6x5.7/∅324x9 $L = 15.4\ m$ $F = 12.8\ m^2$	56	56	10,080,000
2.	Pipes and valves			260,000
3.	Steel structures and pipe supports			780,000

3.1.2 Increasing the precipitation efficiency

An improvement of the precipitation efficiency may be attained as a result of several technological measures:

- reducing the amount of impurities in the process liquor,
- using the whole line of the precipitator tanks to provide an appropriate precipitation time, including the last tank presently used for caustic storage (so when cleaning one of them and using the last one as a filter feed tank, 14 of them would always be full and operating, this would require the setting up of a 2000 m^3 tank or two Nos. of 1000 m^3 each for storing caustic lye to be made of a DIN quality steel).

- utilizing an interstage cooling to provide an adequate temperature profile for the precipitation,

- increasing the quantity of seed added to the precipitation,

- classifying a part of the precipitated slurry (if the more active, finer part is utilized as seed).

3.1.2.1 Process liquor purification

During the data processing if became clear, that there is a significant reserve in the evaporation capacity, which may even be increased by the reduction of the quantity of non-technological water. It is an alumina plant experience, that a part of the impurities of the process liquor (oxalate, V_2O_5, etc.) is precipitated together with the hydrate during precipitation. (Their damaging effect is, that they block the most active locations of the seed from further crystallization, from the crystal growth and agglomeration.)

These impurities dissolve well in the hot (90-95 °C) water used for hydrate washing, in this way their concentration is proportionally higher in the hydrate wash water than in other process liquors. With the increase of the Na_2O concentration the solubility of the oxalate is sharply decreasing, therefore, by the concentration of the hydrate wash water through evaporation the oxalate may be removed from the process. The impurities contents of the Korba spent liquor and hydrate wash water were examined (see Table 3.1.2.1-1). As the analitycal results show, the sample was not representative, the V_2O_5 content of the spent liquor

Table 3.1.2.1-1

LIQUOR ANALYSES

		Alkaline hydrate wash water /collected on 29/9/88 /	Spent liquor /collected on 24/8/88 /
Na_2O_c	gpl	48.7	129.5
Al_2O_3	gpl	20.6	71.0
Na_2CO_3	gpl	11.7	23.9
C_{org}	gpl	3.43	6.17
-in $Na_2C_2O_4$	gpl	1.54	3.3
Molar ratio		3.88	3.00
P_2O_5	gpl	0.53	0.35
V_2O_5	gpl	0.64	0.53
F	gpl	0.90	0.90

analysed by ALUTERV-FKI

was 1.7–1.9 g/l in the studied period. The relative oxalate content of the hydrate wash water is 1.24 times and its relative V_2O_5 content is 3.2 times higher than those of the spent liquor (calculated for the same caustic Na_2O concentration).

One of the 22 t/h capacity evaporation units of the Korba Alumina Plant (there are two units with capacities of 22 t/h each and one with the capacity of 36 t/h) is suitable for the separate evaporation of the hydrate wash water. The organic C content of the hydrate wash water subjected to evaporation (concentrated to 250 gpl caustic Na_2O) would be 16.7 gpl, in which the oxalate content would amount to 7.9 gpl expressed as carbon, or 44.1 gpl as oxalate. The temperature of the process liquor leaving the evaporation is appr. 70 °C. At this temperature, the relative solubility of oxalate is about 1 % in a liquor of about 250 gpl caustic Na_2O concentration. As it has been discussed in Chapter 2, the P-content of the strong liquor decreases as a consequence of the lime addition in the digestion. The high V_2O_5 level of the strong liquor can be lowered by cooling of the liquor. Taking into consideration the local outdoor temperature, an average 35 °C of the cooled liquor temperature can be envisaged. The required average temperature of the cooling water is 33 °C. A tubular heat exchanger of 55 m^2 should be installed for this purpose. The cooled liquor is settled, the settler underflow is centrifuged. (see Fig. 3.1.2.1-1) By this way the level of contaminants can be lowered to the level in 1985-86 financial year in about 6 months.

As a starting measure we suggest to feed about 15 m^3/h of hydrate wash water into one of the 22 t/h evaporators together with about 20 m^3/h of spent liquor. This would reduce less the capacity of the evaporation and the rest of the hydrate wash water could for some time still be used in the red mud washing circuit. On the basis of practical experiences gained by this method it could be later decided, whether to proceed with the 100 % evaporation of the alkaline hydrate wash water or not. Alternatively the initial tests might also be done in the 36 t/h evaporator, because this is already connected to the existing salt settler and so no modifications would have to be carried out.

The above mentioned solutions may be realized by the installation of the following equipment:

		Number		Price
		Tot.	Operating	in RS
1.	Strong liquor tank V = 25 m^3	1	1	170,000
2.	Strong liquor pump	1	1	40,000
3.	Heat exchanger s = 55 m^2 p = 16 bar Tube bundle typ.	1	1	700,000
4.	Hydroseparator ∅ 6 m x 3 m	1	1	600,000

		Number		Price
		Tot.	Operating	in RS
5.	Slurry pumps $Q = 6\ m^3/h$ $H = 40\ m$	2	1	100,000
6.	Centrifuges $Q = 6\ m^3/h$ Solids content 230 g/l	2	1	250,000
7.	Steel structures and pipe supports			100,000
8.	Pipelines with fittings and valves The tanks belonging to item 6. are available			100,000

3.1.2.2 Improving the operation of the precipitator tanks

The experts of Aluterv-FKI experienced during their site mission, that the controlled pressure of the compressed air varied in the precipitation unit between 4.6 and 5.5 kg/cm^2 because of the fluctuating air consumption of the auxiliary consumers (calcination, Kelly filters, workshop).

According to the log sheets, complete black-outs are also frequent, lasting even longer than 5-10 minutes. Six tanks out of the 14 precipitators were out of operation. Two tanks had been in a clogged state for several months due to an earlier power failure.

The help of Aluterv-FKI's experts was asked by the personnel of the Korba Alumina Plant to describe the detailed method for restarting those tanks and for selecting a high pressure compressor aggregate.

To improve the operation of the precipitation unit, it is suggested that

- the pressure control loops at the compressed air supply of the precipitation and desilication be maintained and put into operation,
- if the pressure level is from time to time still lower than the required one, the consumption of the auxiliary consumers be reduced to such a degree, that a constant pressure air supply of the precipitation be assured,
- due to the frequent power failures a Diesel compressor aggregate should be installed (a high pressure unit),
- for the purpose of complete draining of the conical part of the tanks special drain valves be installed and the pipelines serving for caustic cleaning and for slurry extraction be separated,
- the newly developed wear-resistant air feeder (see attached drawing No.4.4-1) providing a more disperse air bubble distribution be used with the precipitators (presently having a specific power consumption of 26-29 kW per tank), ensuring an energy saving of at least 18-22 %,

- the slurry channel system of the precipitator line be updated by Balco. (At present the slurry flows in a channel mounted horizontally and its level and flow is controlled by means of a hole cut in the bottom of the channel. A separate pipe system with 600 mm diameter serves the purpose of bypassing the tanks and of stepping the slurry, provided with a separate air feeder.) The utilization of a single channel version is suggested which unites the two tasks favourably and the step-over of the material is performed by waste energy, using the departure loss of the air lift pump.

The majority of the above mentioned suggestions was discussed during the site misson by the experts of Aluterv-FKI and Balco.

3.1.2.3 Updating the hydrate filtration and washing

Aluterv-FKI's experts handed over a suggestion to Balco at the plant concerning the increase of the productivity of the seed filters. The hydrate removal from the surface of the filter may be more perfectly accomplished by feeding the compressed air pulse-wise, in this way the capacity of the seed filters would increase and the seed ratio of 2.0 could be kept.

The utilization of the DRYMAX 1938 additive is suggested for the reduction of the moisture content of washed hydrate according to Figure 3.1.2.3-1. By the use of the additive, the moisture content of the hydrate would decrease by some 3 %; so the fuel oil consumption of the calcination could be reduced by about 100 kg/h.

	Number		Price
	Tot.	Operating	in RS
1. Barrelpumps Q = 200 l/h H = 5 m	2	1	800
2. Pipe section with fitting and valves			350

3.1.2.4 Implementation of a hydroseparator and updating the interstage cooling

Experts of the Korba Alumina Plant were impressed during their visit to Hungary by the hydroseparator operating in the MOTIM Alumina Plant and by the Alfa-Laval made spiral heat exchangers used in the Almásfüzitő Alumina Plant for interstage cooling. The implementation of the hydroseparator is suggested according to Figure 3.1.2.4-1. Approximately 50 % of the precipitated slurry should be forwarded to the hydroseparator. The overflow would get to the seed filters after settling. The underflow of the hydroseparator would contain the product hydrate. The level of the compacted hydrate in the hydroseparator shouldbe indicated by a max-min level indicator.

The conceptual drawing and the main technological parameters of the hydroseparator are shown in Figure 3.1.2.4-3.

An interstage cooling system consisting of five serpentine heat exchangers is operating in the Korba Alumina Plant. It is suggested to replace them at the time their due change with spiral type ones having a good heat transfer coefficient, demanding insignificant cleaning and -according to Figure 3.1.2.4-2 much less area for location.

The above mentioned solutions may be realized by the installation of the following equipment:

<u>Variant 3.1.2.4-1</u>

		Number		Price
		Tot.	Operating	in RS
1.	Hydroseparator ø 4 m x 10 m	1	1	800,000
2.	Steel structures and pipe supports			180,000
3.	Pipe section with fittings and valves			50,000
4.	Tank with agitator ø 7x22 m $V = 1000\ m^3$	2	2	1,800,000

Variant 3.1.2.4-2

		Number		Price
		Tot.	Operating	in RS
1.	Interstage coolers (spiral heat exchangers) A = 100 m^2	3	3(2)	3,000,000
2.	Hydroseparator ϕ 4 x 10 m	1	1	800,000
3.	Steel structures and pipe supports			180,000
4.	Pipe section with fittings and valves			100,000
5.	Tank with agitator ϕ 7 x 22 m, V = 1000 m^3	2	2	1,800,000

3.1.2.5 Replacement of the existing P-17 plate heat exchangers

Presently four pcs of P-17 plate heat exchangers are operating in parallel and due to this parallel operation there is less velocity and consequent frequent jamming of them. Moreover, this design is outdated and it is very difficult and expensive to get replacement for the worn-out parts. It is suggested to replace those by the latest model AX-30 plate heat exchangers having four sets each with about 372 m^2 area operating in series (three if them operating and one spare). This would handle 500 m^3/h aluminate and spent liquors, rep. and drop the temperature of the former one from 100 to 58 °C. Because of this series operation the velocity across the plate would by higher and this would reduce scaling to a considerable extent.

	Number Tot.	 Operating	Price in RS
1. Heat exchangers A = 372 m^2, Typ:AX-30	4	3	9,600,000
2. Steel structures and pipe support			200,000
3. Pipe section with fittings and valves			800,000

3.1.3 Reducing the caustic consumption

The target mentioned in the title may be attained beside the improvement of the digestion efficiency (discussed in section 3.1.1) by increasing the effectiveness of the causticization and red mud washing. The efficiency of both operations is reduced by the high quantity of non-technological water getting into the process, and by the low evaporation level of the process resulting in only a low quantity of red mud wash water brought into the system.

3.1.3.1 Improving the causticization

According to the Contract, the possibility of the utilization of the complex causticization of the Korba Alumina Plant should be examined.

The basis of the method in question is the following:

- Lime is added in a quantity of about 3 moles CaO/mole bound Na_2O to a part of one of the washers' underflow and the resulting slurry is heated up to 95 °C. At this point the caustic Na_2O concentration of the liquid phase must not be higher then 40 gpl, that is why the underflow of that whasher stage should be treated which is best corresponding to this condition. The lime added takes part in the following chemical reactions, in sequence:

  - 1. it forms calcium aluminate with the Al_2O_3 content of the liquid phase,

  - 2. it causticizes the Na_2CO_3 content of the liquid phase,

  - 3. it forms calcium aluminium hydrosilicates by being incorporated into the sodium aluminium hidrosilicates of the red mud.

- The newly formed calcium aluminate may be decomposed by treating it with sodium carbonate salt in a later phase. The carbonate content of this salt becomes causiticized and approximately 40 % of the Al_2O_3 content bound into the calcium aluminate will be dissolved again.

- The temperature should be kept over 90 °C during the treatment. The required holding time is around 4 hours.

As it may be understood from the description, the method is only suitable where a significant quantity of sodium carbonate salt is separated from the process liquor. The quantity of separable sodium carbonate is small in the Korba Alumina Plant, as the CO_2 content of the bauxite is very low. The salt separated in Korba to balance the impurities level contains a considerable quantity of V_2O_5, P_2O_5 and oxalate, which, when causticized in the red mud washing line, would get into such equilibrium conditions, that most of its damaging components would get back into the process. Because of these reasons the utilization of the complex causticization is not suggested for the Korba Alumina Plant. Instead of it the optimization of the conditions of the traditional atmospheric causticization is suggested.

For the improvement of the efficiency of the causticization it is necessary that

- the temperature of the causticization be 95 °C (for the direct steam heating a temperature control is suggested),

- the lime addition be uniform (a ratio control is necessary between the last washer underflow and the lime milk flow-rate based on flow and solids content measurement; the Na_2O content of the red mud must be analysed regularly for determining the optimum lime addition),

- the Na_2O content in the liquid phase of the causticized slurry should not be higher than 20 gpl.

The caustic ization was calculated in two variants of lime addition, namely for 80 kg/t and 140 kg/t, respectively. In the first case 25 %, and in the second case 35 % of causticization efficiency might be achieved. For the improvement of the heat economy of the washing line it is practical to heat the filtrate which was cooled down on the filter up to 90 °C with the excess heat of the digestion pure condensate. The suggested modifications are shown in Figure 3.1.3.1-1. (The figures correspond to the 35 % variant.)

3.1.3.2 Reducing the quantity of non-technological water getting into the process

Approximately 3 m^3/t of non-technological water gets presently into the process, which amount should be reduced to about 1.5 m^3/t.

The method of reduction demands first of all technological measures, the racionalization and strict control of the operation.

The measures are as follows:

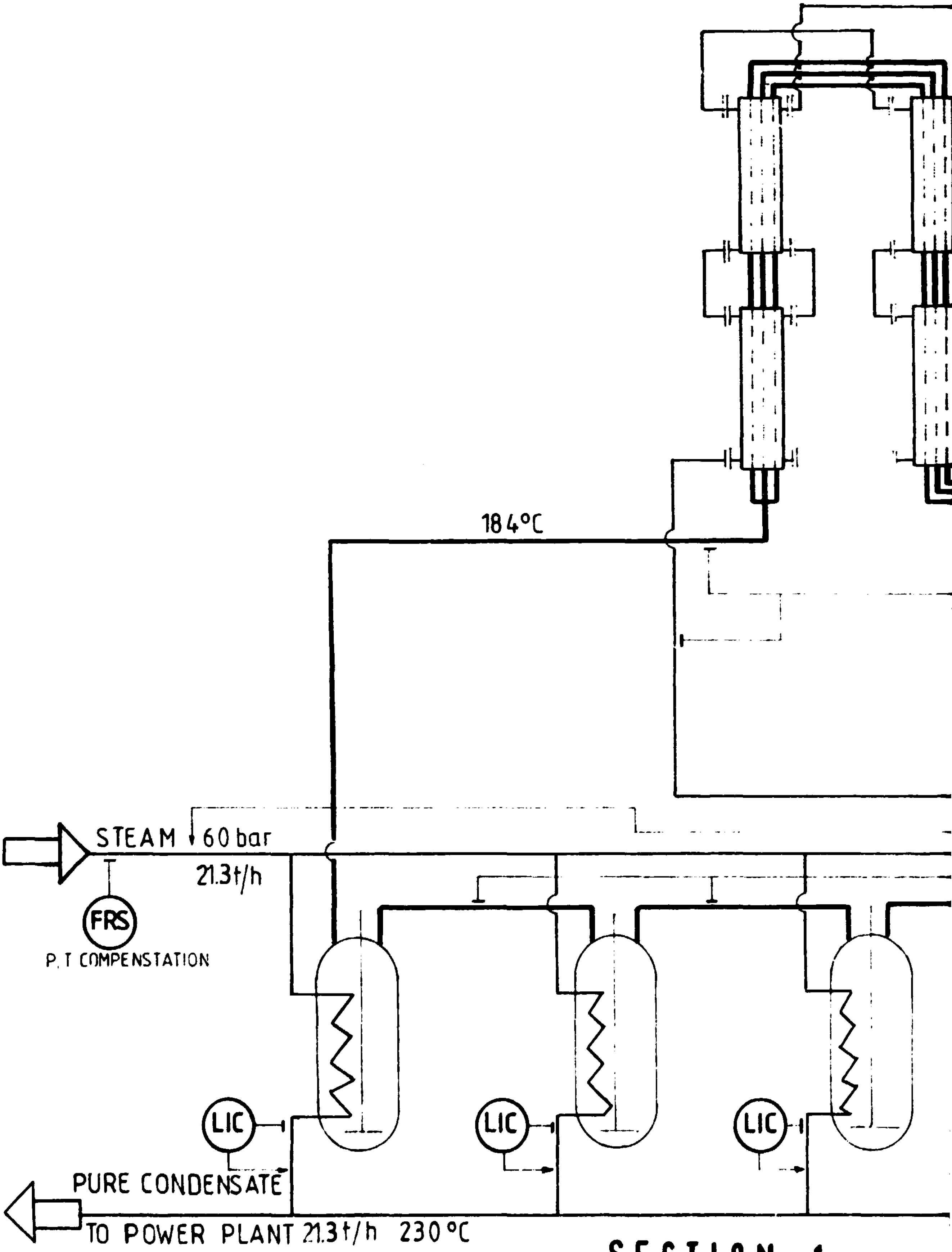
- adjustment and frequent checking of the gland water of agitators and pumps,

- economical utilization of floor washing water and its substitution with spent liquor or washer overflow, wherever possible,
- checking and elimination of leakages, installation of reliable level measurements and controls for the tanks wherever practical or level signalizations for tanks where the level is not controllable,
- feeding the washing liquors with different concentrations into appropriate tanks at the evaporation cleaning operations,
- reparation of the level controls of the flash tank line, regular checking of their operation.

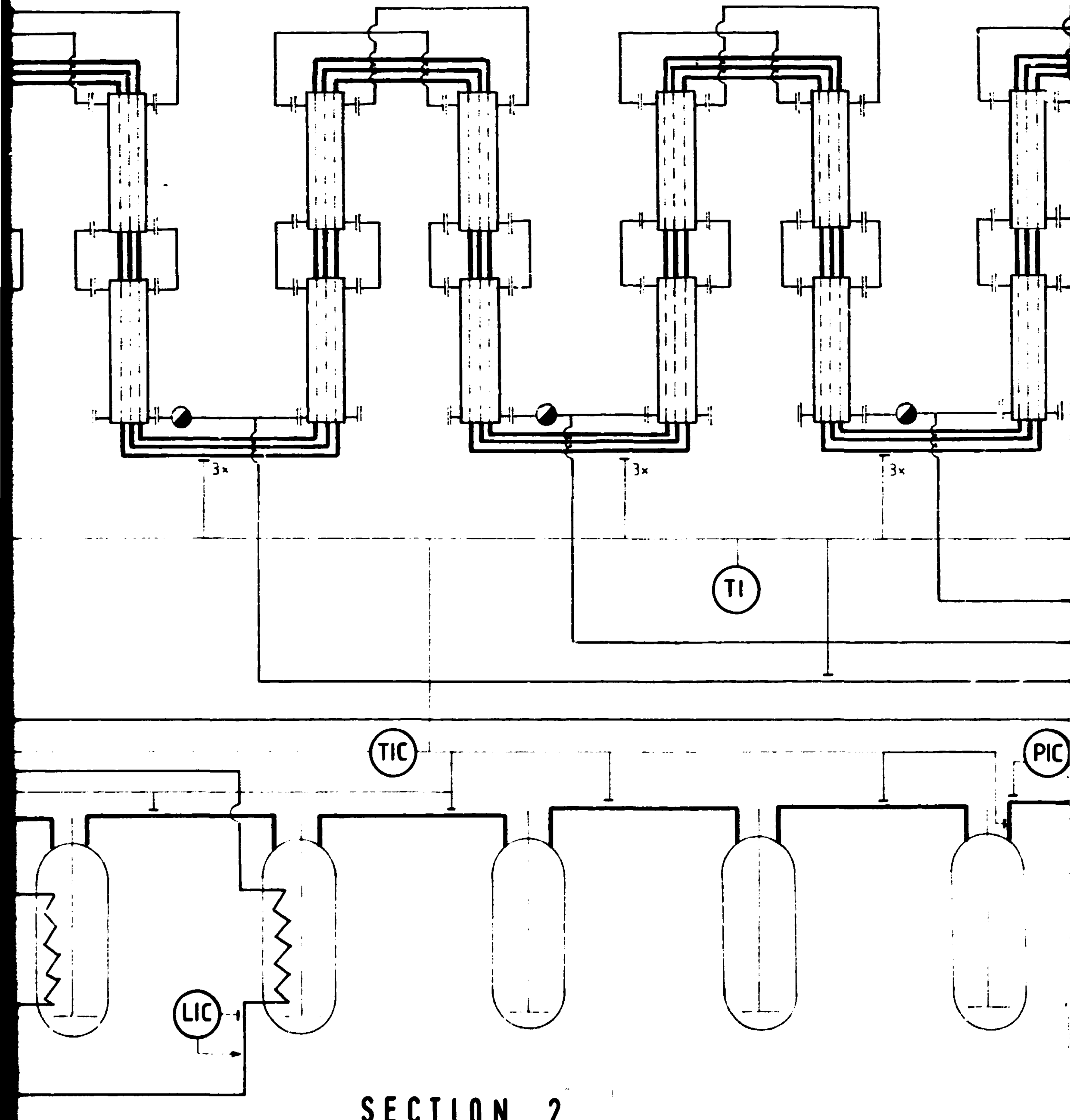
At the preparation of the mud washing water efforts should be made that its temperature be as high as possible.

First of all the red mud washing water should be composed of the alcaline condensate of digestion and of that part of evaporation condensate which is not utilized for hydrate washing.

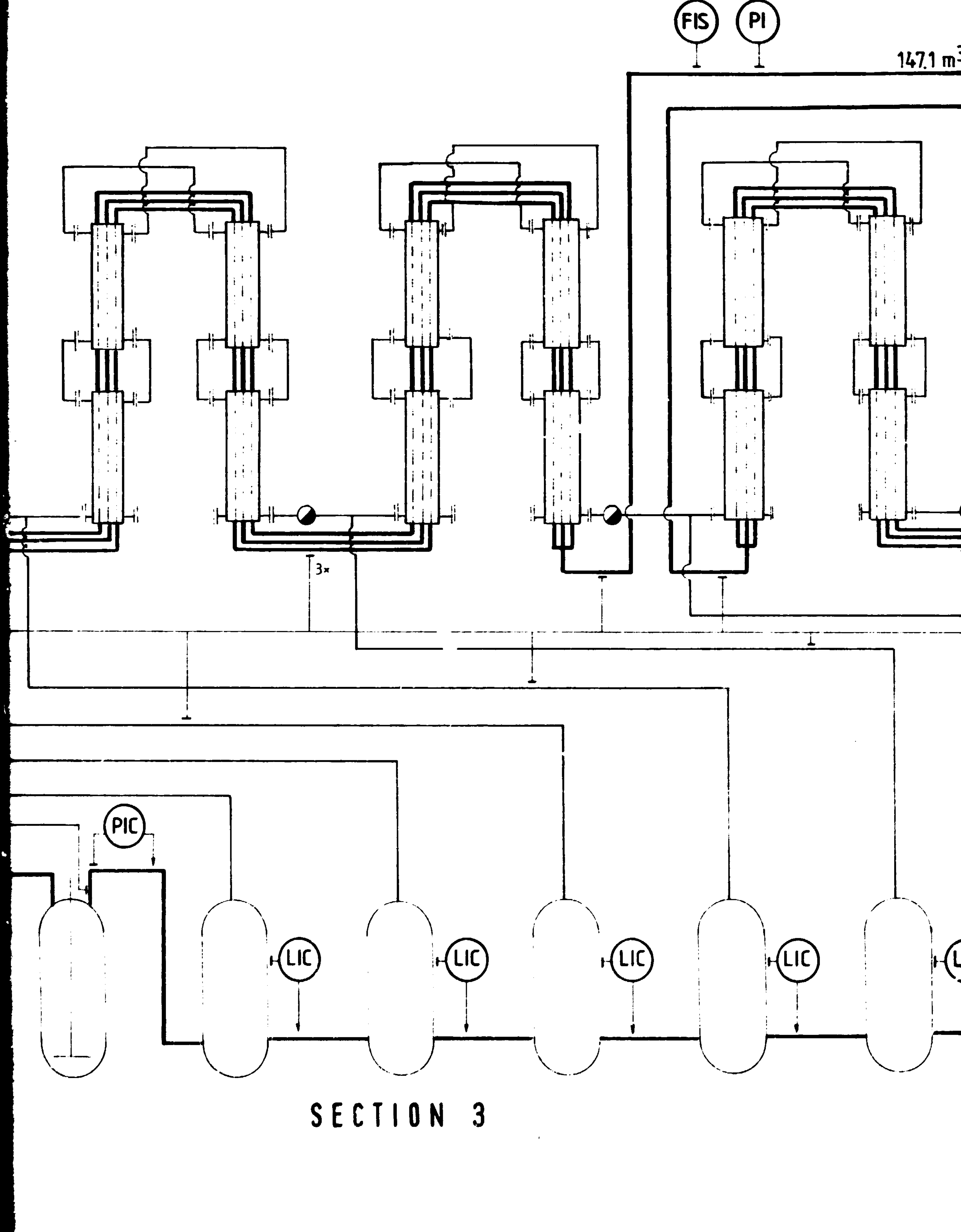
It is practical to use the return water from the red mud pond for lime slaking. This solution has the advantage that the carbonate brought in with the return water becomes causticized and does not increase the carbonate level of the process.


18 4°C
STEAM
60 bar
21.3t/h
FRS
P, T COMPENSTATION
LIC
LIC
LIC
PURE CONDENSATE
TO POWER PLANT 21.3t/h 230°C

SECTION 1

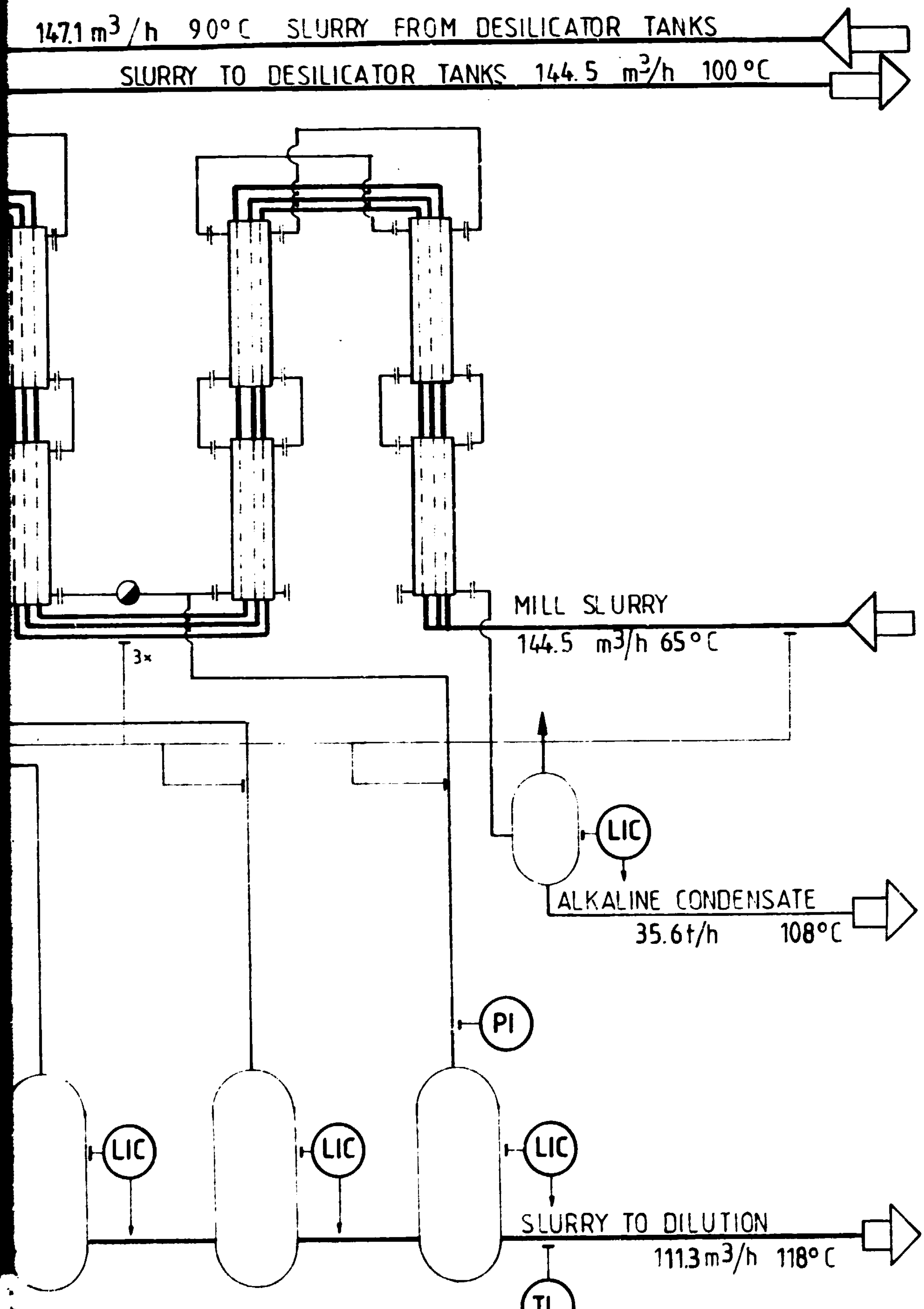


SECTION .2



SECTION 3

Fig. 3.1.1.2. -1



SECTION 4

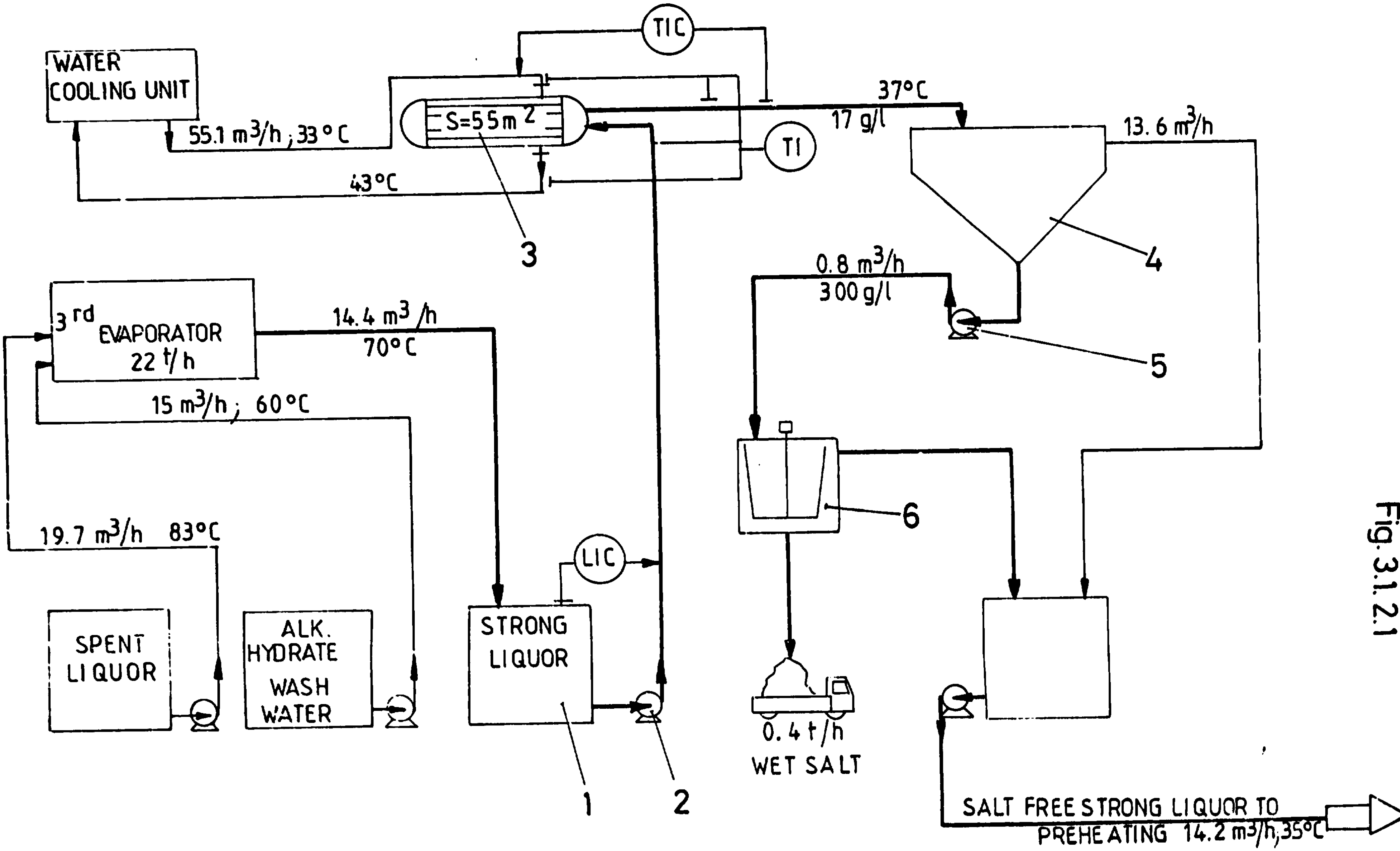

TIC
WATER COOLING UNIT
S=55m 2
55.1 m3/h ; 33 °C
43°C
TI
37°C
17 g/l
13. 6 m3/h
3
4
0.8 m3/h
300 g/l
5
3 rd EVAPORATOR 22 t/h
14.4 m3 /h
70°C
15 m3/h ; 60°C
19.7 m3/h 83°C
6
LIC
SPENT LIQUOR
ALK. HYDRATE WASH WATER
STRONG LIQUOR
0. 4 t/h
WET SALT
1
2
SALT FREE STRONG LIQUOR TO PREHEATING 14.2 m3/h, 35°C

Fig. 3.1.2.1

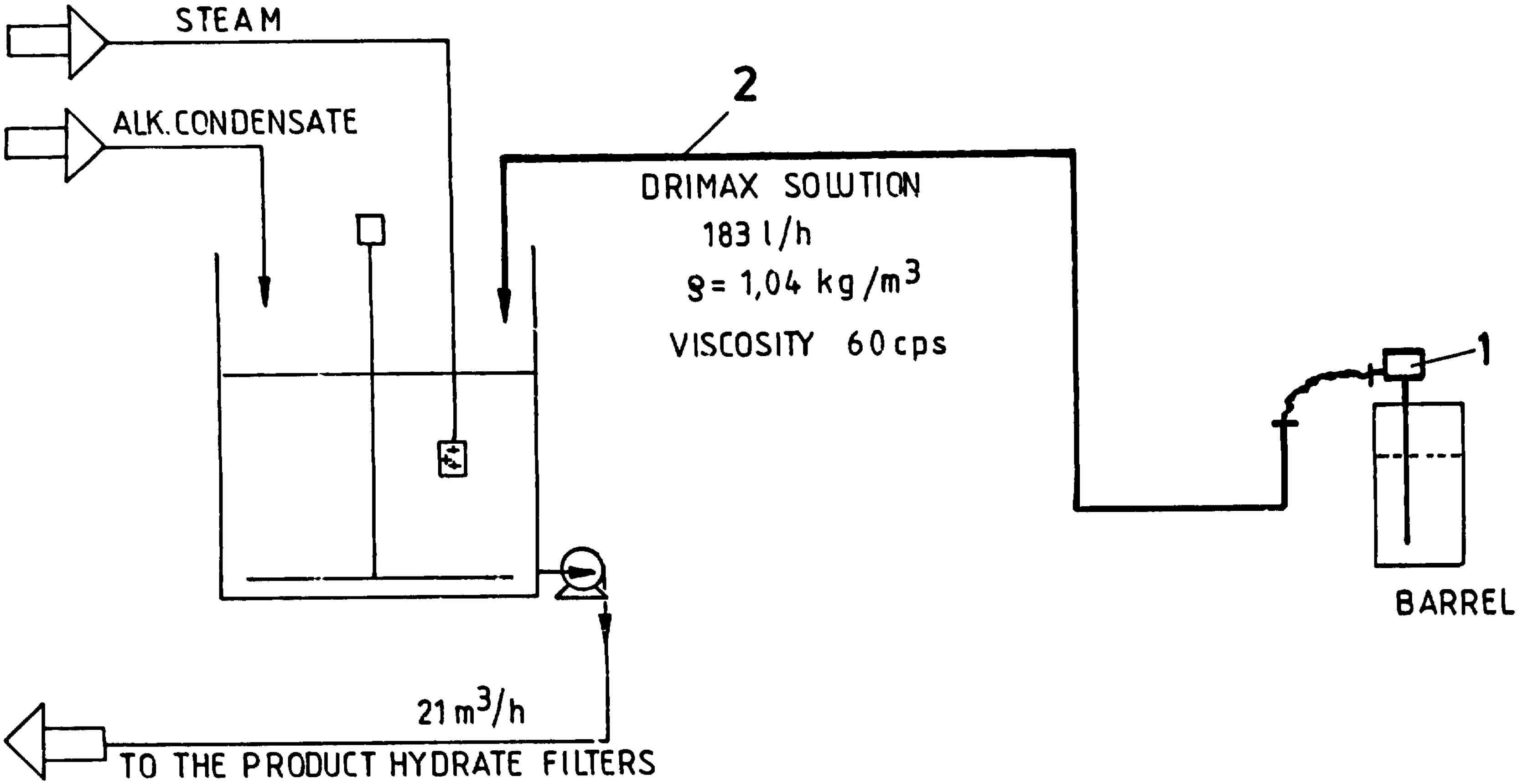
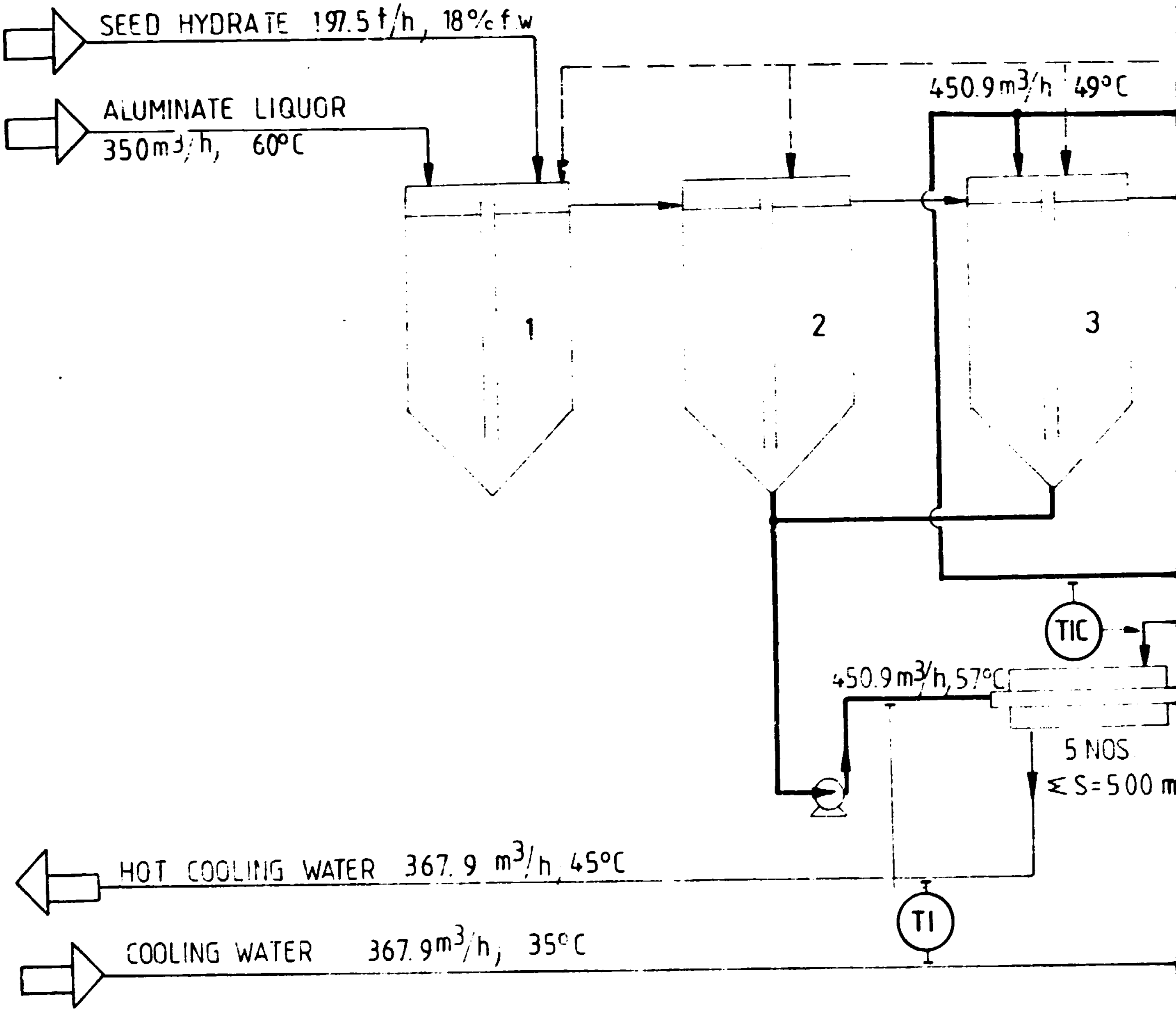

STEAM
ALK.CONDENSATE
2
DRIMAX SOLUTION
183 l/h
ϱ = 1,04 kg/m³
VISCOSITY 60 cps
1
BARREL
21 m³/h
TO THE PRODUCT HYDRATE FILTERS

Fig. 31.2.3-1


SEED HYDRATE 197.5 t/h, 18% f.w
ALUMINATE LIQUOR
350m3/h, 60°C
450.9m3/h 49°C
1
2
3
TIC
450.9m3/h, 57°C
5 NOS
Σ S=500 m
HOT COOLING WATER 367.9 m3/h, 45°C
TI
COOLING WATER 367.9m3/h, 35°C

SECTION 1

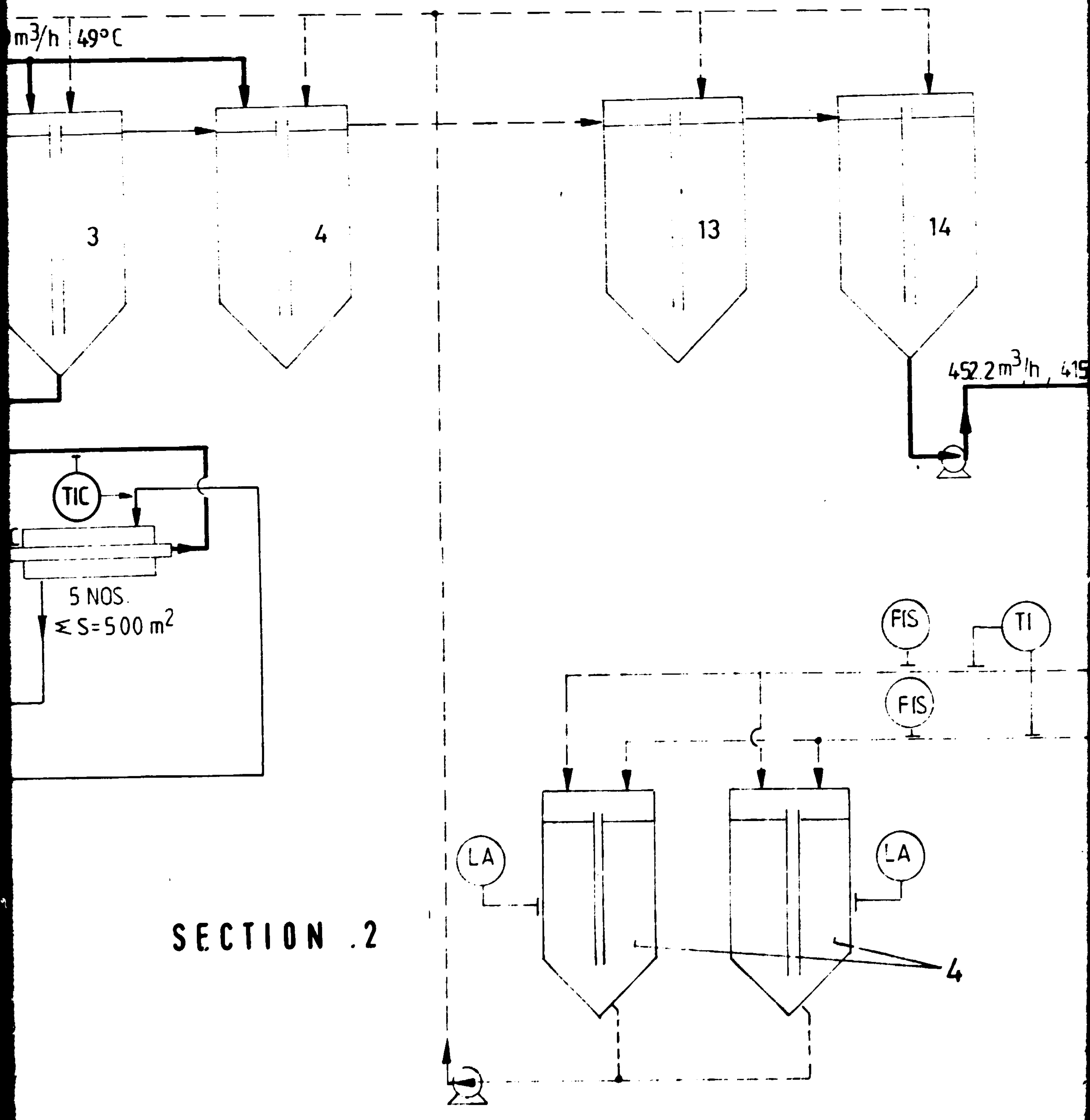
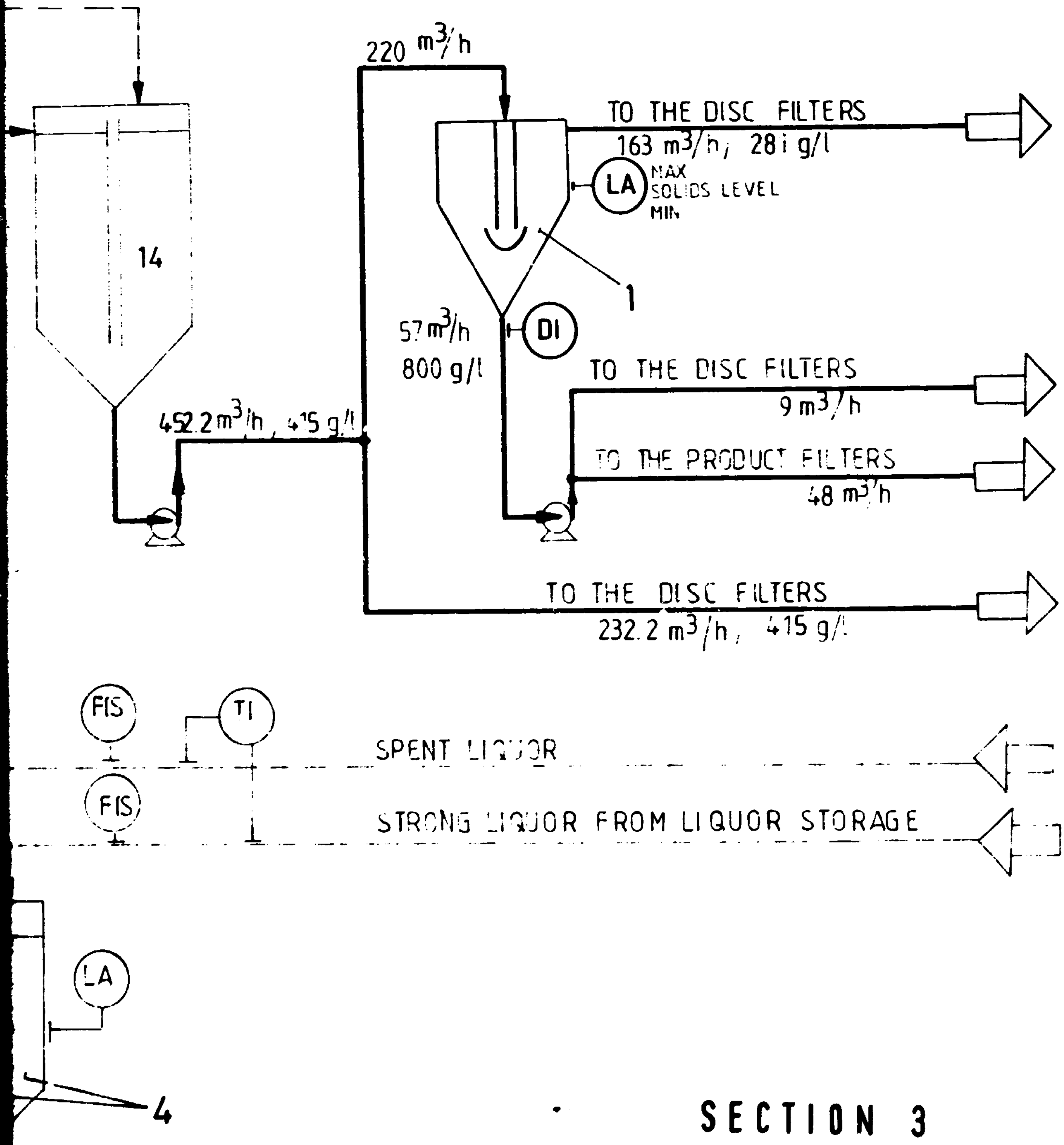
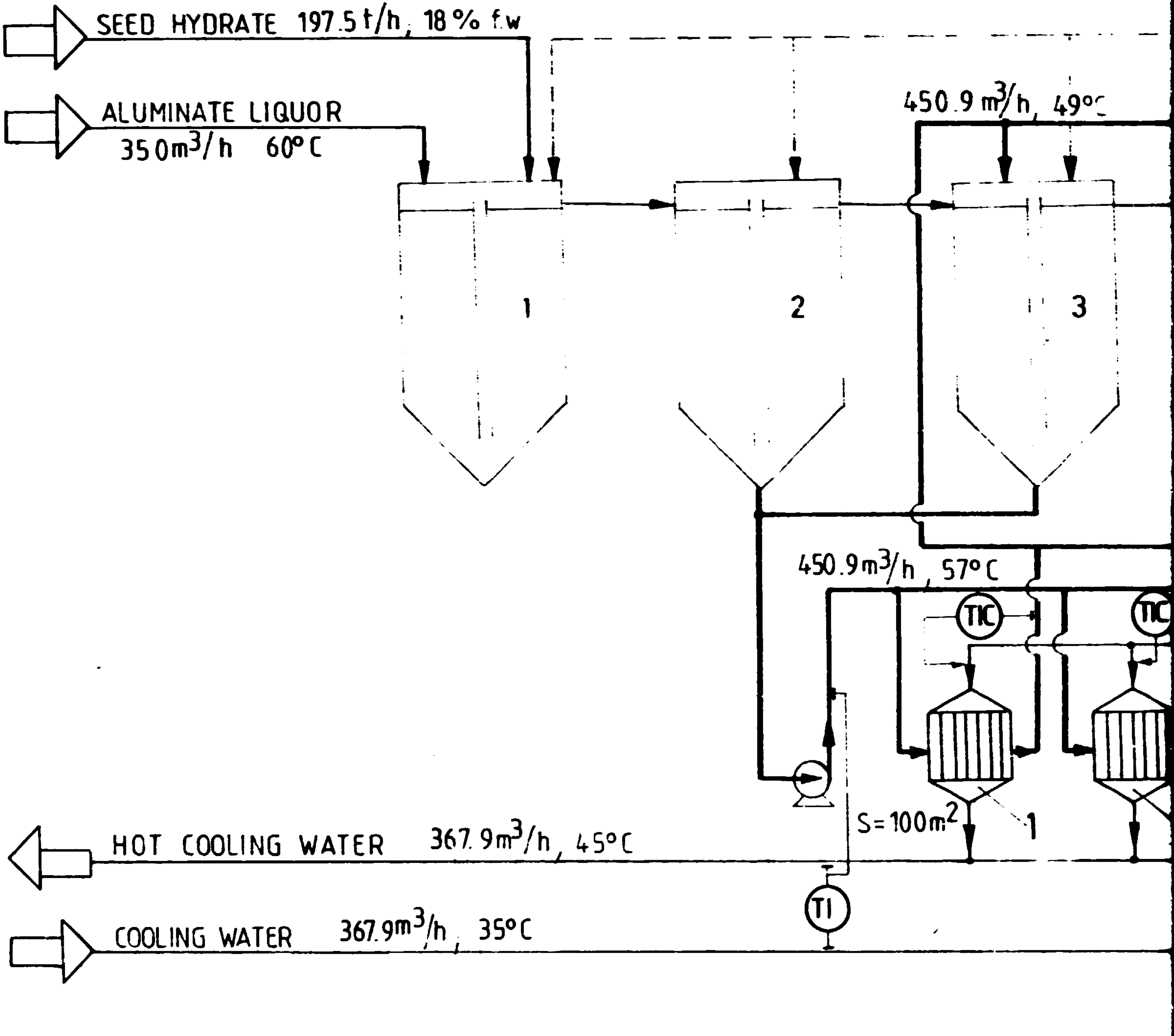
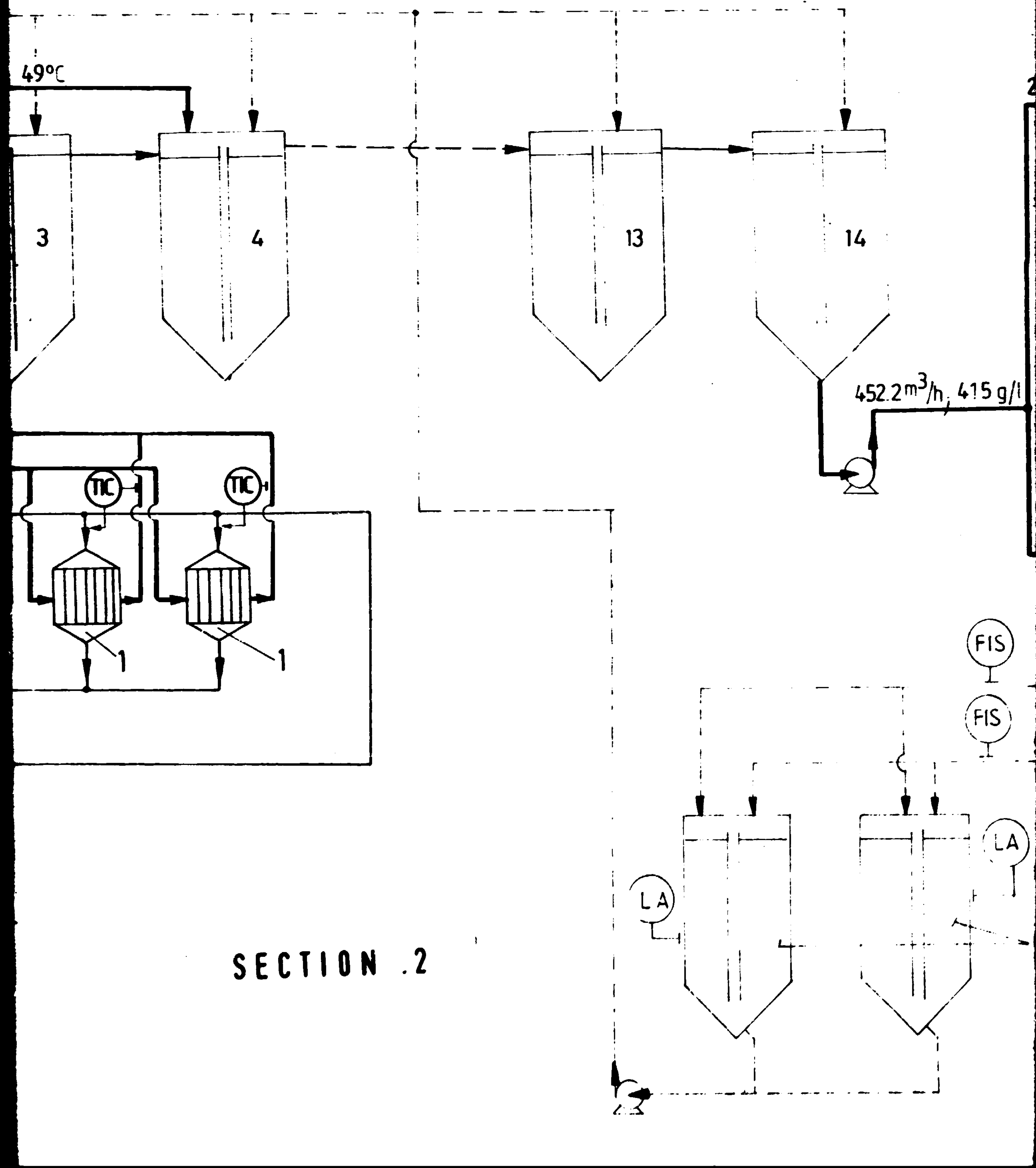

m3/h
49°C
3
4
13
14
452.2 m3/h , 415
TIC
5 NOS.
Σ S=500 m2
FIS
TI
FIS
LA
LA
4
SECTION .2

Fig. 3.1.2.4-1


220 m3/h
TO THE DISC FILTERS
163 m3/h, 281 g/l
LA
MAX
SOLIDS LEVEL
MIN
14
1
57 m3/h
800 g/l
DI
TO THE DISC FILTERS
9 m3/h
452.2 m3/h, 415 g/l
TO THE PRODUCT FILTERS
48 m3/h
TO THE DISC FILTERS
232.2 m3/h, 415 g/l
FIS
TI
SPENT LIQUOR
FIS
STRONG LIQUOR FROM LIQUOR STORAGE
LA
4
SECTION 3

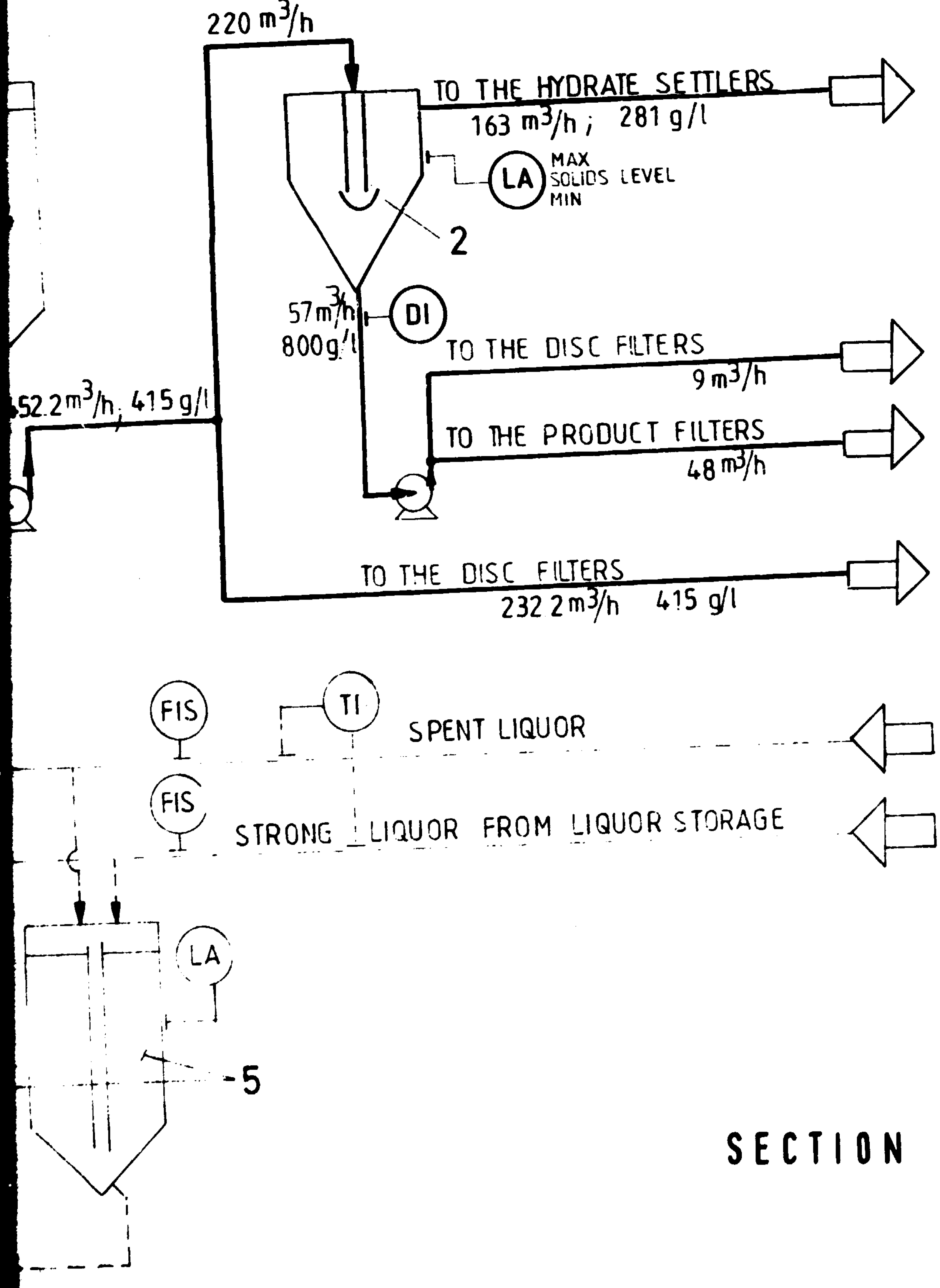

SEED HYDRATE 197.5 t/h, 18 % f.w
ALUMINATE LIQUOR
350 m3/h 60°C
450.9 m3/h, 49°C
1
2
3
450.9 m3/h, 57°C
TIC
TIC
S=100m2
1
HOT COOLING WATER 367.9 m3/h, 45°C
TI
COOLING WATER 367.9 m3/h, 35°C

SECTION 1



SECTION .2

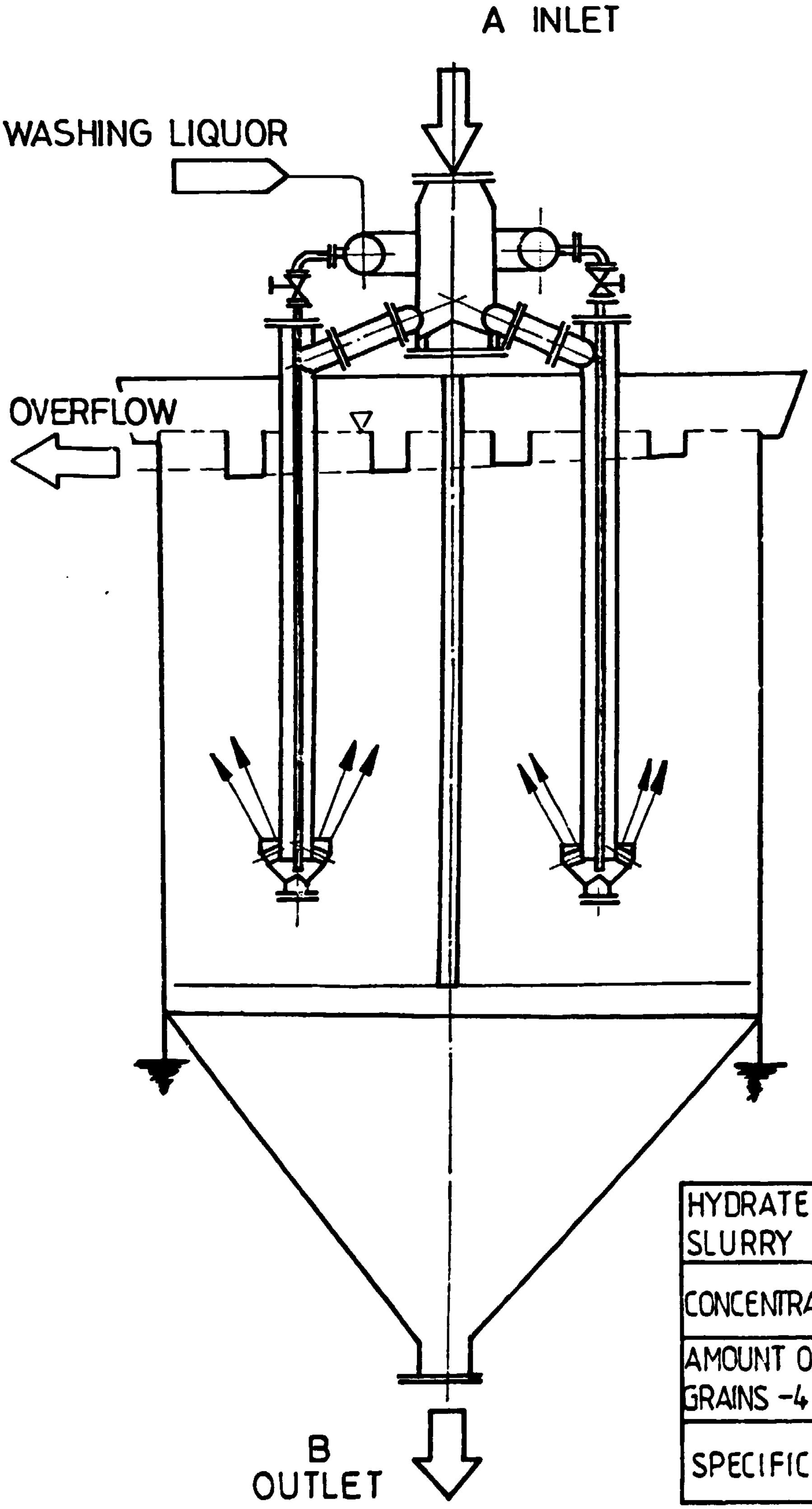
Fig.3.1.2.4-2



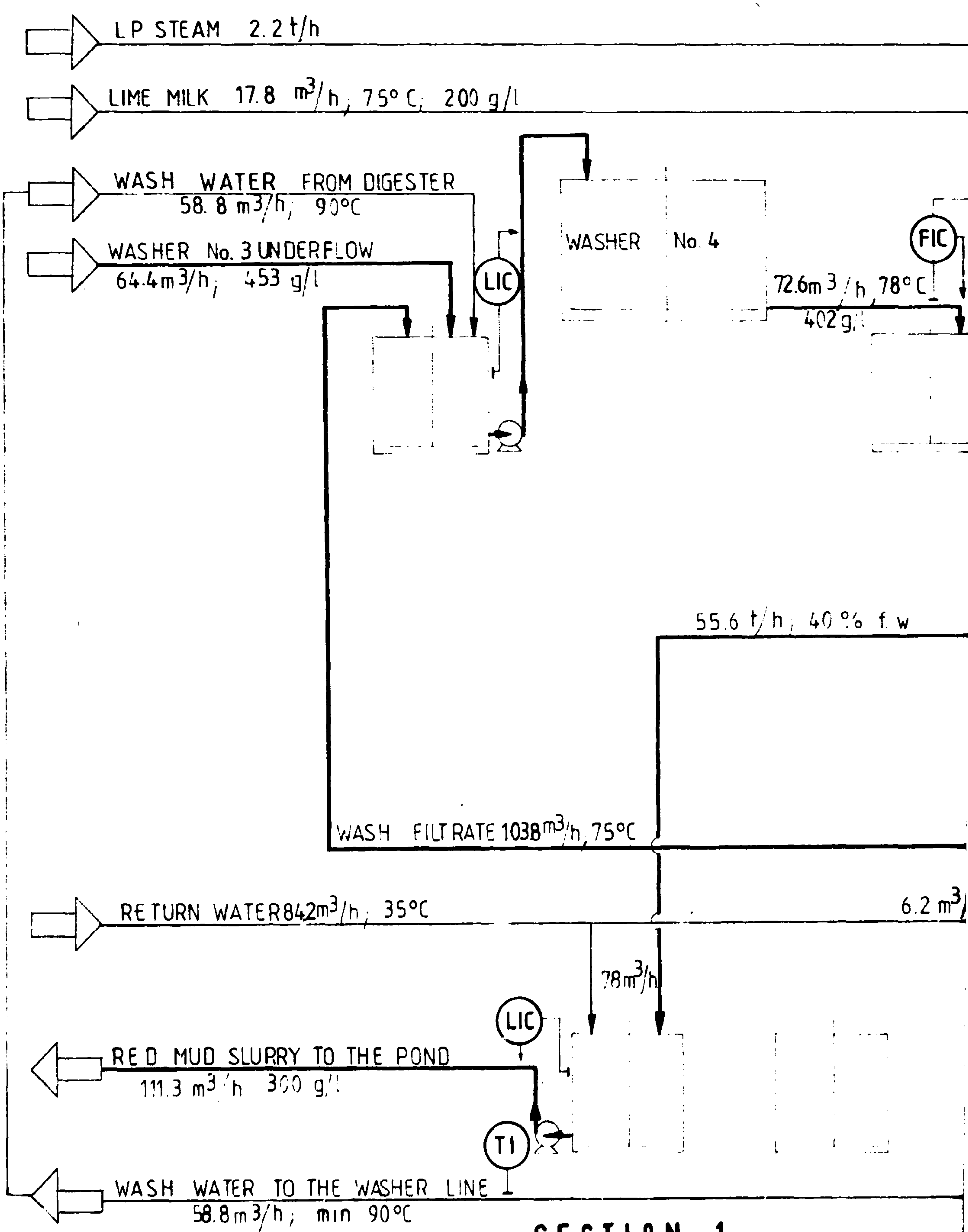
SECTION 3

Fig.3.1.2.4-3

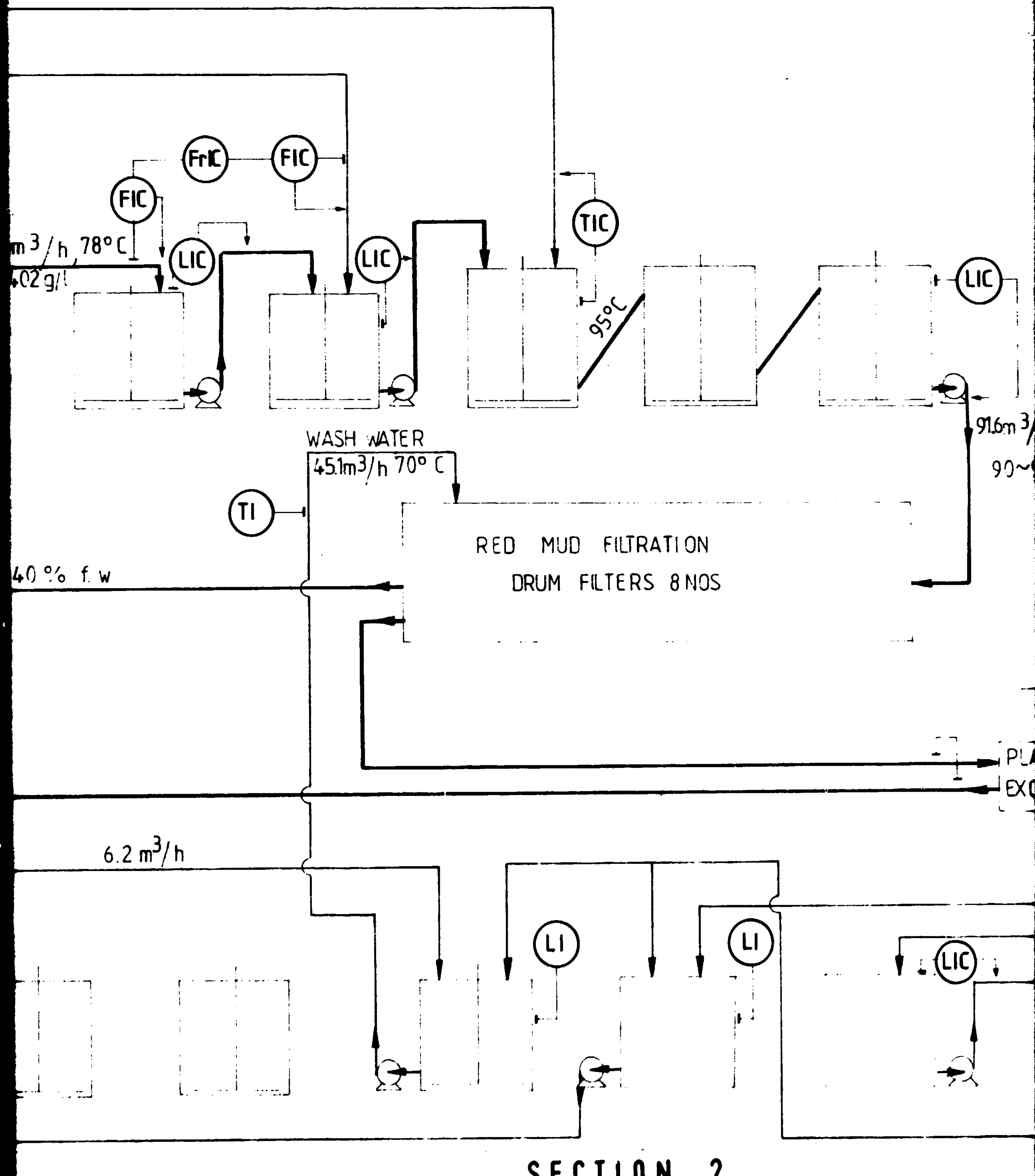
HYDRATE CLASSIFIER
MOTIM TYPE



		A	B
HYDRATE SLURRY	m^3/h	220	57
CONCENTRATION	g/l	400-600	>800
AMOUNT OF GRAINS -45µm	%	20-30	8-14
SPECIFIC LOAD	$m^3/m^2 h$	11-13	–


LP STEAM 2.2 t/h
LIME MILK 17.8 m3/h, 75° C; 200 g/l
WASH WATER FROM DIGESTER
58.8 m3/h; 90°C
WASHER No. 3 UNDERFLOW
64.4m3/h; 453 g/l
LIC
WASHER No. 4
FIC
72.6m 3/h, 78°C
402 g/l
55.6 t/h, 40 % f. w
WASH FILTRATE 1038 m3/h, 75°C
RETURN WATER 842 m3/h; 35°C
6.2 m3/
78 m3/h
LIC
RED MUD SLURRY TO THE POND
111.3 m3/h 300 g/l
TI
WASH WATER TO THE WASHER LINE
58.8 m3/h; min 90°C

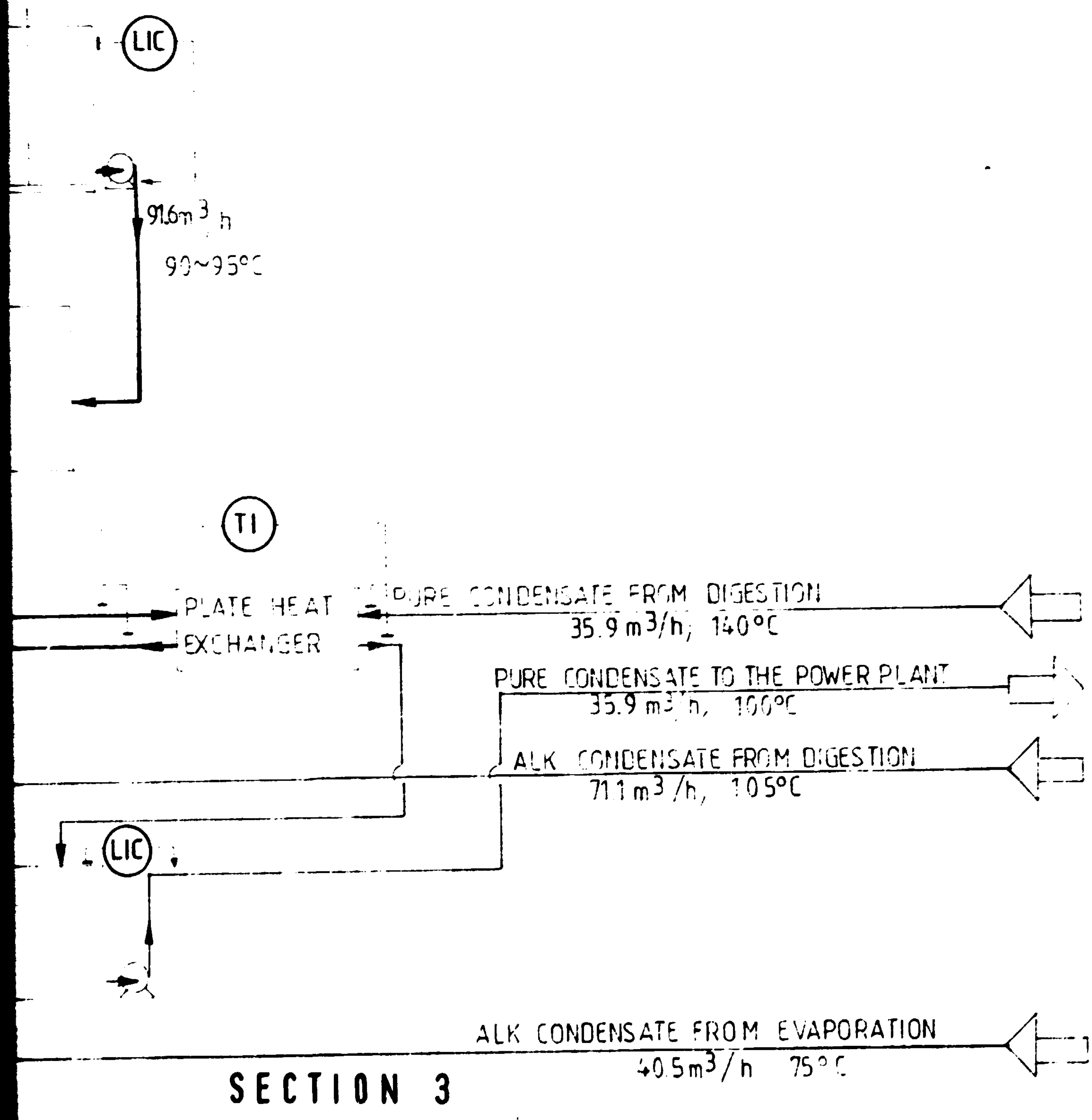
SECTION 1



m3/h, 78°C
402 g/l
FIC
FrIC
FIC
LIC
LIC
TIC
95°C
LIC
91.6m3/
90~
WASH WATER
45.1m3/h 70° C
TI
RED MUD FILTRATION
DRUM FILTERS 8 NOS
40 % f. w
PL
EX
6.2 m3/h
LI
LI
LIC

SECTION .2

Fig. 3.1.3.1-1



LIC
91.6 m3/h
90~95°C
TI
PLATE HEAT EXCHANGER
PURE CONDENSATE FROM DIGESTION
35.9 m3/h, 140°C
PURE CONDENSATE TO THE POWER PLANT
35.9 m3/h, 100°C
ALK CONDENSATE FROM DIGESTION
71.1 m3/h, 105°C
LIC
ALK CONDENSATE FROM EVAPORATION
40.5 m3/h 75°C
SECTION 3

3.1.4 New devices for calcination in alumina production

3.1.4.1 Conventional calciners

The task of the calciner is to remove the adhesive moisture and the crystal water of the aluminium hydroxide, which is carried out at a temperature of 1,000 to 1,200 °C depending on the alfa to gamma ratio of the alumina produced.

Nowadays rotary kilns are still in general use for calcining aluminium hydroxide into alumina where the hot alumina is cooled either in a planetary cooler rotating together with the kiln or in an independent rotary drum cooler.

The heat consumption of the conventional rotary kilns varies between 4,600 and 5,400 MJ for one tonne of alumina, depending on their capacity. This heat demand is met by burning fuel oil or natural gas.

Such kilns have significant surface heat losses, because the long rotary drums can not be properly insulated. The utilization of the heat content of the hot stack gases and of the hot alumina is also poor.

The alumina should be cooled from a temperature of 1,000 to 1,200 °C to an outlet temperature of 80 to 100 °C either in a planetary cooler or in a rotary drum cooler by means of the combustion air. Because of the poor heat transfer between air and alumina the planetary coolers are generally provided with gills while the rotary drum coolers are often sprayed with water to reach the required outlet temperature of the alumina. At these devices a significant part of the heat content of the alumina is not utilized for air preheating, but it is lost.

The stack gases have to be cooled to a temperature of 250 to 300 °C before the inlet of the electrostatic dust collector to make possible the operation of the latter. The gases are cooled by the hydrate fed into the kiln. The heat transfer between these two substances is also poor in spite of the various devices used to improve it, therefore, one part of the rotary kiln used for hydrate preheating and drying is not insulated in order to enable the stack gases to cool to the adequate outlet temperature by means of surface heat losses.

3.1.4.2 Reduction of heat consumption of calcination by cyclones

From the above facts it is evident, that the heat consumption of the calcination can be reduced by proper utilization of the heat content of the alumina and of the stack gases, as well as by a reduction of the length of the rotary kiln, since a long drum leads to significant surface heat losses because of the limited possibilities of its thermal insulation.

The technical developments for improving the utilization of the heat content of the alumina and of the stack gases, as well as for the reduction of the surface heat losses started almost simultaneously.

Cyclones were installed for a better utilization of the heat content of the stack gases. These cyclones took over the tasks of preheating and of drying from the rotary kilns. The cyclones insulated both inside and outside have improved the utilization of the heat content of the stack gases. The hydrate is preheated to a temperature of about 500 °C in them, a significant part of its adhesive moisture and crystal water is removed, while the stack gases can be cooled to a temperature of 140 to 180 °C depending on the fuel characteristics. The heat consumption can be reduced by 500 to 700 MJ per tonne of alumina by the cyclones installed for the utilization of the heat content of the stack gases. That means that the specific heat consumption of a rotary kiln provided with a cyclone type preheating system is 3,900 to 4,500 MJ per tonne of alumina, depending on the capacity of the calciner.

Such calciners are in operation in one of the Hungarian alumina plants with capacities of 500 to 550 MTPD. The heat consumption of these calciners is 3,890 MJ per tonne of alumina in case of the optimal load, while the yearly average amounts to 4,020 MJ per tonne of alumina.

The heat content of the alumina can also be utilized in cyclones. These cyclones are usually connected to a fluid bed water cooler to ensure the optimal preheating of the combustion air. A good heat transfer can be achieved between the alumina and the air in cyclones provided with proper inner and outer insulation. This results in an outlet temperature of the alumina of 200 to 250 °C, while the combustion air is heated to a temperature of 620 to 670 °C. The thermal energy saving is about 500 to 570 MJ per tonne of alumina when using cyclones as alumina coolers.

There are calciners in operation in another Hungarian alumina plant provided with cyclones for alumina cooling. Their capacity is similar to that of the previously mentioned calciners, whereas their average net specific consumption is 4,000 MJ per tonne of alumina.

Installing cyclone systems to both ends of a rotary kiln, a thermal energy saving of 1,000 to 1,270 MJ per tonne of alumina can be achieved, which means that the hydrate can be calcined with a heat consumption of 3,400 to 3,500 MJ per tonne in a rotary kiln with a capacity of 500 TPD or above that.

The above data are valid, if the hydrate has 10 to 12 per cent adhesive moisture and the calcination temperature is 1,100 °C.

The alumina leaves the cyclone cooler system at a temperature of 200 to 250 °C. It is further cooled in a fluid bed water cooler to a temperature of 80 to 100 °C. The amount of the termal energy recovered by the fluid bed water cooler is not included in the figures given above. The hot water obtained from the fluid bed water cooler can be utilized in the alumina production process, which results in additional savings in the heat consumption.

The specific heat consumption of the calciners is considerably influenced - beside the characteristics of the alumina and the type of the calciner - by the inner and outer insulations. The body of the kiln has to be insulated so that the temperature at its outer surface should not exceed 250 °C and the temperature of most of the kiln surface should be below 200 °C, whereas the surface temperature of the cyclones provided with inner and outer insulation should not exceed 60 °C and be 40 °C as an average.

3.1.4.3 Details of the solutions

Utilization of the heat content of the stack gases

The first rotary kiln provided with a cyclone system for the utilization of the heat content of the stack gases was constructed in 1965 in the frame of the reconstruction and expansion of a Hungarian alumina plant. The technical designs were prepared by Polysius (FRG) on the basis of their experiences gained in the cement industry. The so-called Dopol tower system designed by them is shown in Fig. 3.1.4.3-1.

As the flow-sheet shows, the mixture of the aluminium hydroxide and the collected dust is fed into the outlet stack gas pipe of Cyclone II. and heated to a temperature of 150 to

170 °C. After that it is precipitated in Cyclones Fig. 3.1.4.3-1. The hydrate dried in the cyclone system is fed into the rotary kiln at a temperature of 500 °C, while the temperature of the stack gases is 140 to 180 °C after the Cyclones I. before entering the electrostatic dust collector.

Some improvements have been carried out on the cyclone system since it was first commissioned and these have resulted in a troublefree, safe and well controllable operation of the cyclone system.

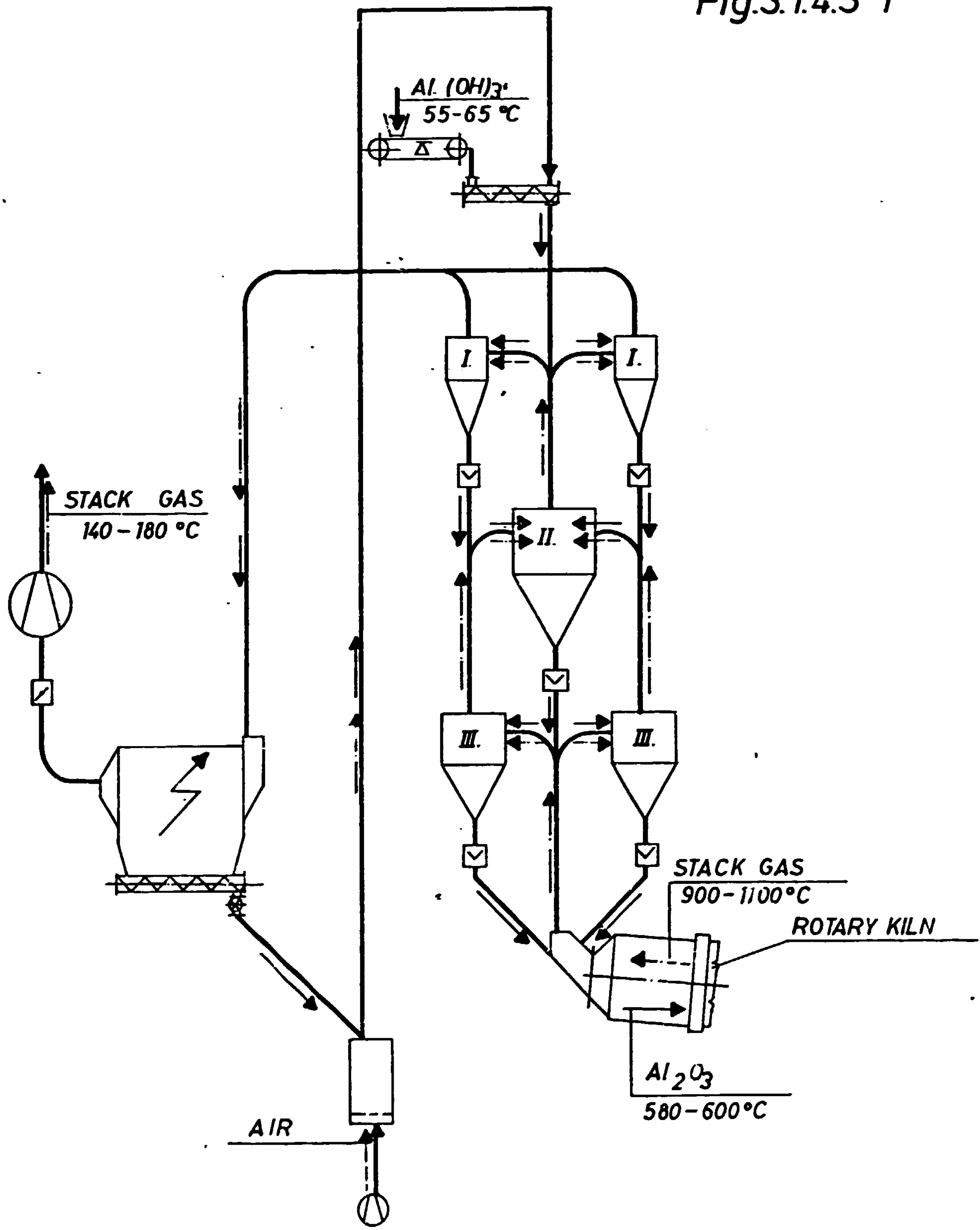
As a practical solution the product hydrate filters are installed at the top of the cyclone tower, but it is also possible to charge the filtered hydrate to the feed pipes of Cyclones I. with a bucket elevator or some other transportation system.

Utilization of the heat content of the alumina

Rotary kilns provided with cyclone systems for the utilization of the heat content of the alumina were constructed in 1972 in another Hungarian alumina plant. The final system developed after serveral modifications is shown in Fig. 3.1.4.3-2.

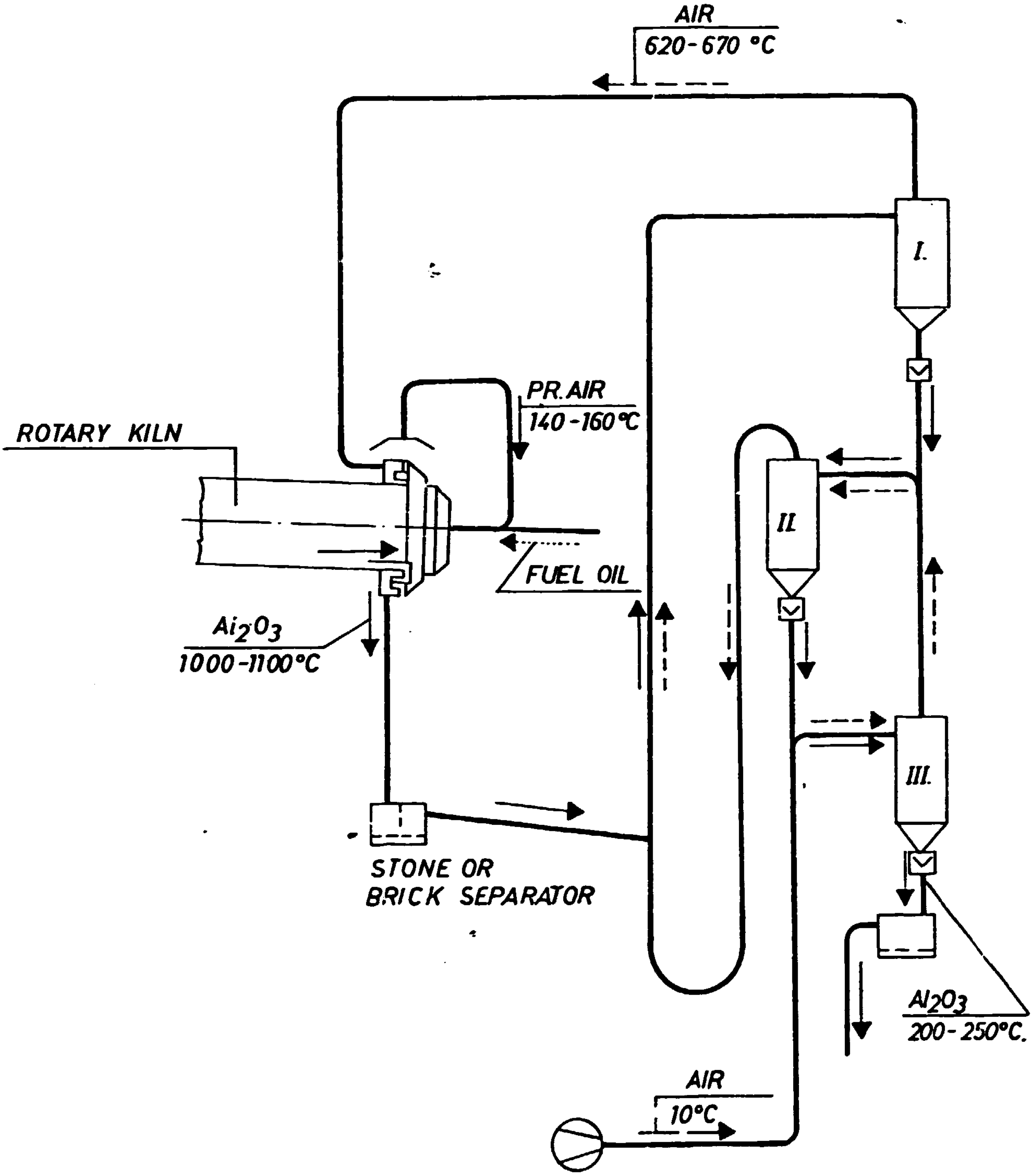
The calcined alumina discharged from the kiln is fed to the outlet combustion air pipe of Cyclone II. - usually through a so-called stone or brick separator - where its heat content is transferred to the combustion air and it is precipitated in Cyclone I. The air leaving Cyclone I. at a temperature of 620 to 670 °C is used for the firing.

Fig.3.1.4.3-1



CYCLONE PREHEATER SYSTEM

Fig.3.1.4.3-2



CYCLONE COOLING SYSTEM

After the counter-current heat exchange between the alumina and the combustion air the alumina leaves Cyclone III. at a temperature of 200 to 250 °C. It is usually further cooled by a fluid bed water cooler.

The system presented in this figure operates trouble-free even with a floury type alumina.

Rotary kiln

Significant reduction in the length of the rotary kiln can be achieved by using cyclone systems, because in this case the cyclone systems take over the drying, preheating and partially also the calcining functions of the kiln. As an example the length of a 100 m long rotary kiln can be reduced to 50 m by the installation of cyclone systems at both ends of the kiln, which means that the surface heat losses can be reduced by 40 to 50 per cent (taking into consideration the surface heat losses of a planetary or a rotary drum cooler).

Determination of the optimum length of a rotary kiln is an engineering job in case of a new kiln design, whereas the extent of the shortening of an operating kiln to be modified has to be determined on the basis of its original size and the location of its bearing rollers and driving unit.

An informatory heat balance of a 1,000 MTPD rotary kiln installed with preheater and cooler cyclone systems is the following:

Heat Losses		
	Stack gas	649 MJ/t
	Radiation, convection	544 MJ/t
	Discharge product	198 MJ/t
	Reaction heat	1,965 MJ/t
	Unaccounted	93 MJ/t
	Total:	3,449 MJ/t
Heat Input		
	Fuel combustion	3,370 MJ/t
	Hydrate	79 MJ/t
	Total:	3,449 MJ/t

3.1.4.4 Stationary calciners

Our cyclone systems described above can be used for the modification of conventional rotary kilns, primarily of alumina calciners, possibly of cement calciners, but these systems can be used at the designing of a new calciner, too.

Further reduction of the heat consumption can be achieved by the complete elimination of the rotary kiln. This can be achieved by using stationary (fluid bed, flash, gas suspension) calcination systems, the heat consumption of which is less by a further 300 to 500 MJ per tonne of alumina than that of our system suggested here, as a result of further possibilities in the reduction of the surface heat losses.

A disadvantage of stationary calciner systems can be the higher attrition of the alumina, which is particularly detrimental in the production of sandy alumina.

3.1.4.5 Advantages of the Aluterv-FKI cyclone system calciner in alumina production

1) The system gives a possibility to modify in an economical way the conventional rotary kilns having high heat consumption.
2) The heat consumption of the modified calciner kilns is close to that of the most up-to-date stationary calciners and the capital costs of the modification are significantly less than those of a new stationary calciner.
3) As a result of the modification the capacity of the kiln can be increased by nearly the same percentage as the heat consumption is reduced by.
4) The stability of the rotary kiln improves as a result of its shortening increasing the life of the brick lining and consequently reducing the maintenance requirements.
5) The modification of operating rotary kilns can be carried out at a minimal production loss, since the towers with the cyclones can be constructed above the operating kilns and they have to be stopped only for the time required for cutting both ends of the kilns and for installing the connection pipes to the cyclone systems.

3.2 Suggestions for the improvement of maintenance

Aluterv-FKI gave suggestions in 1978 for the organizational formation of the maintenance division of the Korba alumina plant based on its experiences gathered in Hungarian alumina plants. Data regarding the amount of working hours spent on the maintenance of the main equipment were also handed over together with the time required for the smaller maintenance works as a function of the cycle time in the form of tables.

The Korba Alumina Plant established its maintenance organization in the past 8 years and by the beginning of 1988 the working hours spent on the maintanance of the equipment are close to, in some cases are equal to the values of the Hungarian practice. This fact proves the gradual improvement of the maintenance activities.

3.2.1 Reconditioning of worn-out parts in the alumina industry

Almost in all plant units of the alumina plants the parts of the equipment are exposed to severe wear and abrasion. The worn-out parts can cause failure of equipment, of plant units or even of a whole technological line.

To reduce the harmful effects of wearing and abrasion different kind of hardsurfacing techniques have been tried and introduced in the Hungarian alumina plants.

The aim was twofold:

- to increse the cycle time of equipment, and
- to decrease the cost of maintenance.

Wear-resisting layers can be made by one of the following surface treatment technologies:

- Hardening: Induction hardening
  Flame hardening
  Electron beam hardening

- Thermo-chemical diffusion handling:
  Chromating
  Carbo-titanizing
  Nitriding
  Carbonization
  Vanadizing

- Galvanic plating:
  Hard chromium plating
  Hard nickel plating

- Hardsurfacing:
  Powder dispersion
  Metalpowder dispersion
  Hardfacing
  Metal cladding

The main features of hardsurfacing techniques applicable for both the manufacture of new parts and reconditioning of wornout ones are summarized in the Table 3.2.1-1.

The techniques listed in Table make it possible to create a wear-resisting layer on a fairly tough base material.

On the basis of practical experiences highly alloyed welding materials gave the best results in case of wear and abrasion caused by bauxite minerals and hydrate.

If machining of the treated surface is required, such type of hardsurfacing technique should be selected, which gives a surface quality machinable by machine tools available in the central workshop of the Alumina Plant.

The economy of reconditioning of worn-out parts is as follows:

- It is more reasonable to rebuild the worn-out 10 to 20 % than to throw away the still sound 80 to 90 % of the part.
- By using regular reconditioning the stock of spare parts can be reduced.
- By selecting a proper hardsurfacing technique the use of expensive alloys for base material is not necessary. Application of structural steel is satisfactory.

- By hardsurfacing the equipment parts the cycle time of the equipment exposed to severe wear and abrasion is extended.

It was found in many cases that not the hardness of the surface exposed to severe wear but the percentage of harder components of the layer was decisive as regards resistance to wear.

Such harder components are different kinds of carbides. The good resistance of carbides against abrasion can be used only when the carbides are fused in a tough base material. In this case the hardfaced layer can not be chipped.

3.2.2 Hardsurfacing techniques used in the Hungarian alumina plants

In the Hungarian aluminium industry the following hardsurfacing techniques are used for both reconditioning of worn parts and the manufacture of new ones:

a) Hardfacing with arc welding
b) Metal powder dispersion

a) Hardfacing with arc welding

This technique is performed the same way as the conventional arc welding.

The weldability of highly alloyed steel causes many problems (such as crevice, chipping, sensitivity against corrosion, etc.).

Table 3.2.1-1

TECHNIQUES FOR HARDSURFACING
(Metal coating)

Manufacture of new parts		Reconditioning
Thin layer technique	Thick layer technique	Thin layer technique
Thickness of coating: 0.05-0.3 mm	Thiskness of coating: 1.5-3.0 mm	Thickness of coating: 0.05-3.0 mm
1. Thermal metal coatings: - diffuse chromating, rather sophisticated technology hardness: 1200 HV resistance to wear: good corrosion resistance: good 2. Nitriding: - hardness: 700 HV - resistance to wear: good - brittle surface 3. Metal powder dispersion: cold dispersion hot dispersion "Castolin" type dispersion Main features: - hardness: max. 62 HRC 760 HV - manual technology (it can be mechanized) - compressive strength of the layer is low - composition: low melting point metal with hard metal powder additives - metals of high melting points can not be processed - low impact energy of the powder - output rate: 4-8 kg/h - the surface generaly requires machining 4. Plasma powder dispersion: Every metal, metal carbide, nitride, oxide, boride, silicide can be dispersed Main features: - plasma ensures a unique high temperature - low heat input - very high impact energy (10 to 20 times higher than that of the metal powder dispersion technique) - automatically controlled technical parameters - the technique may be repeated, it ensures the same quality in different layers - porosity of the surface is low - hardness: 1200-3700 HV - machining of dispersed layers generally not required - output rate: 4 to 8 kg/h with metal 2 to 4 kg/h with carbides	1. Metal powder welding: Mostly used for reconditioning of press dies. (Metal powder welding is not used in the maintenance work of the Hungarian alumina plants). 2. Electric arc welding: - with coated electrodes - with pipe electrodes Main features: - the quality of work depends upon human factors - hardness: max. 60 HRC 700 HV - hihg welding rate - good reliability of the hardfaced layer, no peeling off - after hardfacing machining is required, in case of great hardness machining is difficult - deformination, warping can occur - thickness of the layer 1.5-8 mm - main fields of its application: reconditioning of equipment parts not requiring machining. 3. Hardsurfacing (CO_2, Mich.gas). Main features: Technology of hardsurfacing is well mechanized, independent from human factors - hardness: 60 HCC 700 HV - high welding rate - good reliability of the technology - rough surface, machining is required; in case of great hardness machining is difficult - heat load is high but uniform and can be controlled - no significant warping occurs - thickness of the layer: 1.5-8 mm - in case of highly alloyed steel cracks may occur - output rate: 1 to 3 kg/h 4. Hardsurfacing with powder filled electrodes: - main features are the same as for point 3 but the rate is higher, more than 10 kg/h.	1. Metal powder dispersion Castolin type 2. Hard chromating 3. Combined technology: - hand or machinized hardsurfacing - mechanical cleaning of the surface - plasma dispersion

Remarks: In case of hardsurfacing of new parts the base material can be selected in such a way that a special underlayer is not required. In case of reconditioning the use of underlayer solves the problem of weldability.

Such types of electrodes should be selected, which ensure the forming of an austenitic-ferritic structure, as this structure has favourable characteristics such as:

- it is not inclined to hot crack,
- has great elongation and good toughness,
- has high strength,
- has good corrosion resistance,
- its hidrogen absorpiton characteristic is good.

Some practical examples of reconditioning techniques used in Ajka and Almásfüzitő Alumina Plants are given in Tables 3.2.1-2 and 3.2.1-3.

The electrodes used in Hungarian alumina plants are shown in Table 3.2.1-4.

Table 3.2.1-2

REPAIR OF EQUIPMENT PARTS EXPOSED TO WEAR AND ABRASION IN THE AJKA ALUMINA PLANT

Serial No	Denomination of the part	Material of the part as per Hungarian standard	Material of the part DIN equivalent	Preheating oC	Method of repair	Repairing material type	Repairing material Manufacturer	Remarks
1.	Hammers of hammer mills	A50	St 50-2	250	Hardsurfacing by electric arc welding	Tricarbide 60	Csepel Fémüvek Hungary	Use of underlayer necessary
2.	Interconnecting shafts of digesters	C45	C45	150	Castolin type metal powder dispersion	10009 Borotex metal powder	Castolin Eutectic	For underlayer Castolin type material is used
3.	Piston rods (for hydraulic system of the mobil crane type CK)	CrNi steel	X10 CrNiTi 189	150	Eutalloy type metal powder dispersion	Eutalloy Cromtech 10680 metal powder	Castolin Eutectic	
4.	Control valves		-	250	Castolin type metal powder dispersion	Durchoc N metal powder	Castolin Eutectic	
5.	Shaft ends for disc filters	A50	St 50-2	200	Castolin type metal powder dispersion	10009 borotex metal powder	Castolin Eutectic	Use of underlayer necessary
6.	Nozzles for gas burners of rotary kilns	A60	St 60-2	200	Castolin type metal powder dispersion	10009 Borotex metal powder	Castolin Eutectic	
7.	Screw conveyors for hydrate feed to rotary kilns	KL2	HII	250	Hardsurfacing by electric arc welding	Tricarbide 60 electrodes	Csepel Fémmüvek Hungary	Use of underlayer necessary

Table 3.2.1-3

REPAIR OF EQUIPMENT PARTS EXPOSED TO WEAR AND ABRASION IN THE ALMÁSFÜZITŐ ALUMINA PLANT

Serial No	Denomination of the part	Material of the part as per Hungarian standard	Material of the part DIN equivalent	Preheating oC	Method of repair	Repairing material type	Repairing material Manufacturer	Remarks
1.	Shafts of the digesters (parts between the gearbox and the shaft of the agitator)	A70	St 70-2	150	Castolin type metal powder dispersion	10009 Borotex	Castolin Eutectic	In case of deep wear hardfacing Böhler FOX DUR 600 electrode is required
2.	Thrust bearings of digesters (bush and sleeve)	A60	St 60-2	250	Hardsurfacing	FOX DUR 600 electrodes	Bühler	For medium layer 18/8 CrNi electrodes, for underlayer Böhler Fox A7 electrodes
3.	Feeding screws for rotary kilns	A38	USt 37-2	250	Hardsurfacing	FOX Super Dur W80 electrodes	Böhler	For underlayer Bühler FOX A7 electrodes
4.	Hydrate flush mixes blades before the feeding screws	A38	USt 37-2	250	Hardsurfacing	FOX Super Dur W80 electrodes	Böhler	For underlayer Bühler FOX A7 electrodes
5.	Running surface of crane wheels	cast steel		-	CO2 hardsurfacing	DCMS-JG electrodes	Csepel Fémmüvek Hungary	Made by single purpose machine
6.	Places for bearings on shafts of electric motors	A60	St 60-2	-	CO2 hardsurfacing	VJH1; EMK6 electrodes	Csepel Fémmüvek Hungary	Made by single purpose machine
7.	Half shafts of gearboxes	A60	St 60-2	-	CO2 hardsurfacing	VJH1; EMK6 electrodes	Csepel Fémmüvek Hungary	Made by single purpose machine
8.	New valve seats	A45	St 45	150	Hardsurfacing	SAS 2JG electrodes	Böhler	Made by single purpose machine Protecting gas: argon

Table 3.2.1-4

Following electrodes are used in the Hungarian alumina plants:

Electrodes Type	Manufacturer	Hardness of layer HRC (HB)	Chemical composition	Remark
EK B18/8s	Csepel	-	C: 0.08 Mn: 1.4-1.6 Cr: 19 Ni: 9 Nb+Ta: 0.8	used for underlayer
Ef Mn14	Csepel	250-400 HB	C: 1.2-1.4 Mn: 12.00 16.00	used for overlayer
EK B Mn Cr	Csepel	150-200 HB	C: 0.2 Mn: 23 Cr: 10	used for overlayer
Tricarbid 60	Csepel	55-60	C: 3.5-6 Mn: 4.0-5.5 Cr: 17-22 W: 4-5.5 V: 4-5.5	used for overlayer
FOX SAS 2A	Böhler	-	C: 0.003 Mn: 0.9 Cr: 20 Ni: 10 Nb. killed	used for underlayer Gives austenitic structure
FOX Super dur W70Cr	Böhler	68-72	WC: 70 CrC: 10 Fe: 20	used for overlayer
FOX Super dur W80	Böhler	60-65	WC: 80 Fe: 20	used for overlayer
FOX dur 600	Böhler	54-58	C: 0.5 Si: 2.3 Mn: 0.4 Cr: 9.0	used for overlayer
FOX Lecurit 62	Böhler	63-64	C: 4.7 Cr: 30	used for overlayer
Gridur 46	Messer-Griesheim	55-60	C: 0.5 Cr: ∼ 9 Si: 1.5	used for overlayer
Gridur 34	Messer-Griesheim	42	C: 0.05 Ni: 60 Cr: - W: -	

Table 3.2.1-4 (cont.)

Electrodes Type	Manufacturer	Hardness of layer HRC (HB)	Chemical composition	Remark
OK 61 81	ESAB	-	C: 0.06 Mn: 1.5 Cr: 20 Ni: 10 Nb: 0.7	used for underlayer
OK 85 58	ESAB	47-52 53-57	C: 0.35 Mn: 1 Cr: 1.5 W: 8 Co: 2 Nb: 0.8	
OK 83 65	ESAB	58-63	C: 0.7 Mn: 0.4 Cr: 2.0	
Castolin 680	Castolin Eutectic	360 HB	Cr/Ni type	used for underlayer and for overlayer
P5	Avesta	-	C: 0.03 Cr: 22 Ni: 14 Mo: 2.5	used against corrosion load

b) Metal powder dispersion

The following metal powder dispersion techniques are used in the Hungarian aluminium industry:

- Castolin metal powder dispersion,
- Eutalloy process,
- plasma powder dispersion.

3.2.2.1 Castolin metal powder dispersion

The castolin technique can be regarded as a conventional hardfacing method, using oxi-acetylene welding torch to spray pulverous alloys on the surface of equipment parts. In the gas mixture the metal powder warms up, gets melted and bounds to the equipment part. The Castolin technique can be performed both for rotating and plain surfaces. Generally the technology consists of two phases. At first the underlayer is sprayed on the surface which promotes the adhesion at the equipment parts. In the second phase the overlay is sprayed which bounds well to the underlayer.

The castolin technique is a "cold" metal powder dispersion method as the layers are not bound by diffuse adhesion to the base material. This hardsurfacing technique can be used in those cases where no dynamic loads are expected.

3.2.2.2 Eutalloy process

The Eutalloy process is a hardsurfacing method using a special welding torch to spray pulverous alloys on the surface of equipment parts. Due to the pressure drop the gas mixture draws the metal powder through the flame. The metal powder warms up in the flame and collides with the surface of the base material. The powder gets melted and bounds to the base material.

The Eutalloy process uses a specially designed welding torch. The oxigen and acetylene get into a mixing chamber. The design of this chamber differes from the standard solutions.

The pulverous welding material is mixed with the gas through a nozzle connected to the chamber.

A plastic holding tank for powder can be connected to the welding torch by a bayonet clutch. The other parts of the welding torch, finger pipe, nozzle etc. are of standard solution.

Before welding it is necessary to preheat the surface, because the powder adheres only to a hot surface. In most cases it is sufficient to preheat the parts by the welding torch, in case of edge welding the preheating may be omitted. For preheating of big and complicately shaped parts a heat treatment in a furnace is recommended. Important rule: preheating might be done only up to 350 °C, the appearing of a blue colour.

When the blue colour appears, the flame should be kept at a distance of 40 to 50 mm from the part and the surface should be sprayed uniformly with metal powder. The quality of the first layer is very important, as it protects the base material from further oxidation, which would take place at higher temperatures.

Further work depends upon the thickness of the layer to be built up. If only a thin layer is needed (a minimum 0.05 mm thick layer can be dispersed), only the first protecting layer is melted in without any further powder feed. The work should be done carefully, because the welding material reaches its proper mechanical characteristics only by melting.

Good melting occurs when the colour of the part is dark purple. As the melting of the welding material can be observed easily, harmful overheating can be avoided. The core of the flame should be kept at minimum 5 to 10 mm from the surface, otherwise the material can get overheated. This technique is basically different from the Castolin metal powder dispersion methods, where the adhesion of the molten metal can not be increased beyond a certain extent. Here a good diffusion bond takes place in the course of melting which follows the metal powder dispersion phase.

It follows from the above that, when building up a thicker layer, the powder should be fed batchwise, as in the course of dispersion of the powder the surface cools somewhat, thus the dispersed layer has to be molten in.

The above discussed practical advises should be followed thoroughly, when building up edges.

Summarizing the advantages of the Eutalloy technique, it can be stated that this technique is one of the most applicable methods for welding highly alloyed steels.

The working temperature of the Eutalloy technique is only about 800 ^{o}C compared to the 3500 ^{o}C local fusion temperature of arc welding, the heat load of the welded material is lower and the heat distribution is more uniform.

Using this technique to worn out parts thrown away earlier they can be reconditioned with small effort. As the thickness of the layer can be easily controlled by the powder feed, little machining is required after welding.

The following powders give good wear and abrasive resitant layers:

- Borotec 10009
  Nickel based alloy

This alloy has good fusion characteristics and good edge firmness quality, thus it is used for reconditioning of cutting tools and shears. Its resitance to wear is excellent, 10 times better than that of hardened C60 steel. Bending and drawing tools may also be rebuilt with this material.

Chemical composition: Ni: 72 %
Cr: 15 %
Si: 4 %
B: 4 %
C: 0.7 %

Hardness: Max 275 HV (62 HRC)
Thermal heat expansion coefficient: 13.6×10^{-6} °C^{-1}

- BronzoChrom 10185
  Nickel based alloy

This alloy has good impact, corrosion and wear resistance characteristics. Because of its good fusion characteristics it is used to build up very thin (0.05 mm) layers.
Chemical composition: Ni: 91 %
Cr: 4 %
Si: 4 %

Hardness: Max 410 HV (42 HRC)
Thermal heat expansion coefficient: $16.2 \times 10^{-6}\ ^{o}C^{-1}$

- ChromTec 10680
  Nickel based alloy

This is used in those cases, when the higher hardness of the part is not essential. As the viscosity of the fused material is high, it is excellent for repairing damaged edges and angles.

It is used mainly for preparing underlayers in those cases where a thick layer is required.

Pretreatment of the surface

A good quality layer can only be obtained, if the surface to be handled is perfectly clean.

Before applying the Eutalloy powder dispersion technique the surface has to be machined by using a tipped tool (marked K10). The best surface can be provided by grinding in radial directions. Sharp edges should be avoided, transitions to be machined at a 30 ° angle. As the pretreated surface has to be kept clean, the powder dispersion has to be performed as fast as possible.

Preheating and dispersion

Prior to dispersion the surface of the part has to be preheated up to about 100 to 350 °C. The welding torch has to be kept in such a distance from the part that the end of the flame just touches its surface. The welding torch should be moved slowly and uniformly and the flame should be kept perpendicular to the surface of the part. In case of cylindric parts a good dispersion can be attained by chucking the part in a lathe. The part can be rotated with a speed of 15 to 20 r.p.m. Using automatic feed, the feed rate may be 5 mm per rotation.

The thickness of the dispersed layer

By using the RW Eutalloy and Castolin devices the thickness of the layer can be controlled accurately. However, it should be kept in mind that after melting the thickness of the layer will decrease by about 25 per cent.

To ensure a good fusion and bond the maximum thickness of a layer dispersed in one working phase should not exceed 0.5 mm. If a layer thicker than 0.5 mm is required, the final thickness should be attained by dispersing two or more layers. In this case the second layer should be dispersed and then fused after the fusion of the first layer.

This process should be continued until the final hardfacing thickness is reached. Due to economic considerations the thickness of hard surfaces obtained by the Eutalloy technique is generally between 0.6 to 1.5 mm.

3.2.2.3 Plasma powder dispersion

The plasma metal dispersion has been developed from the metal powder dispersion technique begun in the early years of this century. The physical concept of "plasma" was introduced by LANGMUIR in 1923 for the special gas condition in which the gases become conductive by separating the electrons from the atoms. This condition is often called as the fourth state of materials and in most cases is connected with high temperature.

The plasma metal dispersion operates with partly ionized and fully ionized plasmas. Thus there are electrons, ions and neutral particles in the plasma.

The first plasma dispersion devices were developed in 1939 already, but devices good for practical application appeared in the market only after 1956. From a laboratory technique the plasma metal dispersion has become a practical procedure widely used in the different fields of industry.

The plasma dispersion technique is used to improve the quality of the surface of an equipment part to such an extent, which can not be achieved with the base material only. This technique is used not only for a single part, but for parts of serial production. In case of a single part a manually operated dispersion burner is used, which allows the combination of different kinds of materials, e.g. the dispersion of high melting point materials or anti-corrosive materials onto mild steel, the dispersion of ceramics onto metals or vice versa, moreover the dispersion of plastics onto other materials, or metals onto plastics.

The dispersion techniques widely used in the industry are the arc, the flame and the plasma dispersion techniques. The main features of these three dispersion techniques are drawn up in Table (3.2.1-5). It can be seen from the table that out of the three dispersion techniques the plasma dispersion offers the best quality of layer and a wide variety of applicable materials.

The modern plasma dispersion device available in the market has an electrical output at the burner of up to max. 70 kW (max. 100 kW at the rectifier) using two-atomic nitrogen gas for plasma gas.

The gas velocity exceeds 1000 m/s and the velocity of the powder particles reaches 600 m/s. Due to the high kinetic energy of the particles the layer has a good adhesive bonding to the surface of the base material.

High performance devices work with pure argon or argon-hydrogen and argon-helium mixtures. The main advantages of the media are that they ensure a stable arc and low working tension. Though the entalpy of argon and helium is lower than that of the two-atomic gases, however, the temperature of the plasma is at a given energy higher than in case of nitrogen and hidrogen.

A further advantage of the plasma technique is that all the working parameters (gas pressure, gas flux, electrical output, powder load) can be adjusted before putting the device into operation. After pushing the start button the control system of the equipment adjusts and controls the working parameters to the prescribed values.

Table 3.2.1-5

THE MAIN FEATURES OF THE FLAME, ELECTRIC AND PLASMA DISPERSION TECHNOLOGIES

Technologies	Material for dispersion		Available energy for		Heat resistance against the environment of the material to be dispersed	Output	Required "know-how"
	shape	material	melting	acceleration			
Flame dispersion	wire	mostly metals	T=3000 °K energy density 10^3-10^4 W/cm^2	low V=150 m/s	great oxidation and carbonation	low, metal 5 to 8 kg/h ceramic 1 to 2 kg/h	little
	powder	every kind		V=30-60 m/s			
Electric arc dispersion	only wire	conductive metals	T=4000-10000 °K energy density 10^4-10^5 W/cm^2	low V=150 m/s	very strong	high, 15 to 25 kg/h	little
Plasma dispersion	only powders	every kind	T=10000-20000 °K energy density 10^5-10^6 W/cm^2	high V=600 m/s	weak	medium, metal 4 to 8 kg/h ceramic 2 to 4 kg/h	much

versatility

quality

The fluctuations in the plasma flux are compensated automatically within a range of ± 2 %. If the deviations in the cooling of the plasma device, in the gas pressure or in the plasma flux are more than ± 2 %, the device stops automatically.

This stoppage is also controlled automatically. The output of the device decreases gradually and the final switchoff phase ends in a pure argon plasma gas flow. After the switch-off, the controlled argon after-flow serves for cooling the nozzle and the electrode, thus increasing the service life of these two parts.

The plasma dispersion technique may be used for the manufacture of new equipment parts as well as in the maintenance work for reconditioning the worn-out parts.

The plasma dispersion technique is mostly used in the following cases:

- wear resisting protecting layers,
- heat isolating layers,
- inlet layers,
- corrosion resistant layers against hot gases.

In the following paragraphs some practical applications are shown and the metals, carbides and oxides are discussed in more details.

Tungsten

Tungsten has the highest melting point (3410 °C) among the metals, its density is 19.3 g/cm^3. It has good heat and electric condictivities and its resistance against contact erosion is fairly good.

Tungsten can be well dispersed by argon, argon-hydrogen or nitrogen plasmae.

Molybdenum

Its melting point is fairly high (2620 °C), its density is 10.2 g/cm^3, much lower than that of tungsten, which is advantageous in certain cases.

Because of this and some other reasons molybdenum is widely used in the plasma dispersion technique. Molybdenum has good dispersion characteristics, gives a thick splashed structure with good friction and abrasion resistance features.

In the manufacture of tools the substandard parts can be repaired by the molybdenum plasma dispersion technique. Instead of manufacturing new parts the repair of substandards parts offers an economical advantage.

Carbides

Pure carbides are rarely used in the dispersion technique. They are mostly used with binding materials as in the synter technique, too.

For tungsten carbide, the most important hard carbide the binding material is cobalt in most cases, for chromium carbide a chromium-nickel alloy.

For the dispersion of hard carbides argon or an argon-hydrogen mixture are used, but pure nitrogen can also be applied. As the decomposition temperature of tungsten carbides is exceeded in the plasma dispersion technique, the composition of the layer will be different from the composition of the powder.

The quality of the layer will depend mostly upon the heat flux of the plasma gas into the metal (carbide) powder, the holding time of the powder in the plasma gas, the particle size of the tungsten carbide and its cobalt content.

Fine powders decompose easier than the powders with coarser particles. As the carbon loss in the layer can exceed 50 % in some circumstances due to decomposition, the remaining carbon content can decrease below 3 %.

The hardness of the layer depends partly on the starting cobalt content and partly on the dispersion conditions. Some informative value for the hardness of a tungsten carbide dispersion with a 12.5 % cobalt content is 1000-1300 kg/mm^2 measured by the Vickers method in case of 100 to 300 g load. Chromium carbide dispersions are softer, the carbide is Cr_3C_2 which decomposes during dispersion, the hardness is only 800 kg/mm^2.

Hard metal splashing is mainly used to provide wear and abrasion resistant surfaces; tungsten carbide with cobalt as binding metal is more advantageous is case of room temperature, but for higher working temperatures and corrosive conditions Cr_3C_2 carbide with Cr-Ni binding materials is more favourable.
Mechanical stirrers for the chemical industry are coated with tungsten carbide and cobalt materials, but chromium oxide coatings are also widely used.

Oxides

The application of oxides is increasing in the plasma dispersion technique. In case of wearing load aluminium oxide alloys with 3, 10 or 40 per cent titanium oxide additive work well. With the plasma dispersion technique these alloys ensure about 2400 HV microhardness, which is much better than the maximum hardness of 1100 HV attainable with hard chromium coating.
Chromium oxide can be easily used in the plasma technique, it gives a good, wear resistant coating. The method is applicable in pump manufacturing.
The layers dispersed by titanium-oxide (TiO_2) are not as hard and wear resistant as the layers made from Al_2O_3 or Cr_2O_3 oxides, but the structure of the coating is dense and very fine surfaces can be machined by grinding and polishing.
The plasma dispersion technique has been used succesfully in the Almásfüzitő Alumina Plant, for reconditioning of worn-out parts.

3.3 Suggestions regarding the instrumentation and control system

3.3.1 Review of the present state of instrumentation

Aluterv-FKI's experts studied the instrumentation and process control system of the Korba Alumina Plant during their site mission. Basically pneumatic type instruments are used except for the temperature measurements throughout the plant. There are some problematic measurements regarding the sensing method of parameters to be measured (dip tube with air bubbling at level measurements, weir boxes at flow measurements, dip tube systems at density measurements, etc). Other problems arise in connection with the pressure measurements with the pressure tapping devices (oil seal pots are frequently clogged). At some applications the control valves are the source of problems regarding their life duration (settling and washing area). With the present manpower and equipment (the number of qualified personnel, the lack of calibrating devices, repairing instruments, machines, etc.) it is very difficult to cope with the requirements of keeping all the instruments in operation.

3.3.2 Updating the present instrumentation and control system

Taking into consideration the future development, the existing instrumentation system should be modified using electronic instruments instead of the pneumatic ones. The sensing and signal processing is based on the standard 4-20 mA signal range and after an electro-pneumatic signal conversion the remaining control valves could get their control signal using pneumatic actuation. The signal conversion could be done by using either e/p signal converters or control valves equipped with electro-pneumatic positioners.

At some control points the complete elimination of the control valves by replacing them with r.p.m. regulated pump drives would provide the following benefits: savings in electric power consumption regarding liquor/slurry delivery, reduction of pump wear as the pumps would run on a lower r.p.m. when a smaller quantity of fluid had to be delivered, not to mention the elimination of wearing and maintenance of the control valves. This method may favourably be applied where slurries of abrasive nature or liquors of huge quantities are to be pumped through a pipe system having considerable dynamic losses, as the flow-rate of the fluid would be controlled, instead of increasing or reducing the hydraulic resistance of the piping system (as in case of control valves), with the modification of the r.p.m. of the pump drive.

Based on the experiences gained during our site mission the following suggestions are made regarding the updating of the instrumentation:

Control rooms

- The modernized control system is suggested to be constructed using electronic central devices having the capability to couple them to a higher level system (microprocessor system, data logger) in a later phase. The rooms need to be renovated and updated as well, having new control panels and equipment. The area control rooms are suggested to be organized in the following way:

No. 1 - Ball-mil. Desilication and Digestion control room.
No. 2 - Settling and washing, Red mud filtration and Lime handling control room.
No. 3 - Hydrate filtration and precipitation control room.
No. 4 - Calcination control room.
No. 5 - Evaporation I., II. and III. control room.

Two new control rooms are to be contstructed one for the instruments of the thre evaporators, and another one for the instruments of the hydrate filtration and precipitation. The remaining control rooms need renovation. Provision is kept for installing new, six channel recorders for each control rooms as well as totalizers for accounting the important material inputs and outputs of the plant units.

A serious cooperation is needed with the power electric department and the mechanical engineering department to determine those equipment that are to be remotely operated (some pumps need closed delivery valve when started, it means remote operation of the corresponding valves). The display of the running status for the main equipment of the plant unit is suggested to be brought into the control rooms. The cost estimate for the above mentioned items is included in the appropriate lines of the cost estimation table.

Level measurements

- For measuring the level in atmospheric tanks the dip tube system is used. With its frequent clogging and scaling it demands a lot of time-consuming maintenance activity. It should be replaced by extended diaphragm type transmitters and transmitters having a wafer element in their diaphragm type sealing system. The use of these transmitters is suggested in the evaporation, red mud filtration, hydrate filtration.

- For measuring the level in the flash tanks in digestion, dP transmitters are used with steam purging. Though this method seems to be more reliable as compared to the previous one, it needs checkings. The gamma-ray level measurements using radio-isotopes are much better according to the experiences gained with their application in the Hungarian alumina plants. Because there is no direct contact between the process liquor and the sensing parts of the level measuring loops, this method is not going to give maintenance troubles for the instrumentation personnel. The use of multi-channel central devices provide the benefit of cost reduction as there is the possibility for having common power supply units for the central instruments for a whole line of flash tanks.

- For measuring the level in hydrate bins and alumina silos the testing of a special istrument designed for this kind of application was suggested. The name and address of such firms were given to the experts of the Korba Alumina Plant to make contact with them.

Flow measurements

- For measuring slurries and liquors either weir boxes or segmental orifice plates are used. These kinds of measuring devices have the following problems: they need regular and frequent checking and cleaning, they are not sufficiently accurate and reliable. Instead of them the utilization of electromagnetic type flowmeters is suggested. When sizing and selecting them the following requirements must be met: the proper velocity range of the flowing fluid must be assured (1.0-1.5 m/sec at maximum flow-rate), the lining of the flowmeters must be made of polytetra-fluorethylene (TEFLON), the sensing electrodes must be made of corrosion and abrasion resistant materials, preferably with the provision of the possibility of boiling away the deposits (self cleaning), the applied electromagnetic field must be of the switched d.c. kind and, last but not least, prior to their installation for protecting the inside lining they are to be equipped with protective flanges.

- For measuring the H.P. steam consumed by the digestion, the steam parameters (pressure, temperature) must be evaluated whether they are sufficiently constant. It is advisable to use automatic pressure and temperature compensation, if the mentioned parameters fluctuate resulting in the variation of the specific volume of the steam to be measured. Special computing devices are available from several instrument manufacturers for this purpose. On the basis of the state equations of the steam the computing device computes the flow-rate taking into consideration the actual values of the auxiliary parameters (pressure, temperature).

- For measuring the fuel oil consumption of the calcination, instead of the rotameter presently used, the application of a rotary piston type flowmeter is suggested. This type of instrument works reliably and accurately in the Hungarian alumina plants if the following requirements are met.

At least two fine filters should be installed parallel to each other and upstream of the flowmeter, one of them works, the other is a hot spare. Should the working filter get clogged, it may be cleaned after the flow is valved to the other one. Whatever fine the filter used, the protection of the flowmeter is best assured by the proper treatment of the fuel oil to be measured (the daily fuel tank should be heated, be high enough to let the sand and other solid particles settle down, and the suction pipe of the fuel oil pump should be installed at a height, where the oil is sufficiently clean).

- For measuring solids flows (bauxite and hydrate) it is suggested to replace the existing weighing belts. The type of weighing belts proven to give the best results in the Hungarian alumina plants are eletronic ones having practically no moving parts.

Pressure measurements

To avoid the problems caused either by the clogging of the oil pots' input section or by the escape of the oil the pressure tapping points should be modified to make them suitable for the use of the diaphragm type sealing system. A lot of time-consuming maintenance work may be avoided regarding the cleaning of oil pots and the proper refilling of oil pot - inpulse pipe - transmitter chamber systems. The pressure measurements may be made reliable with manufacturer made diaphragm type sealing systems.

Density measurements

The presently used density measurement systems utilize the double dip tube method provided either with a reference chamber or with transmitter zero elevation. The problems mentioned about the dip tube system at the level measurements stand for the density measurement systems as well, regarding the primary elements of them. For a more reliable and accurate solution the application of gamma-ray density measurements was suggested. This method has all the benefits which come together with the fact that no parts of the instruments are wetted by the process liquors. These kinds of instruments are succesfully used by the Hungarian alumina plants, as they are of the so-called "install and forget it" type instruments regarding maintenance work.

Temperature measurements

The majority of the temperature measurements in the Korba Alumina Plant uses a two wire system. Even if the platinum sensors are perfect and accurate, the measurement is loaded with an error which is due to the fact, that the ohmic resistance of the cable (which depends on the ambient temperature) between the resistance bulb and the central instrument is included only in one arm of the Wheatstone bridge of the instruments. This means that not only the sensor's resistance is changing (with the measured temperature), but the cable's resistance, too (the latter with the ambient temperature). With the suggested three wire system the resistances of two wires of the cable are included into separate arms of the Wheatstone bridge, respectively, assuring a complete compensation of the variations of the ambient temperature. The third wire is coupled to the high input impedance instrument, that is why the change of the wire's resistance has no effect on the accuracy of the measurement.

The suggested modification of the equipment in the digestion (tube-in-tube type heat exchangers) would require the application of a new type of temperature sensor. This is a kind of surface temperature measuring device, having a very small heat capacity. They are installed to the outside surface of the small diameter slurry pipes of the tube heating elements providing proper heat contact and requiring proper heat insulation, using of course a three wire cable system.

1. Surface temperature sensors

MINCO THERMAL RIBBON
(thin, flexible resistance thermometer detector)
range: -200 to 200 °C

CAREL COMPONENTS LTD
24 ENDEAVAOUR WAY, WIMBLEDON
LONDON, SW19 8UH
TEL.: 01-946 9882, TELEX: 928717
FAX: 01-946 6259

2. Resistance thermometers for measuring surface temperatures

Order No.: 7MC1032-1AA-Z

SIEMENS INDIA LTD
Representative Offices in Bombay, Ahmedabad, Banglore,
Calcutta, Hyderabad, Madras, New Delhi

Control valves

- Two methods are used in the Hungarian alumina plants for modifying the flow-rate of different slurries. One of them uses special slurry valves, the other the r.p.m. regulation of the delivery pump.

Unfortunately these control valves are not commercial types, they were developed by the Hungarian alumina plants. The basis of these control valves is the traditional hand operated straight valve body, provided with a seat and a plug made of tungsten carbide (WIDIA) and a pneumatic servo motor equipped with an electro-pneumatic positioner. The seat and the plug are similar to those which are used in the control valves of the flash tank level control loops. It is suggested for the experts of the Korba Alumina Plant to make a trial with this kind of control valve. As they have a source for these kinds of seats and plugs (flash tank level control valves) it would be practical to construct and try at least one of such control valves.

As for the r.p.m. regulation of slurry pumps, two kinds of regulators are used in the Hungarian alumina plants. One of them is the so-called hydraulic coupling, which is installed between the electric motor and the pump. This is a kind of hydro-motor, its loaded shaft is driven by the oil filling which transfers the momentum of the electric motor. There is a slip between the constant r.p.m. of the driver shaft and the changing r.p.m. of the driven one, which depends on the level of oil filling. By changing the level of the oil filling the slip, that is the r.p.m. of the driven shaft may be changed, which means a change of the pump r.p.m. as well.

The other type of r.p.m. regulating unit controls directly the r.p.m. of the driving motor of the pump using frequency converter power electronics (thyristor).
Both of the regulators are working satisfactorily in Hungarian alumina plants.

- The control valves used for the high pressure condensate level controller loops are single seated angle body valves. When sizing these control valves the vapour formation within the valve body must be taken into consideration. Some supporting information was mailed to the Korba Alumina Plant including a method for the sizing of control valves for flashing service.

With the suggested equipment modification (tube-in-tube heat exchanger) a new method is suggested for the condensate separation. Though this is specified by the mechanical engineering side, it is one kind of a regulator without auxiliary energy, its operation is based on the theory of thermodynamics. The valve begins to open, when "cold" condensate enters it. When the condensate begins to run out, some "hot" steam enters the valve, and it begins to close. As a result, it lets through only the condensate flow and stops the flow of the steam.

The manufacturer of a so called STEAM GARD device is:

ENGINEERING RESOURCES INCL. USA

They have representatives in Austria:

CIRKEL-A
Verfahrenstechnische Gerate G.m.b.H.
1150 Wien, Pillergasse 13-15
TEL.: 0222/872674
TELEX: 134436 cirk a

3.3.3 Instrumentation in the service of process control

For improving the technological operations, modifications are suggested for some measuring and control loops. Basically with the suggested system modification (from pneumatic to electronic) the new, more accurate and reliable measurements and controls will significantly improve the control of the process. The cost of instruments needed is included into the cost estimation.

Preparation of digesting liquor

The most important target is to provide a stable, defined concentration of the digesting liquor. This may be achieved by blending the components in the desired ratio. An updated ratio control at the liquor preparation would assure the

quality of the digesting liquor based on the reliable measurements and control of the flow-rates of the components, taking into account the laboratory data with a slight modification of the ratio value when it is necessary.
The proposal for the control system is given in Fig. 3.3.3-1.

Control of the bauxite to digesting liquor ratio

With the new measuring devices for bauxite and liquor flows (a reliable weighing belt and magnetic flowmeters) the desired ratio value may be assured. The periodic laboratory analysis data regarding bauxite and liquor quality and the target molar ratio must be taken into consideration when adjusting the ratio of the components. The installation of a flow control loop is suggested for assuring a proper lime addition. The flow-rate of lime milk would be determined by the mass flow of bauxite in the adjusted slurry. This signal may be produced from the flow-rate signal of the slurry (volumetric flow) multiplied by a signal which is proportional to its solids content (a gamma-ray density measuring loop would provide the latter).

Control loops for the dilution

The purpose is twofold. The proper settling of the solids is to be assured in the settlers and the hydrolysis losses are to be reduced by adding some spent liquor. Flow control loops are suggested for the first washer overflow and for the spent liquor cascaded with a dilution density control loop. The two liquor flows should be connected into a ratio control loop. This ratio may be modified, if necessary, based on the data of the periodic lab analyses.
The proposal for the control system is given in Fig. 3.3.3-2.

The protection of plate heat exchangers

The plate heat exchangers are very sensitive to overpressure. That is why the reliable operation of the pressure control loops is very important on both sides. At present only pressure measuring loops are there, their completion to complete control loops is suggested.

Compressed air system

It is very important to assure a reliable compressed air supply of the precipitator line and of the desilication. First priority should be provided for them. (See the corresponding section for mechanical engineering.)

Measuring loops for the hydroseparator

For the reliable operation of the hydroseparator the installation of a solids level alarm loop is suggested to indicate if the level of the compacted solids is between the maximum and minimum values. The utilization of a gamma-ray level alarm is suggested. For measuring the solids content of the hydroseparator underflow a gamma-ray density measuring loop is suggested, namely the type which may be installed around the slurry pipeline.

Instrumentation for the interstage cooling

For achieving the desired temperature profile of the precipitation line, interstage coolers are provided. To improve the efficiency of the cooling, the utilization of temperature control loops is suggested.

Lime milk preparation

The quality of lime milk prepared may be assured by controlling its density. For density measurement the utilization of a gamma-ray density transmitter is suggested. To prevent the clogging, scaling of them a thorough flushing is to be provideg utilizing remote controlled valves (on-off operation).

Causticization, wash water preparation

The lime milk added is to be controlled in the function of the solids mass flow-rate of the last washer underflow. The flow-rate signal multiplied by the signal of the solids content will provide the set point of the lime milk flow controller. The utilization of a temperature controller is also suggested at this point for achieving the desired 95 ^oC causticization temperature using LP steam for direct heating.

3.3.4 Suggestions for installing a communication system

To make the plant operation more effective, the installation of a communication system is suggested. It is intended to accelerate the exchange of information between the field operation personnel and the operators in the control rooms. The provision of interconnection of the control rooms is also suggested in the frame of the communication system.

This system is succesfully operating in the Hungarian alumina plants with the following features: it is organized around the control rooms with its central units located in the same and its field units are placed in the vicinity of the technological points where the operators are performing their activities (closing and opening valves, starting and stopping pumps, etc.). There is a possibility to communicate with the individual field operators or to give general announcements concerning the plant unit. The field operators are able to report to the control room only, because they cannot communicate with each other directly. The number of central units is equal to the number of control rooms and depending of the size of plant units 2 to 4 field units are sufficient for each control room. To help the field service activities of the instrumentation personnel the utilization of at least 6 to 8 walkie-talkie units is also suggested.

3.3.5 Suggestions for the installation of a management info system

After updating the instrumentation (that is, all the instruments regarding the main technological parameters are electronic types) the possibility of installing a data logger is given. This data logger collects the data of 30-40 measured or controlled parameters. The system has the capability of performing calculations to provide information about the state of the plant in a condensed form for the plant management. Printing and logging devices are also included as well as the possibility to check any processed value collected at any time.

3.3.6 Improving the maintenance activity of the instrumentation personnel

Manpower

The existing personnel is not sufficient to perform all the duties of maintenance work regarding the instrumentation of the whole alumina plant and the steam plants. The total number of the existing measuring and control loops is more than 700. The number of skilled technicians and helpers together is around 40 according to the data given by the experts of the Korba Alumina Plant, which figure is lower than sanctioned by the Local Management. For a rough comparison , the instrumentation personnel of the Ajka Alumina Plant is give here.

The task of this personnel is in connection with the alumina plant instrumentation (the number of measuring and control loops is around 400).

Electronics technicians	15
Instrument technicians	11
Fitters	12
Lathe operator	1
Helpers	4
Total	43

It must be noted that whenever it is necessary the other departments give helping hands at the installation and dismantling of control valves, magnetic flowmeters, etc. The

Ajka Alumina Plant has also some kinds of service contracts for the maintenance of some special equipment.

With the future electronization of the instrumentation of the Korba Alumina Plant the employment of electronics technicians is suggested.

Referring to the new types of instruments suggested, their demand for maintenance work seems to be reduced. With this in mind the personnel is suggested to be doubled gradually including the employment of electronics technicians.

Equipment

The equipment of the instrumentation repair shop should be modernized for assuring the effective maintenance work. Not only special calibrating devices are necessary (temperature bath for calibrating resistance thermometers and thermocouples, calibrating devices for checking and setting pressure and dP transmitters), but also special tools for making the fitting, installing activities easier and quicker. The installation of an instrument lathe is suggested for making fine parts of the instruments to be repaired. When ordering the new electronic instruments it is suggested to purchase special calibrating devices, tools for their quick calibration directly from the manufacturer (for example a flow simulator for calibrating magnetic flowmeters).

3.3.7 Suggestions for the priorities of the activities

Considering the high cost of the updating of the instrumentation and control system, it should be done gradually. A proposed priority order is given as follows:

- modernization of the instrument repair shop for repairing, testing and calibration of electronic instruments including specialized measuring devices to be installed during the updating process, walkie-talkie units, etc., (Approximately: 1,100,000 Rs)

- modernization of digestion instrumentation, of important operational instruments like steam flow, pressure, temperature, slurry flow; installation of a loudspeaker network for the digestion area, communication links between the digestion control room and the control rooms of connected processes (evaporation, grinding, settling and washing), (Approximately: 16,000,000 Rs)

- modernization of the instruments of the precipitation, hydrate filtration area, first of all of vital instruments like PHE overpressure protection, liquor flowmeters, loudspeaker network for the area, communication links to the control rooms of connected processes, (Approximately: 10,000,000 Rs)

- modernization of the instrumentation of the settler and washer area, loudspeaker network, communication links to other control rooms, (Approximately: 6,000,000 Rs)

- modernization of the instrumentation of evaporation and the remaining plant units, loudspeaker network, (Approximately: 12,400,000 Rs)

- modernization of the instrumentation by providing ultimately the data logger as a management reporting system, (Approximately: 1,200,000 Rs).

3.3.8 Cost estimation for updating the instrumentation

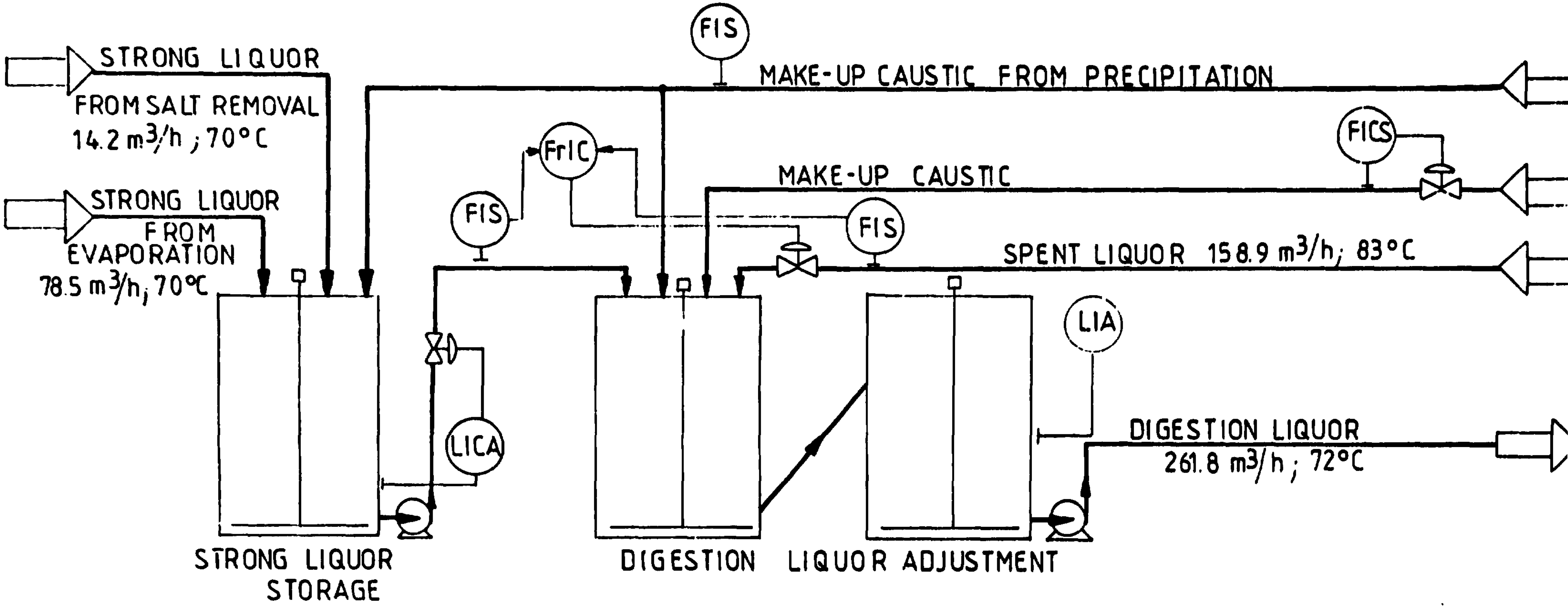
Instrumentation

Denomination	Estimated unit price (in Rs)	Number of devices	Total price (in Rs)
- electronic transmitters (p, dP)	50,000	90	4,500,000
- gamma-ray level measurements	300,000	14	4,200,000
- gamma-ray level alarms	150,000	2	300,000
- gamma-ray density measurements	300,000	10	3,000,000
- magnetic flowmeters	150,000	16	2,400,000
- rotary piston type flowmeters	150,000	2	300,000
- alumina silo level measurements	50,000	2	100,000
- cable modification for Pt 100	4,000	70	280,000
- new temperature measurements	10,000	40	400,000

Denomination	Estimated unit price (in Rs)	Number of devices	Total price (in Rs)
- variable speed drives	200,000	10	2,000,000
- level control loops for digestion pure condensate	50,000	8	400,00
- central instruments			
- indicators, controllers	40,000	120	4,800,000
- 6 channel recorders	60,000	10	600,000
- electronic totalizers	30,000	50	1,500,000
- two new control rooms (panels etc.)	400,000	2	800,000
- renovation of three control rooms	400,000	3	1,200,000
SUB TOTAL FOR INSTRUMENTATION			26,780,000
ADDITIONAL 5 % FOR INSTRUMENTS NOT COVERED			1,340,000
INSTRUMENTATION			28,120,000

Denomination	Estimated unit price (in Rs)	Number of devices	Total price in (Rs)
Equipment for instrument repaire shop			1,000,000
Loudspeaker network for plant units	90,000	10	900,000
Communications links between control rooms	75,000	10	750,000
Walkie-talkie units	12,000	8	96,000
Data logger for 40 analog signals			1,200,000
CRT terminals + printers for control rooms	350,000	5	1,750,000
Mounting materials (20 %)			6,764,000
Erection, installation costs (15 %)			6,100,000
GRAND TOTAL FOR UPDATING			46,680,000

ROUNDED FIGURE: 46,700,000 Rs



DIGESTION LIQUOR ADJUSTMENT (PROPOSAL) CONTROL SYSTEM

Fig.3.3.3-1

CONTROL SYSTEM FOR DILUTION
(PROPOSAL)

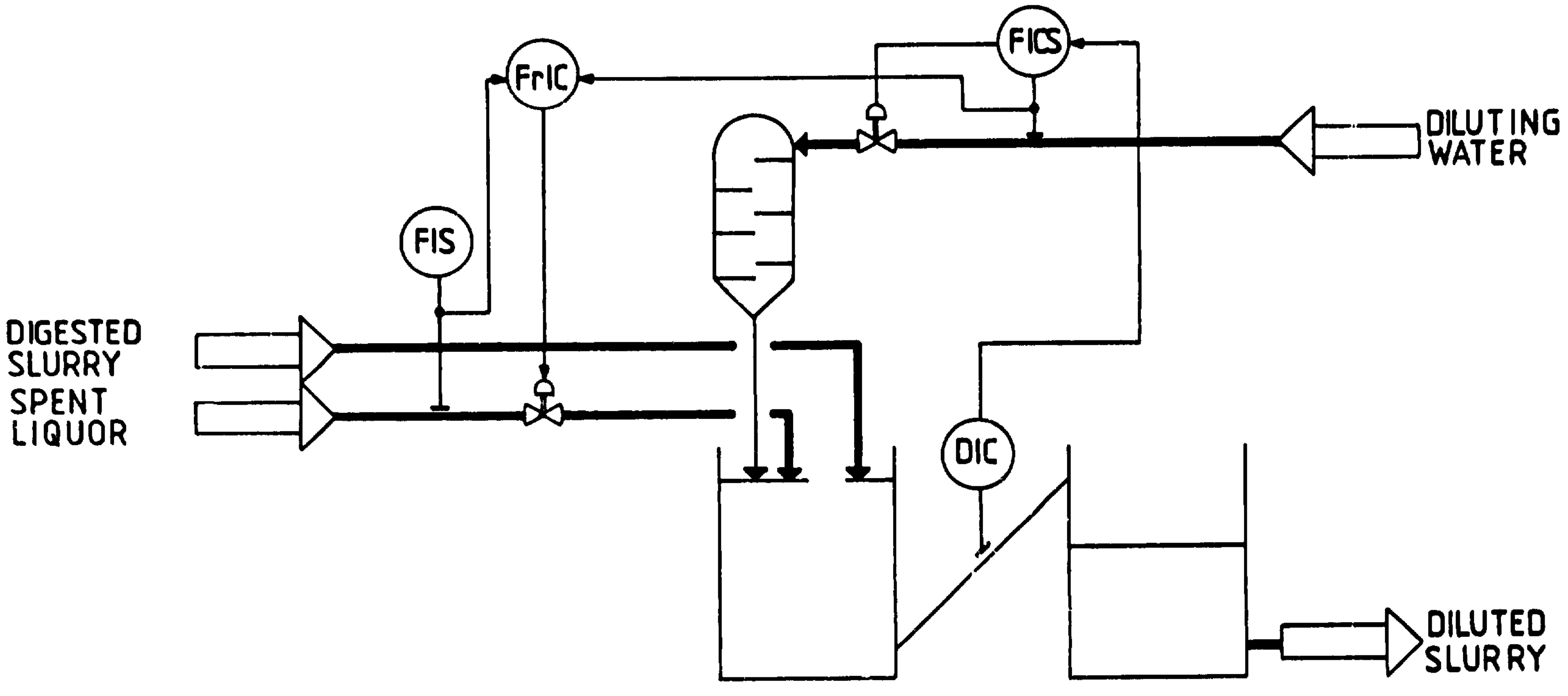


Fig. 3.3.3-2

4. Technical material handed over to Balco at Korba

4.1 De-aeration of the digesters

The gradual leading off from the digesters of the inert gases formed during the process may be achieved by a small size hole bored into the output slurry pipe. If this de-aerating hole is not prepared during the maintenance work (if the slurry pipe is replaced), then the gas phase formed during the operation will occupy the upper part of the digester displacing the slurry to be digested.

The heating coils of the digesters containing some gas phase are not able to heat up the slurry sufficiently resulting in the sharp degradation of the digestion yield. That is why the slurry pipes of all heated digesters must be checked during maintenance. The location and shape of the de-aeration hole is shown in Fig. 4.1-1.

4.2 Replacement of the slurry preheaters with tube-in-tube type heat exchangers

The cycle time of the vertical floating head type heat exchangers is not satisfactory. Because of the frequently required cleaning, the difficult maintenance work and the low heat transfer coefficient the replacement of the preheaters with tube-in-tube type heat exchangers was suggested.

Drawings were handed over to Balco in Korba (Figs. 4.2-1, 4.2-2, 4.2-3) containing the equipment and piping information about the tube-in-tube type heat exchangers suitable for replacing the floating head type preheaters. Based on the technical material handed over the replacement of two preheaters will be carried out by the experts of the Korba Alumina Plant at one of the digesting lines.

4.3 Cleaning of the precipitator tanks

It is practical to clean the first six tanks of the continuous precipitator line by taking them out of operation every half year, while the remaining tanks should be treated similarly every year.

A horizontal preheater with 200 m^2 heating surface is used for preheating the liquor utilized for the caustic cleaning of the precipitators. For dissolving the scalings the 2000 m^3 volume of the precipitator is completely filled up with caustic washing liquor and it is circulated during the cleaning process. The cleaning of the tanks lasts presently for 14 to 18 days.

A sectioned washing chamber cleaning method was suggested, at which the use of 500 to 800 m^3 washing liquor is sufficient.

The construction of the sectioned washing chamber system is shown in Fig. 4.3-1 enclosed. With this method the solution of scaling is more effective because of the high velocity (0.8-1.5 m/sec) at the contact surface between the scaling and the liquor phase, so the cleaning of the tank may be finished in a shorter period.

4.4 Updating the air-lifts of the precipitator tanks

By the modernization of the air-lifts in Hungarian alumina plants the overall efficiency of the circulation was improved during the past decade and the electric power consumption was reduced by 18-22 %.

The result of the modernization has two components:

- the air feeder made of wear-resistant material according to the enclosed drawing No. 4.4-1. (The modified air feeder is capable of providing a more disperse air bubble distribution, so the slip of the slurry relative to the air is reduced. The average size range of the air bubbles is 6-12 mm. Previously the air feeders were made of non-wear-resistant material and so at the end of the cycle time they were worn out. Because of the worn out feeders the flow conditions got out of tune and the air consumption increased.)

- the reduced air pressure before the air feeder. (Previously the pressure of the compressed air was between 5.2 and 5.4 bar. According to the operational experiences the pressure of the compressed air may be reduced by 0.25 bar without the degradation of the agitation.)

These two modifications result in the decrease of the electric power consumed.

4.5 Increasing the productivity of the seed filters

According to the practice of the Hungarian alumina plants the seed material is filtered by disc filters from the precipitated slurry. The previous construction of the disc filters used compressed air with a slowly rising pressure (up to 0.4-0.5 bar) to remove the hydrate.

This slowly rising air pressure did not assure the removal of hydrate from the whole area of the disc segments. Because of this the productivity of the filter was smaller, averaging between 0.8 and 1.0 t/m^{2}h.

The modernisation of the filters resulted in the increase of the productivity. It was achieved by an impulse-like pressure increment (0.4-0.6 bar), applied at the beginning of the separation of the hydrate cake. The compressed air almost completely removes the seed material by its dynamic effect from the surface of the filter segments. This method of filter cake removal is especially suitable in cases when the moisture content of the hydrate is high or it varies with the time. The cake separation system operating according to the new method is shown in Figs. 4.5-1 and 4.5-2.

The capacity increase is about 15-20 % but it depends on the size distribution of the hydrate particles, of the moisture content of the hydrate and of the quality of the filter cloth used.

Care should be taken that the pressure of the compressed air does not exceed the maximum 0.6 bar value, othervise the filter cloth of the filter segments may be torn.

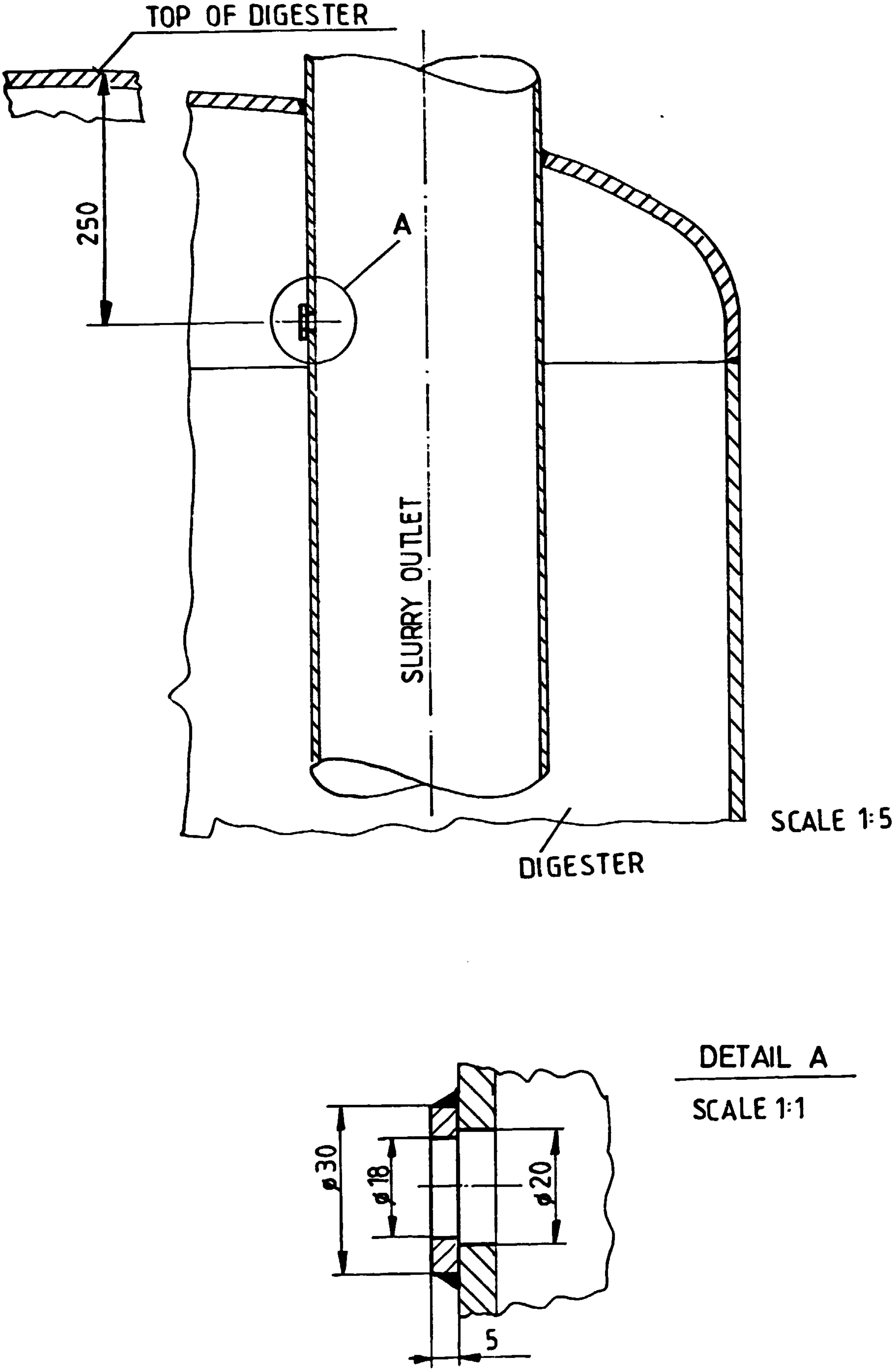

TOP OF DIGESTER
250
A
SLURRY OUTLET
SCALE 1:5
DIGESTER
DETAIL A
SCALE 1:1
ø30
ø18
ø20
5

Fig. 4.1-1

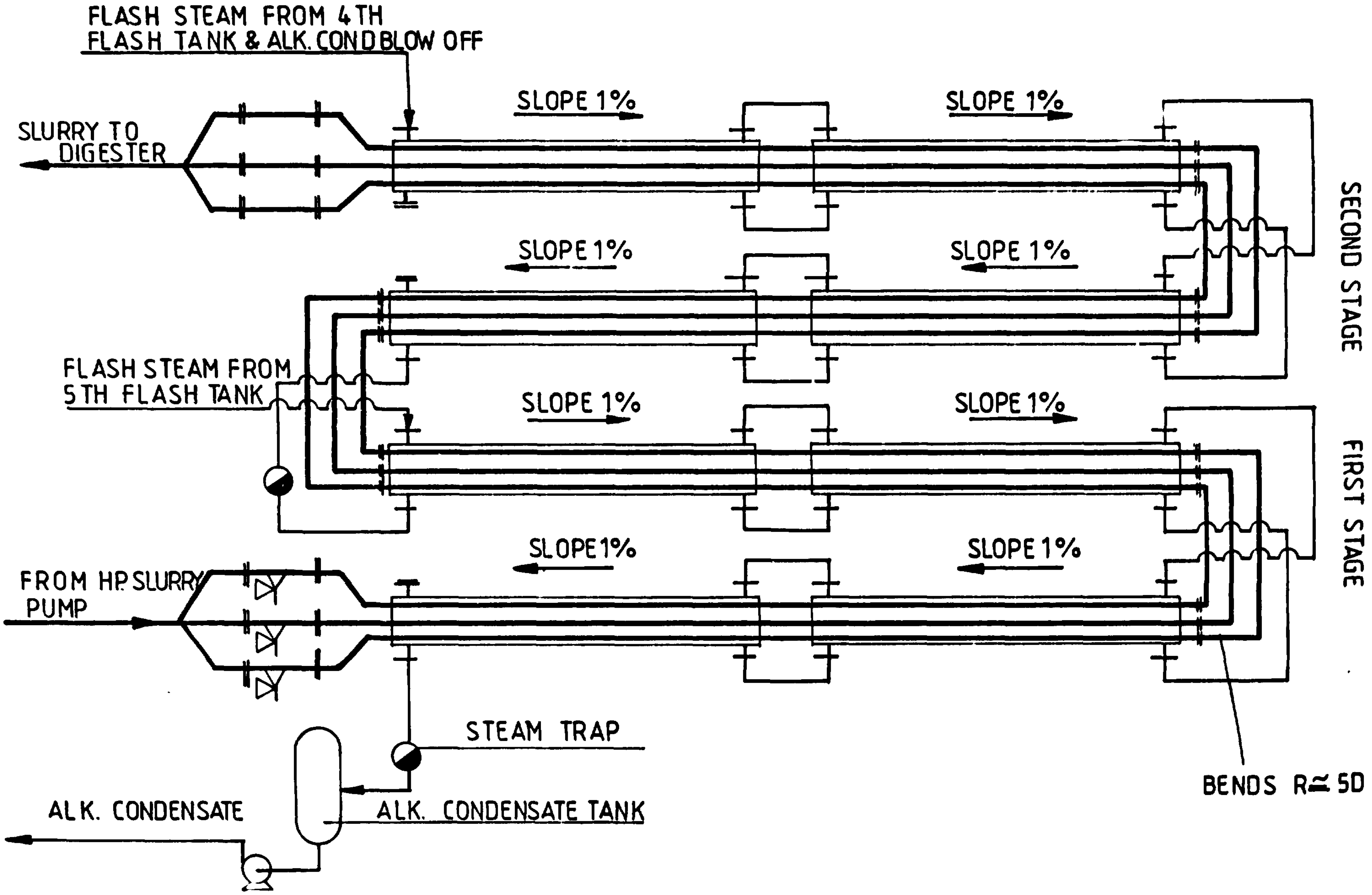

FLASH STEAM FROM 4TH FLASH TANK & ALK. COND BLOW OFF
SLOPE 1%
SLOPE 1%
SLURRY TO DIGESTER
SECOND STAGE
SLOPE 1%
SLOPE 1%
FLASH STEAM FROM 5TH FLASH TANK
SLOPE 1%
SLOPE 1%
FIRST STAGE
SLOPE 1%
SLOPE 1%
FROM HP SLURRY PUMP
STEAM TRAP
BENDS R≃5D
ALK. CONDENSATE
ALK. CONDENSATE TANK

Fig. 4.2-1

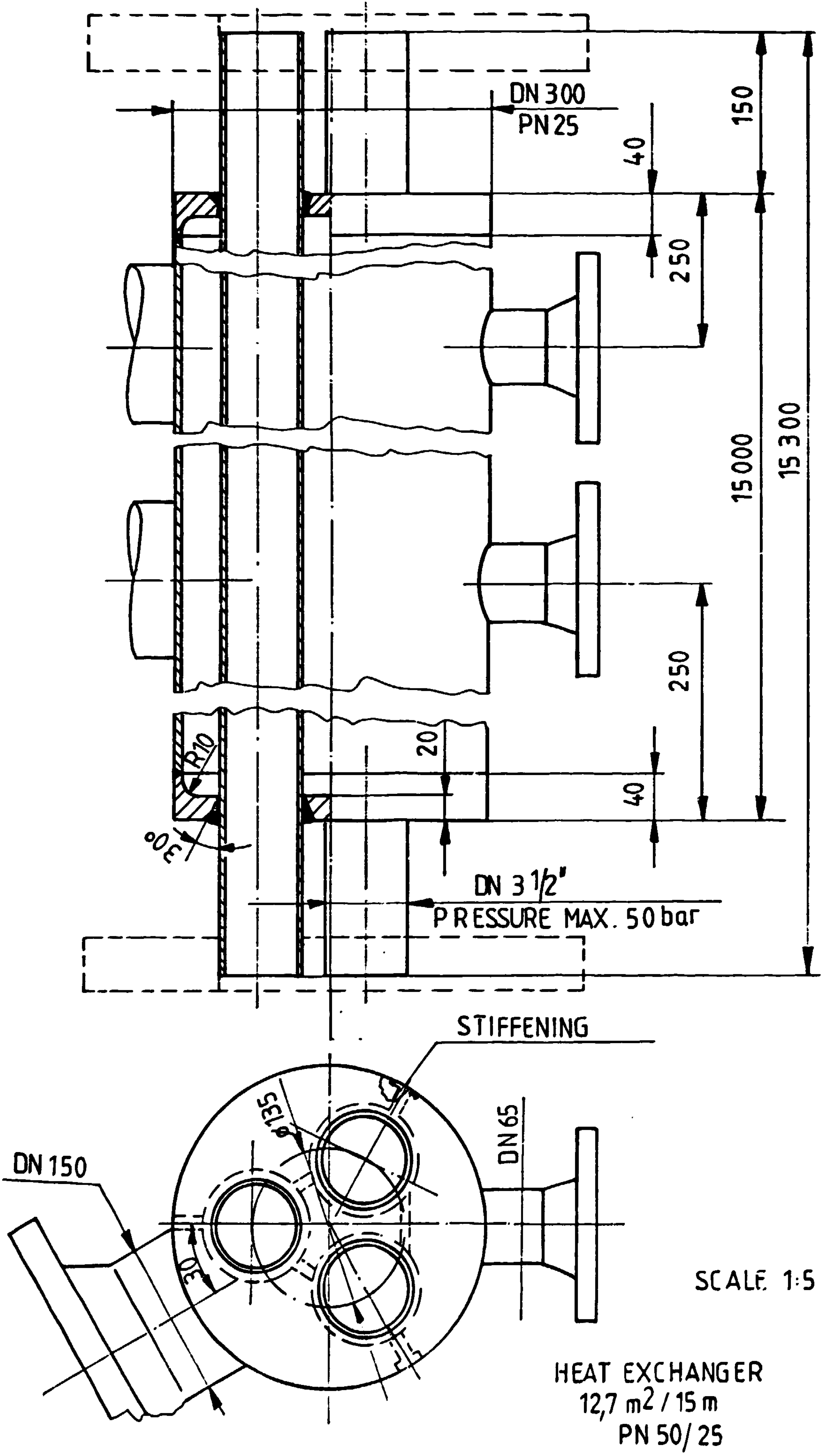
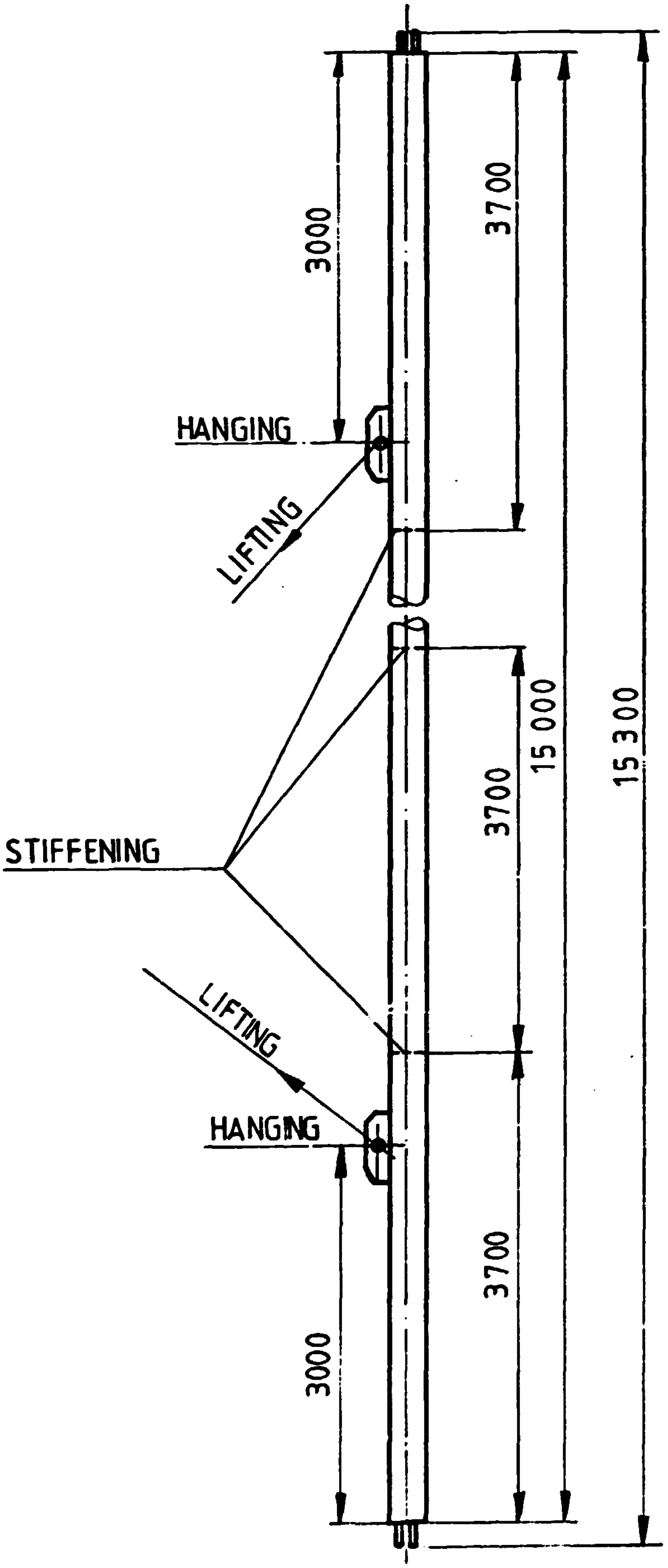
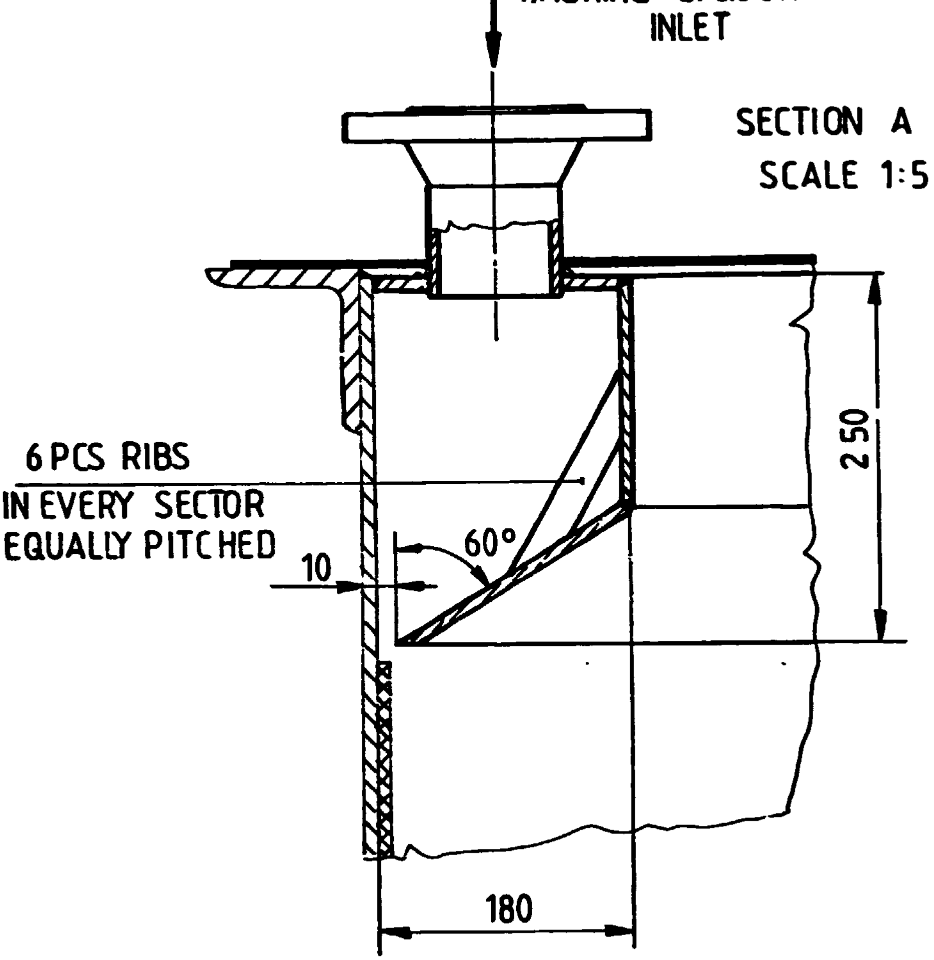

DN 300
PN 25
150
40
250
15 000
15 300
250
20
40
R10
30°
DN 3 1/2"
PRESSURE MAX. 50 bar
STIFFENING
ø135
DN 65
DN 150
30
SCALE 1:5
HEAT EXCHANGER
12,7 m2 / 15 m
PN 50/25

Fig. 4.2-2



SCALE 1:50

Fig. 4.2- 3


INLET
SECTION A
SCALE 1:5
6 PCS RIBS
IN EVERY SECTOR
EQUALLY PITCHED
60°
10
250
180

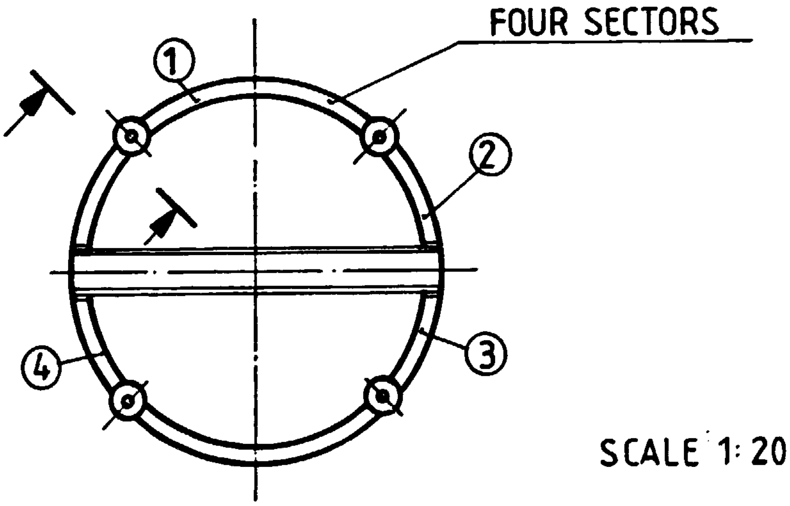
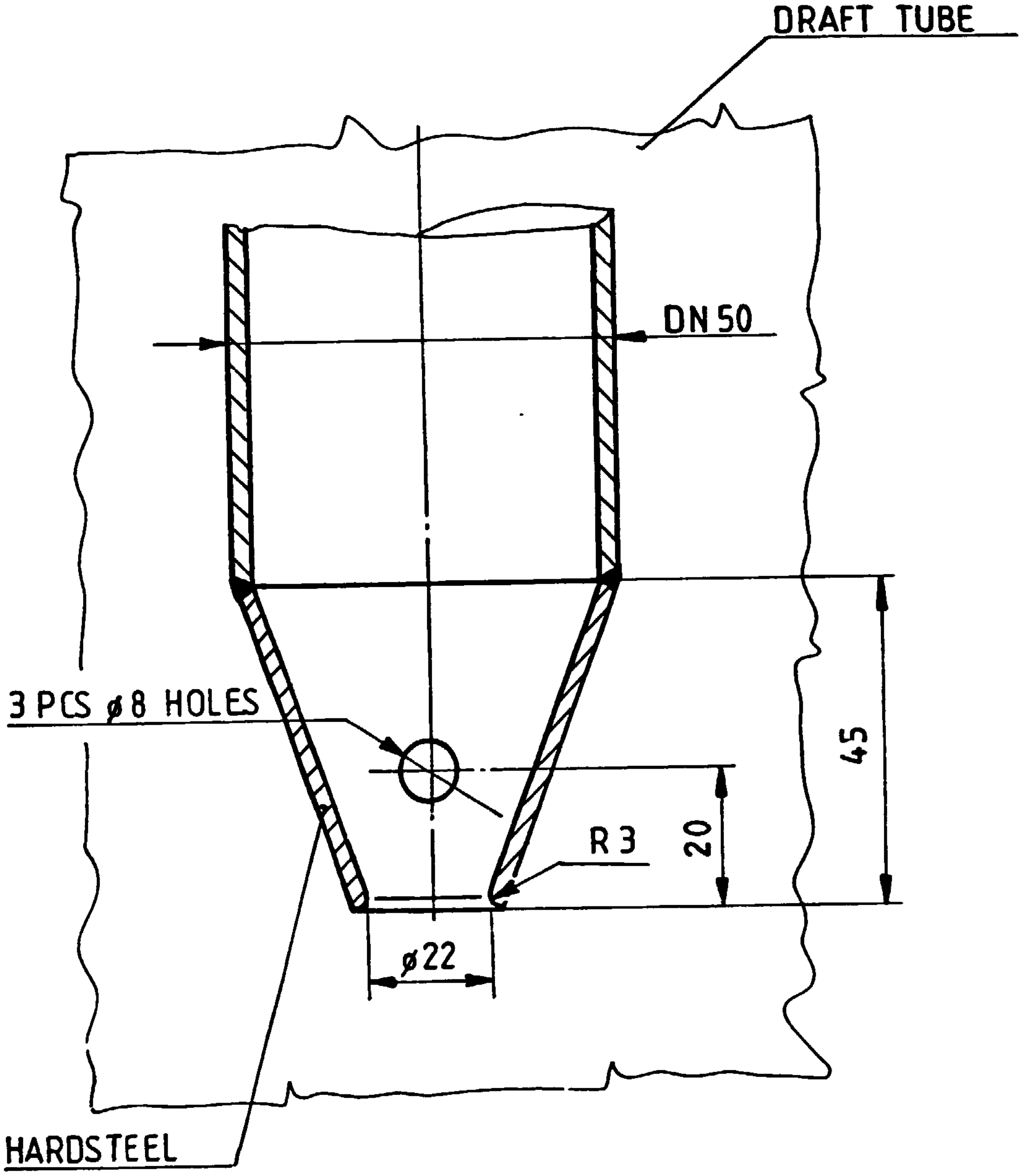

FOUR SECTORS
①
②
③
④
SCALE 1: 20

Fig.4.3-1



SCALE 1:10

Fig. 4.4-1

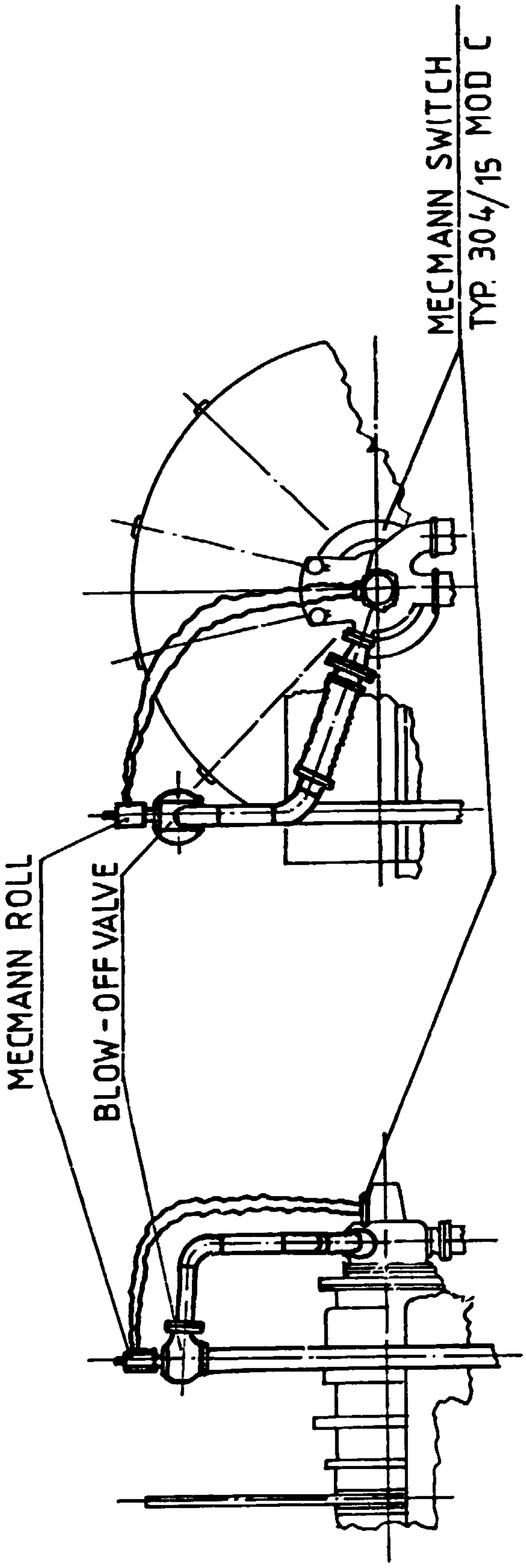
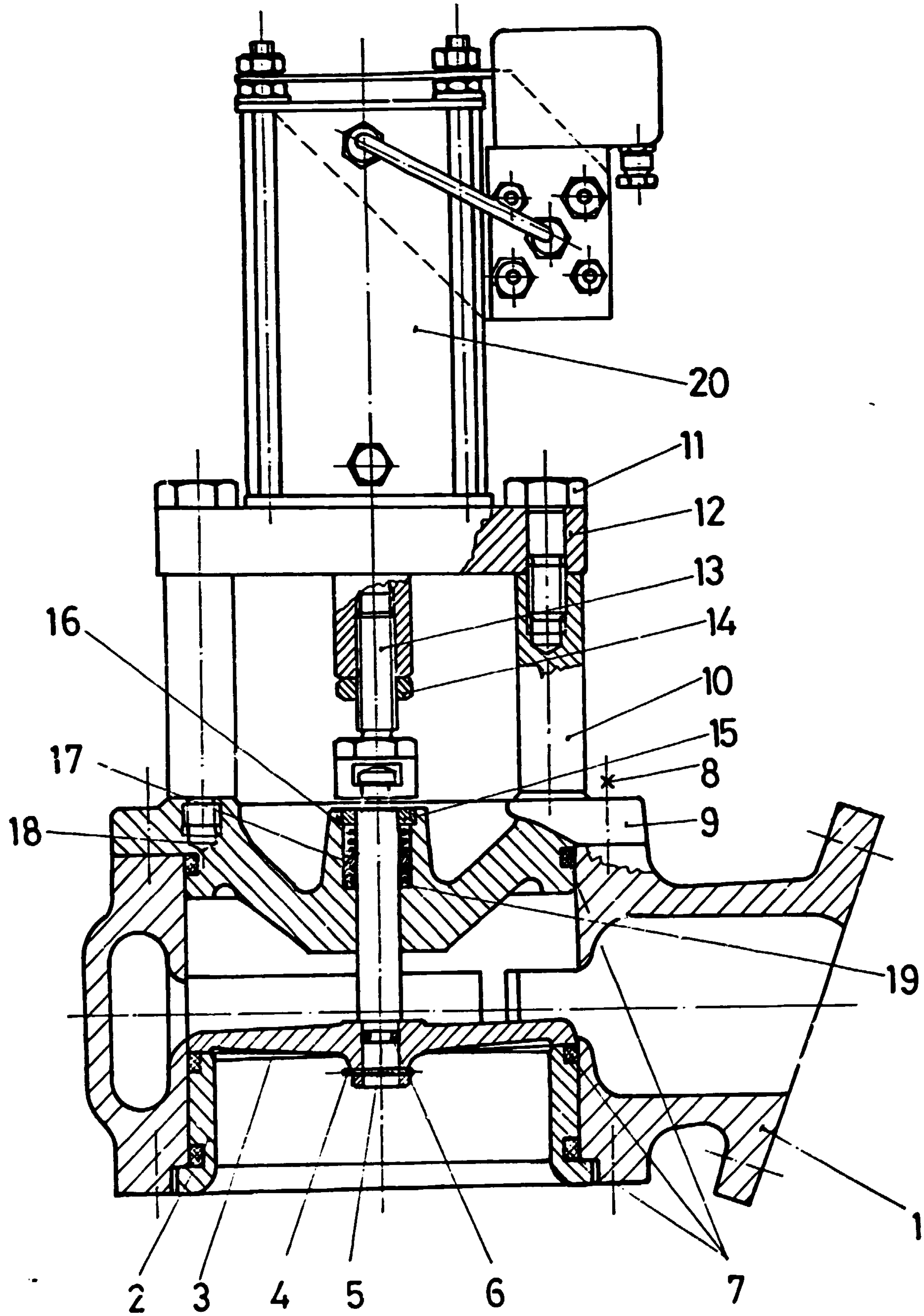


Fig. 4.5-1



ITEMS SEE ON PAGE No. 2/2

Fig. 4.5-2/1

1	VALVE BODY
2	SEAT
3	DISC
4	VALVE PLUG STEM
5	COTTER PIN
6	"O" RING
7	"O" RING
8	HEXAGON FULLY THREADED BOLT
9	COVER
10	POLE
11	HEXAGON BOLT
12	MECMANN CYLINDER FIXING
13	ROD HEAD
14	HEXAGON FLAT NUT
15	PACKING FIXING BOLT
16	PACKING EXPANDER
17	UPPER SUPPORTING RING
18	SEALING "V" RING
19	LOWER SUPPORTING RING
20	MECMANN CYLINDER

Fig. 4.5-2/2

5. Technical materials sent to BALCO

5.1 Start-up and operating procedure for continuous digester lines after maintenance

(According to the practice of Hungarian alumina plants)

After finishing the maintenance work but prior to filling up the continuous line the digesting system must be thoroughly checked once more. The functioning of fittings, safety valves, gaskets, rapture disks, contact manometers, other manometers, thermometers must be checked and - if required - the maintenance must be completed.

If the line is suitable for safe use and all the equipment necessary for its running are ready for start-up, the digesting line must be started in the following sequence:

5.1.1 Leakage test

Compressed air or steam is to be used for the test. The use of liquids is not allowed as they are incompressible and the draining of the system is also problematic.

The leakage test is to be done in several steps. First the preheaters and digesters are filled with compressed air or steam up to a pressure of 10 bar from the delivery pipe of the H.P. pump. The leakage of the system must be checked at this pressure. The test lasts for 20-25 minutes. Then assuring a free flow for the steam or compressed air to the flash tanks the flash tank line is to be pressurized at a pressure level of 3 to 4 bar. After the leakage test the pressure of the system is to be released through the last flash tank and through the pressure shock absorber of the H.P. pump.

5.1.2 Start-up with slurry

After setting the overpressure protection of the piston pump the delivery of the slurry can be started in such a way that the air can leave the system. The agitators of the digesters must be started. The whole digester line must be gradually filled up with slurry. After filling up the digesters the slurry pump must be stopped, the hand-operated valve located after the pump must be closed carefully.

According to the Hungarian practice there is a hand operated air purging valve located in the slurry line between the last digester and first flash tank. During the filling-up period this valve must be opened to allow the air/gas phase to leave the system. When the system is completely filled with slurry, the line pressure hand valve must be closed and the heating of the digesters can be started. The heating-up procedure is to be controlled cautiously in such a way that the pressure and temperature be raised uniformly up to 30 bar and 238 °C. Should the pressure of the digester line rise suddenly in spite of the slow heating, with the very cautions opening of the hand valve at the end of the line some slurry must be released to the flash tanks (15°-20° opening of the hand wheel of the hand valve accompanied by a soft hissing sound of the flowing slurry). If necessary, the heating should be reduced. For the time of slurry releasing both the hand valve and the automatic control valve (the latter switched to hand operation) can be opened very slowly. Then the control valve must be closed again. At this time all the vapour valves of the flash tanks are closed.

Subsequently the piston slurry pump and the pressure shock absorber should be pressurized and the pressure of the filled digester line must be released towards the flash tank line very slowly with the gradual opening of the hand valve at the end of the digester line. Then the piston slurry pump can be started with its lowest RPM and the line-pressure control valve can be opened very slowly. If the filling up of the digester line was not proper and a considerable amount of air/gas was left in the system, the control valve would have to handle slurry and gas phases alternately, which would make the spindle of the control valve vibrate and strong cracking sounds would be heard. This vibration can reach a level when the plug of the valve can break off. That is why the proper operation of the line pressure control valve has such a great importance, that is, no gas or air phase is allowed to remain in the system.

As long as the hand valve is not opened fully, the control valve is not allowed to be opened. When starting the slurry flow from the digesters towards the flash tanks the automatic control valve should be manually operated.

Following these the heating control loop must be switched from hand operation to automatic control in the control room.

If the temperature and pressure of the slurry attains the prescribed values, the RPM of the piston pump can slowly be raised. The slurry flow should be increased so that the digestion temperature and pressure would not drop.

The vapour discharge from the flash tanks must be started with the highest pressure tank and must be continued gradually towards the lowest pressure one paying attention to the pressure in the flash tanks. The vapour valves must be opened carefully to avoid the escape of the liquid phase together with the vapour.

The slurry level in the flash tanks must be set by manual control and only after balancing it can the controller be switched to automatic operation.

The flash tank pressures should be set to the following values in a 240 °C digestion:

1st flash tank: 16-20 bar
2nd flash tank: 10-14 bar
3rd flash tank: 8-9 bar
4th flash tank: 6-7 bar
5th flash tank: 5-6 bar
6th flash tank: 3.5-4.5 bar
7th flash tank: 2.5-3.5 bar
8th flash tank: 1-1.5 bar
9th flash tank: 0.2-0.5 bar

The quality of the vapour must be checked every 5 minutes as long as the equilibrium is reached following a start-up. Should the vapour be contaminated with slurry, the opening degree of the corresponding vapour valve must be reduced. Finally the vapour valve must be set in such a way that the vapour be pure and the automatic control loop be able to control the level in the flash tank.

The pressure gauge should be continuously and thoroughly checked during the whole filling up period!

5.1.3 Continuous operation of the digestion and flash tank line

The operation of the digester line must be checked with the pressure and temperature data of the individual vessels. Changes can happen in a short period of time, so the supervision of the line is essential. The observed data should be recorded in the shift log.

The level of slurry in the digesters is very important considering the residence time and the scaling of the heating coils (the scaling on the heating coils is more extensive, if the digester is not full of slurry). Generally the digesters are de-aerated through a hole bored into the inside section of the slurry transfer pipes.

5.1.4 De-aeration of the heating coils

The de-aeration of the heating system is necessary for maintaining the heating efficiency and for checking the operation of the flash tanks.

Before the de-aeration procedure the accident-free flow of the pure or contaminated vapour must be assured to the collector, then either the upper or the lower de-earator valve of the heating coil mus be carefully opened. Should the escaping vapour be contaminated with slurry, the procedure must be stopped and the vapour contamination must be eliminated by the careful partial closing of the corresponding flash tank's vapour valve. If the coil is de-aerated because of insufficient heating, the normal temperature values will be restored after finishing the de-aeration. Otherwise the de-aeration procedure must be repeated.

The digester pressure control valve must be controlled in such a way that in case of manual operation the pressure of the line remain constant.

The level control loop must control the level in the flash tank properly. Should the latter increase in spite of the control loop, the vapour valve must be slightly closed. If it sank. the vapour valve should be opened a bit further. If the level control loop of the first flash tank does not work properly, the level alarm of the loop means that the flash tank is filled up with slurry. If the level alarm did not work, the pressure measured on the flash tank or the contamination of the vapour leaving it would indicate its filled-up state. In this case the vapour valve of the first flash tank must be closed to a sufficient degree so as the proper slurry level in it be reached again.

The stuffing box of the digester agitator is cooled with industrial water. The efficiency of the cooling must be checked frequently to avoid overheating.

Two to four digesters are heated with high pressure (70 bar) steam. In normal operation there is an automatic temperature control in the function of the slurry temperature. In case of some disturbances this control loop must be switched to manual operation and according to the slurry temperature the steam control valve must be opened or closed.

If the feeding of the slurry stops due to some reason, the automatic temperature control loop will close the steam valve (loop in AUTO mode) so the pressure cannot increase above the allowed one.

5.2 Technical data on the slurry piston pumps with liquor flushing

The technical material containing drawings and descriptions was mailed to the Korba Alumina Plant.

5.3 Technical literature on instrumentation

The prospectuses were mailed to the Korba Alumina Plant.

5.4 Method for increasing the classification efficiency of conventional hydroseparators

The prospectus was handed over to the experts of the Korba Alumina Plant during their visit to the MOTIM Alumina Plant.

6. Economic effects of the proposed process modifications

The enclosed Table 6-1 includes the summarized investment costs of mechanical, civil and electric engineering, steel structures and instrumentation.

The economical issue of the proposed process modifications are the following respectively:

General modifications (improvement of the digestion yield, molar ratio, replacement of the LP and HP heat exchangers to tube in tube elements, improvement of the red mud causticization efficiency, improvement of the instrumentation of the plant, reduction of the non-technological water inlet) result the following change specific consumptions

Wet bauxit, t	-0.160
Caustic soda, t	-0.0143
Burnt lime (80 % CaO),t	+0.1032
Flour, t	-0.0001
Steam, t	-0.031
Fuel oil, t	0
Power, kWh	-15.5

An additional effect of these modifications is an increase in the plant capacity by 27 thousand ton alumina per year.

The material and energy costs decrease by 78.48 Rs/t alumina i.e. 14,666,500 Rs/year.

The sum of the investment costs are 13,100,000 Rs. (see Table 6-1, variants 3.1.1.1, 3.1.1.2-A, 3.3.2-1, 3.3.2-2)

A decrease in the steam consumption is a consequence of the replacement of the three recuperation digesters per line.

The specific material and energy consumptions change as it follows:

Wet bauxit, t	0
Caustic soda, t	-0.0004
Burnt lime, t	0
Flour, t	0
Steam, t	-0.639
Fuel oil, t	0
Power, kWh	0

Increase in the alumina production capacity is 1.26 thousand tons alumina per year.

The material and energy costs decrease by 66.4 Rs/t alumina, i.e. 12,491,800 Rs/year. The excess investment costs is 5,430,000 Rs.

Installation of the liquor purification results in further alumina production capacity increase by 5.95 thousand tons alumina per year.

The specific material and energy consumptions change as it follows:

Wet bauxit, t	-0.001
Caustic soda, t	+0.00018
Burnt lime, t	0
Flour, t	0
Steam, t	-0.118
Fuel oil, t	0
Power, kWh	-4.29

The material and energy costs decrease by 13.73 Rs/t alumina, i.e. 2,644,700 Rs/year.

The required additional investment cost is 2,730,000 Rs (see 3.1.2.1 variant in Table 6-1).

Modernization of the precipitation described in Chapter 3.1.2.4 makes it possible to increase the plant capacity by an additional 5.92 thousand tons alumina per year.

The expected changes in the specific consumptions after the modernization of the precipitation is the following:

Wet bauxit, t	-0.002
Caustic soda, t	+0.00029
Burnt lime, t	0
Flour, t	0
Steam, t	-0.092
Fuel oil, t	0
Power, kWh	-19.8

The material and energy costs decrease by 21.05 Rs/t alumina, i.e. 4,210,000 Rs/year. The required additional investment costs is 4,820,000 Rs.

5 kg fuel oil per ton alumina has been envisaged by using the DRYMAX dewatering additive in a dosage of 0.122 kg/t alumina. The material and energy costs decrease by 14.48 Rs/t alumina, i.e. 2,896,400 Rs/year. The required investment costs only 1,150 Rs.

Table 6-2 shows the common effects of the individual process modifications on the specific consumptions.

The following unit prices have been taken into account in the economic calculations.

Wet bauxit	193.72	Rs/t
Caustic soda	6255	Rs/t
Burnt lime	744	Rs/t
Flour	5070	Rs/t
Steam	100.03	Rs/t
Fuel oil	3483	Rs/t
Power	0.67	Rs/t
DRYMAX 934	1000 GBP/t i.e. 24000 Rs/t	

Table 6-1

Investment cost
in Rs

Plant unit	Civil engineering	Mechanical engineering	Steel structure	Instrumentation	Engineering costs	Component total	Remark
3.1.1.1	-	-	-	-	-	1,650,000	
3.1.1.2	200,000	5,960,000	520,000	650,000	730,000	8,060,000	A variant
	300,000	10,340,000	780,000	850,000	1,220,000	13,490,000	B variant
3.1.2.1	170,000	1,960,000	100,000	300,000	200,000	2,730,000	(Purification)
3.1.2.3	-	1,150	-	-	-	-	(Dry max)
3.1.2.4-1	460,000	2,650,000	180,000	1,100,000	430,000	4,820,000	1. variant
3.1.2.4-2	550,000	5,700,000	180,000	1,200,000	760,000	8,390,000	2. variant
3.1.2.5	400,000	10,400,000	200,000	300,000	-	11,300,000	(Heat exchangers)
3.3.2-1	560,000	100,000	50,000	800,000	120,000	1,630,000	2 pcs of new control rooms
3.3.2-2	200,000	80,000	150,000	1,200,000	130,000	1,760,000	Renovation of 3 pcs of control rooms

Table 6-2

SPECIFIC AND YEARLY MATERIAL AND ENERGY CONSUMPTIONS AND COSTS

	Units	Base 1987.February-November		Proposed		Differences of	
		specific	yearly	specific	yearly	specific consumption	specific costs in Rs
Alumina	t	-	159,900	-	200,000	-	-
Wet bauxite	t	2.759	441,160	2.596	519,300	-0.163	-31.58
Caustic soda	t	0.1085	17,350	0.09427	18,854	-0.01423	-89.01
Burnt lime	t	0.141*	22,545	0.2162**	43,240	+0.1032**	+76.78
Flour	t	0.0024	384	0.0023	460	-0.0001	- 0.51
H.P. steam	t	2.640	422,170	1.707	341,400	-0.933	-93.33
L.P.steam	t	0.910	145,550	0.963	192,600	+0.053	+ 5.30
Total steam	t	3.550	567,610	2.670	534,000	-0.880	-88.03
Fuel oil	t	0.124	19,830	0.124	24,800	0	0
Power	Mwh	0.330***	52,767	0.2904	58,080	-0.0396	-26.53
Total							-158.88
Yearly material and energy costs							31,776,000 Rs
Total investment costs							37,380,000 Rs

* in 64.2 % a.CaO
** in 80 % a.CaO
*** estimated