ENERGY CONSERVATION APPROACHES IN PULP AND PAPER INDUSTRY - CASE STUDIES

A DISSERTATION

submitted in partial fulfillment of the
requirements for the award of the degree

of

MASTER OF TECHNOLOGY
IN
PULP AND PAPER

By

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CANDIDATE'S DECLARATION

I hereby certify that the work which is being presented in this thesis entitled "**ENERGY CONSERVSATION APPROACHES IN PULP AND PAPER INDUSTRY - CASE STUDIES**" in fulfillment of the requirement for the award of the degree of Master of Technology and submitted at the Department of Paper Technology, Indian Institute of Technology Roorkee, Saharanpur Campus, a department of the Institute is an authentic record of my own work carried out during the period from August, 2003 to May, 2004 under the supervision of Dr. N J Rao, Shri Ram Kumar, and Shri V S S Kanna Babu.

The matter embodied in this thesis has not been submitted by me for the award of any degree of this or any other University.

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EXECUTIVE SUMMARY

- The energy conservation study is mainly targeted at identifying practical, sustainable and economically viable energy saving opportunities in the area of Pulp mill, Chemical recovery, and Co-generation unit, resulting from detailed study and analyzes of technical parameters. The energy conservation study is carried out using online process measurement data for evaluation of energy saving potential and economic viability.
- This report relates to the energy conservation study based on present process and future possible modifications, which could be implemented in ITC Paperboards and Specialty Paper Division, Unit Bhadrachalam.
- The plant utilizes electrical power through co-generation and state grid. Plant uses coal and black liquor as main fuel for generation of steam. The plant also consumes furnace oil in limekiln and SRB. The energy cost pattern of the unit is as follows:

Particulars	Annual consumption (2003- 04)	Unit Cost Rs.	Annual Cost Rs. Crore	% On Total Annual Cost.
Coal consumption	213,000 MT	1.5 / kg	31.95	83.42
DG power generation	1224,000 kWh	7.08 / kWh	0.86	2.24
Grid Power	15573,000 kWh	3.25 / kWh	5.06	13.21
Furnace oil	360 kl	12 / Lit	0.43	1.12

- During the study, the numbers of brainstorming sessions are carried out with the process personnel and with utility personnel and all the recommendations for improvement in energy saving possibilities have been discussed. The list of recommendations and energy saving is given as below:

Sl. No.	Recommendations	Steam TPH	Power, MW	Furnace Oil, kl/hr
1.	Two stage batch digester steaming	-	0.12	-
2.	Improvement of blow heat recovery system of pulp mill	3.84	-	-
3.	Improvement of steam economy in MEE system	1.51	-	-
4.	Controlling the crystallization of salts in black liquor and increasing the system availability	1.37	-	-
5.	Indirect heating of heavy black liquor up to firing temperature	1.19	-	-
6.	Effect of addition of neutralized sesqui sulphate crystal (i.e. Spent acid from ClO_2 plant) in black liquor as sodium sulphate and increasing in steam generation	0.56	-	-
7.	Optimization of cooling water requirement in surface condenser a. Savings with exact requirement pump capacity b. Saving with existing pump capacity c. Saving by avoiding orifice meter	 - - -	 0.076 0.01 0.032	 - - -
8.	Smelt dissolving tank vapor recovery for process condensate water heating	1.59	-	-
9.	Utilization of condensate available from lime kiln oil heater, SRP oil heater & soot blowing in SRB	0.41	-	-
10.	Boiler blow down and Air preheater condensate tank flash vapor recovery	1.6	-	-
11.	Reduction of furnace oil consumption in lime kiln section	8	-	0.02
12.	Improvement in power generation by modification of cogeneration system	-	3.09	-
13.	Enhancing energy economy with new technological options. a. RDH b. BL heat treatment	 - 3.82	 0.102 -	 - -
14.	**Grand total**	**23.89**	**3.43**	**0.02**

- The identified total annual energy saving potential is around 23.89 TPH of steam, 3.43 MW of power, and 0.02 kl/hr of furnace oil amounting to Rs. 1520 Lakh. The savings works out to be about 40% of the total annual energy cost of the plant.

- The digesters scheduling is analyzed to evaluate possibilities of increasing the productivity of digester. A 20% increase in pulp production capacity exists.
- Economic analysis of the study is done through pay back method by accounting the depreciation of 10 year and tax on investment of 15%. A summary list of recommendations, saving potential, implementation costs, and pay back period is given below:

Sl. No.	Recommendations	Savings Rs. Lakh	Cost of Implementation Rs. Lakh	Pay Back Period, months
Short Term Measures				
1.	Utilization of condensate available from lime kiln oil heater, SRP oil heater & soot blowing in SRB	10.58	Marginal	Nil
2.	Improvement of steam economy in MEE system	35.8	Marginal	Nil
3.	Controlling the crystallization of salts in black liquor and increasing the system availability	112	Nil	Nil
4.	Reduction of furnace oil consumption in lime kiln section	213	2	0.13
5.	Optimization of cooling water requirement in surface condenser	27.84	2.5	1.5
6.	Improvement of blow heat recovery system of pulp mill.	91.24	15	2
7.	Boiler blow down and Air preheater condensate tank flash vapor recovery	38.03	12.5	5
8.	Indirect heating of heavy black liquor up to firing temperature	29.2	12	6
9.	Smelt dissolving tank vapor recovery	37.87	20	7
10.	Effect of addition of sesqui sulphate solution (i.e. Spent acid from ClO_2 plant) in black liquor as sodium sulphate and increasing in steam generation	13.3	10	10
11.	Improvement in power generation by modification of cogeneration system	764.14	600	11
Long Term Measures				
12.	Two stage batch digester steaming	30	50	23
13.	Enhancing energy economy with new technological options. (i.e. RDH, Thermal depolymerization of black liquor)	116	NA	Long Term

- The recommendations of serial number 4,5, & 7 is already implemented successfully, serial number 1,2,3,6,10, & 12 have been started to implement in the unit and remaining will be implemented at the latest stage.

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ABBREVIATION

AA – Active Alkali
AOX – Absorbable Organic Halides.
APSEB – Andhra Pradesh State Electricity Board.
ADMTP – Air dry metric tonne pulp.
AMT – Ash Mixing Tank.
BL – Black Liquor.
BLS – Black Liquor Solids.
BOD – Biochemical Oxygen Demand.
CFB – Coal Fired Boiler.
CW – Cooling Water.
DCS – Direct Control System.
DS – Dissolved Solids.
ECF – Elemental Chlorine Free.
EMS – Environmental Management Systems.
EA – Effective Alkali.
FC – Foul Condensate.
FFFF – Free Flow Falling Film.
GCV – Gross Calorific Value.
GLC – Green Liquor Clarifier
GL – Green Liquor.
HP – High Pressure.
HBL – Heavy Black Liquor.
ITC – Indian Tobacco Company.
ISO – International Organization for Standards.
LHT – Liquor Heat Treatment.
LP – Low Pressure.
MP – Medium Pressure.
MEE – Multi Effect Evaporator.
PCFV – Process Condensate Flash Vessel.
PC – Primary Condenser.
PLFV – Product Liquor Flash Vessel.
RDH – Rapid Displacement Heating.
RAA – Residual Active Alkali.
RT – Retention Time.
SC – Secondary Condenser.
SRP – Soda Recovery Plant.
SRB – Soda Recovery Boiler.
SFT – Secondary Fiber Treatment
SDT – Smelt Dissolving Tank.
SCFV – Steam Condensate Flash Vessel.
TG – Turbo Generator.
TRS – Total Reduced Sulphur.
TS – Total Solids.
TAT – Time At Temperature.
TTT – Time To Temperature.
TPD – Tonne Per Day.
TPH – Tonne Per Hour.
TPY – Tonne Per Year.
WBL – Weak Black Liquor.

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PROPOSED STUDY

The power scenario in the country in the last ten years had been facing shortage of power in many regions through out the year. The shortage of power has a cascading effect on the industrial production and consequently the economic development of the country. Hence energy conservation is the foremost talk of the day and efforts are to bring down the power consumption in the view of present supply / demand position. Even though there is awareness is to use energy optimally, in industries. Barriers like, lack of priority, lack of finance, prevent work being done towards optimization of energy. The aim of this project is to study the potential for energy conservation measures in an integrated Pulp and Paper Industry without affecting production, yield or quality.

Attempts are made to systematically analyze the mill operations from technical aspects and came out with the specific suggestions for improving the performance of the mill with respect to energy use.

The proposed study is related to Pulp mill and Chemical recovery sections,

The methodology adopted for achieving the results, as follows:

- Understand the present plant operation and benchmark current operational levels.
- Compares the results to the original design and understand reasons for variance.
- Carry out engineering studies to determine how to overcome the limitations.
- Evaluates some of the options for improvement and workout its feasibility.

Engineering study are carried out in Pulp mill and Chemical recovery sections as,

- Effect of reducing the steam demand in digester house by two stage of digester steaming.
- Improvement of system performance of blow heat recovery system of digester house.
- Effect of increasing production capacity of digester by proper cooking schedule.
- Controlling the crystallization of salts in black liquor evaporation and increasing system availability.
- Effect of increasing steam economy of evaporation by feed of liquor to 4^{th} effect.
- Effect of increasing steam generation in recovery boiler by indirect heating of liquor to firing temperature and by adding spent acid as crystals from ClO_2 plant as make-up salt.
- Effect of reducing steam demand by recovering of smelt dissolving tank vapor.
- Effect of reducing steam demand by recovering boiler blow down and air preheater condensate tank flash vapor recovery.

- Effect of reducing the pumping energy by minimizing the cooling water flow to surface condenser.
- Effect of reducing furnace oil consumption in limekiln burner by using high intensity magnetic field.
- Increase in power generation by modification of cogeneration system.
- Enhance energy economy with new technological options.

CHAPTER 1

INTRODUCTION

1.1 GENERAL

Presently India rank amongst the top 3rd / 4th developing Nations in energy. Further Indian industrial growth is constant between 7 to 10% per year. The cost of energy in India presently ranks the highest amongst the developing countries.

India is poorly placed in terms of world energy resources, while 16% of world population lives in India, only 0.6% of oil and about the same portion of gas reserves exists in the country. India is endowed with 6% of coal reserves of the world. India is the net importer of energy.

The demand for electricity is increasing gradually year after year. The total installed capacity was around 1300 MW in 1947 and has made an impressive growth to about 90 GW during 2000. In spite of such an impressive growth, the per-capita electricity consumption in India is only 310 kWh, which are one tenth less than that of world average and one fortieth less that of Latin American countries. More over, presently country, as a whole, is facing a shortage of 18.3% in peak and 9.2% in meeting energy requirement. (3).

1.2. NECESSITY OF ENERGY CONSERVATION

Power is a basic input to the industry and forms a substantial part in the production cost. Hence it should be available continuously at a reasonable rate. It not only affects the profit of industry, but also will hit the economy of the country. Hence energy is crucial for every one.

The power shortage can be reduced by,

- Increasing the generation capacity.
- Renewable source of energy.
- Using available energy in the most efficient manner.

Increasing the generation capacity needs improved technology, additional cost and of course fuel resources where India lags a bit. Power generation through renewable has not yet attained the enough maturity. Hence the only way to save the world from the on looking darkness is to use the available resource in a most effective manner. Though energy conservation provides a temporary solution, it is the best solution as governed by the problem.

1.3. INDUSTRIAL SCENARIO

Large-scale industries form the backbone of the country's economic growth. These industries are highly energy intensive units and a energy survey would lead to a substantial potential savings. With the energy conservation bill coming into effect, energy conservation has becomes a mandatory for every industries through several industries, being aware about the importance and its advantages has adopted it on regular basis just like maintenance.

The energy consumption percentage of various industrial sectors is shown in **Table (1),**

Table (1): Percentage of energy consumed by different industries (3)

INDUSTRY	PERCENTAGE CONSUMPTION
Textiles	1.4
Fabricated Metals	1.5
Transportation Equipment	1.6
Lumber and Wood	2.2
Glass, Stone and Clay	4.3
Food	4.7
Iron and steel	12.2
Paper	12.4
Pharmaceutical and Chemical	24.9
Petroleum and Coal	29.5
Others	5.3

1.4. OBJECTIVE

The main objective of the project study is outline a systematic approach for substantial reduction in energy consumption in an integrated pulp and paper industry. The method of approach of this are given below:

- Analysis on energy consumption.
- Analyzing the present process area and finding the problems.
- Identify the potential energy saving area.
- Evaluate the energy saving measure both technically and economically.

CHAPTER 2
LITERATURE SURVEY

2.1. Two stage batch digester steaming (1)

Today's high-energy costs necessitate the optimum production and utilization of process steam. As more pulp and paper mills practice energy conservation, they will be saving steam through methods such as recycling hot air or water for heat recovery. They will also be looking for effective ways to balance their utilization of all sources of steam, including low-pressure steam (4 bar or less).

Co-generation, the coincidental generation of electrical power in conjunction with the production of process heat, has become a key word in most energy conservation programs. Mills will work to boost cogeneration as an alternative to increasing process of purchased electrical power.

Two-stage steaming has been in use for many years for indirect heating of digesters with forced circulation systems. In two stage steaming, the digester is initially heated and pressurized by low-pressure steam. After the digester has been partially brought to cooking pressure and temperature, it is switched over to 10.5 bar steam, which takes the digester to full pressure 7-8 bars and completes the cooking cycle.

By increasing demand for low-pressure steam, this method allows a change in the extraction ratios of mill's turbine generators. The amount of 4 bar extraction may be increased while 10.5 bar extraction is decreased. This measure would boost the turbine's electrical output while using the same amount of input steam. Saving would be realized through the replacement of purchased electricity with less expensive co generated power.

By going for two stage of steaming, this method will increase the cooking time of digester, which intern reduces the productivity of the digester. This disadvantage can come out with following approaches along with two stage of steaming,

- Presteaming of chip with steam packers – Improves bulk density of digester (Higher chips charge) with marginal drop in bath ratio and decrease in specific steam consumption.
- Impregnation of chips with black liquor and cooking up to first stage of steaming and displaced with fresh white liquor,
    - Increases rate of delignification, results in higher yield.

- Spilt alkali dosage with two stage of steaming, increase yield of pulp (i.e. increase in productivity) with improved alkali profile.

➤ Case study of mill with two stages steaming in digester as follows (1):

The potential benefit of adding two stage indirect heating to mill with a computerized digester system.

Production: 1000 a.d. TPD of low yield pulp

Steam usage: 1.27 tonne / tonne a.d. pulp

Steam and electrical generation factors:

Primary steam at 45 bar

70 kWh / tonne of steam at 10.5 bar extraction

110 kWh / tonne of steam at 4 bar exhaust

In this mill, 50% of the 10.5 bar steam could be replaced with low pressure, 4 bar steam. Cogeneration would be boosted 28.6% because the extraction of a turbine would be varied to meet the demand for this 4 bar steam, and an extra 28000 kWh would be generated. Depending on a mill's electrical generation system and the cost of purchased power, the increase in cogeneration could provide a considerable savings. The monthly saving of the mills as,

Net saving = Rs. 11.34 lakh per month

2.2. Displacement batch cooking - RDH (4)

Liquor displacement batch cooking is essentially a batch cook where heat and residual chemicals remaining in the black liquor at the end of cooking are captured for reuse in subsequent batch cooks. This occurs by placing the black liquor in separate pressure accumulators for later use to heat chips and white liquor for later cooks.

Benefits by RDH process,

➤ It saves thermal energy.

➤ It saves chemicals in cooking, washing and bleaching operations.

➤ It produces pulp, which has 15-20% higher tear-tensile strength. Hence better machine runnability is assured due to excellent quality of pulp.

➤ Technology is environmental friendly,

a. Less TRS emissions

b. No-emissions to mercaptan, malodorous gases.

c. Drastic reduction in AOX, BOD & Color.

➤ One stage of in digester helps in minimizing alkali losses.

- RDH pulp results in lower viscosity of black liquor going to chemical recovery.

Comparison of heat consumption between different pulping processes (6),

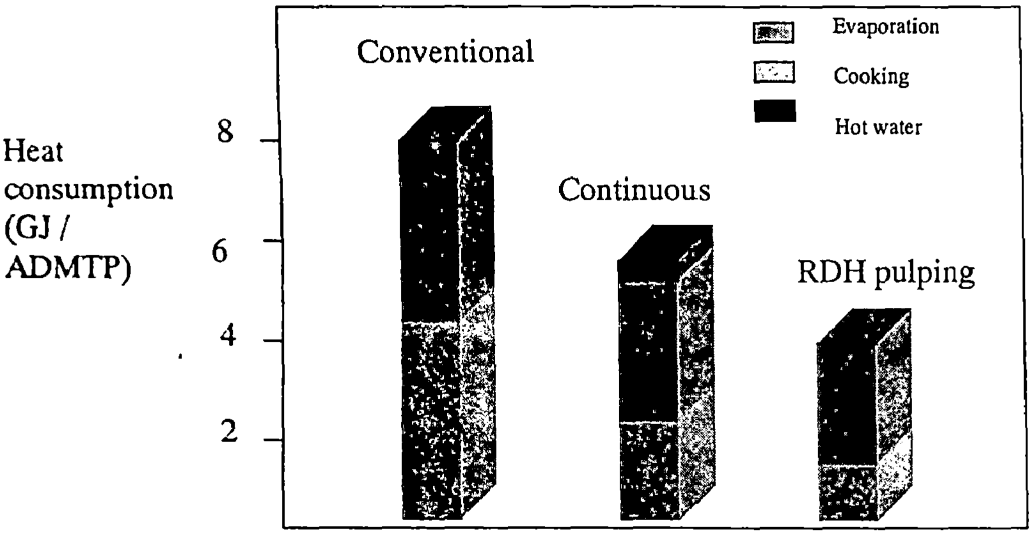


Figure (1): Energy economy comparison with RDH, Conventional batch, & Continuous pulping process.

2.2.1. Pulping process with liquor displacement

In the conventional system both white liquor and black liquor were charged together as soon as chip fill was over. Due to that % of OH^- ions at the beginning was very high compared to SH^- ions, as result of which cellulose was getting degraded and strength was going down. Is the reason why is not able to cook below Kappa no. 25.

In the present system of RDH initially charging was carried by warm black liquor and little amount of white liquor to give more presence to SH^- ions compared to OH^- due to its selectivity towards lignin. As a result of this cooking can be below Kappa no. 20. At the same time achieving the high strength properties compared to the conventional pulp.

Due to the desired properties of high strength and low Kappa number, is able to bring down the bleach demand drastically at the same time maintaining high brightness as well as high strength properties. As a result very low quantity of effluent generation results.

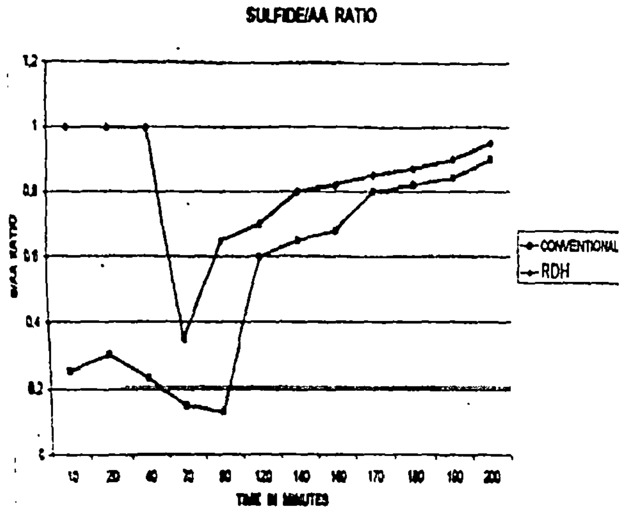


Figure (2): Effect of Sulfide / AA Ratio during cooking time by conventional and RDH process (4).

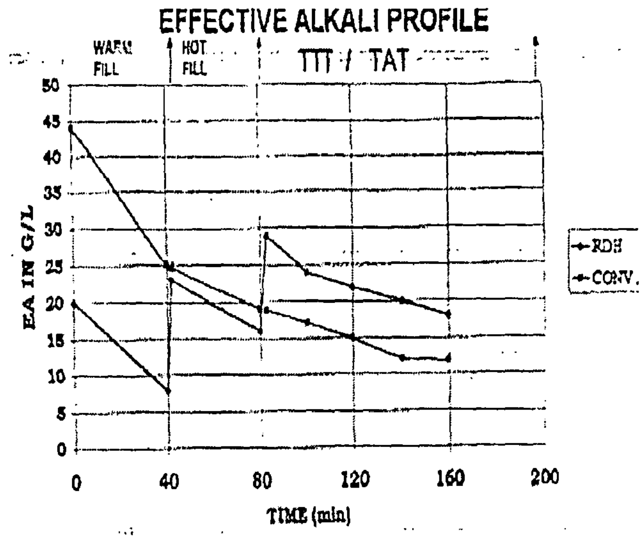


Figure (3): Effect of Alkali profile during cooking time by conventional & RDH process (4).

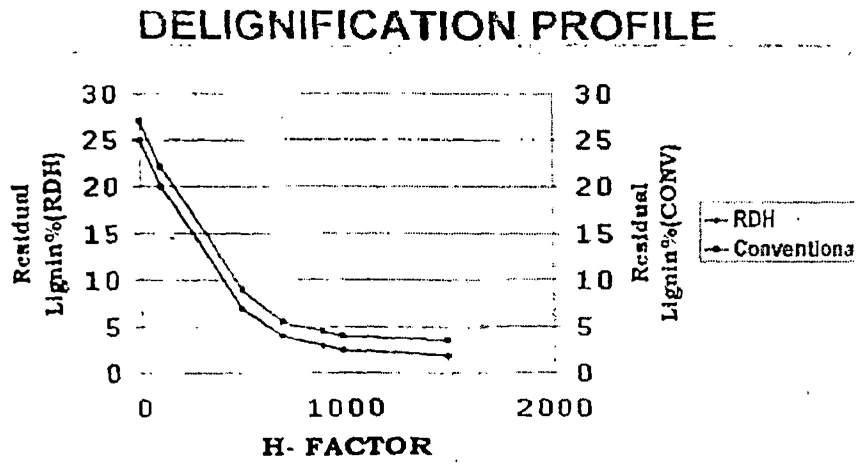


Figure (4): Effect of delignification profile by conventional & RDH process (4).

2.2.2. Cooking schedule of the RDH process,

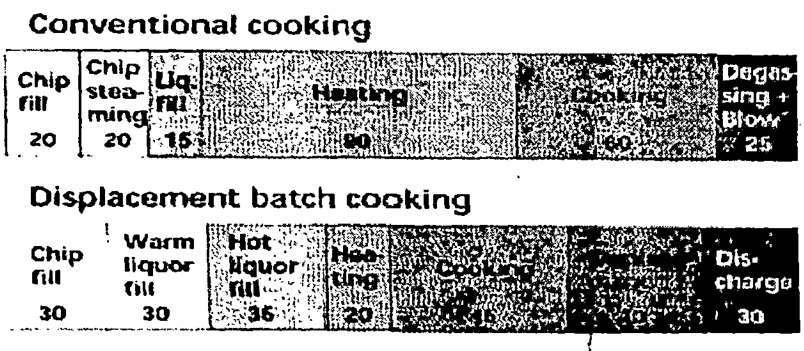


Figure (5): Cooking schedule of RDH process (4).

2.2.3. Uniformity and strength of pulp by RDH (4),

With the laboratory studies made by Sunds Defibrator, the results shown that full liquor displacement batch cooking produced more uniformly delignified chips inside the digester than conventional batch digester.

To minimize the Kappa number deviations between individual pulp fibers, delignification within chips must also be uniform. Figure (7) shows that liquor displacement cooking is better than conventional cooking. In conventional cooking, thicker chips therefore result in higher Kappa numbers and rejects. The Kappa number deviation inside the digester is also considerably higher in conventional batch cooking. The reason for better uniformity in liquor displacement batch cooking are the improved impregnation and chemical and heat distribution in the digester.

With referred to Sund Defibrator laboratory trial, the overall tear tensile strength of mill made, unbleached softwood Kraft pulp averaged only 75% of that of laboratory cooked reference pulp from the same chips. The unit operation responsible for this strength deficit in the conventional industrial scale digester. The hot discharge of batch digesters weakens Kraft pulps severely. The strength deficit is therefore due to a combination of two factors: cooking environment and discharge methods. The fibers undergo mechanical stress, and the pulp strength decreases. Where displacement cooking is of cold blow, delivers better pulp strength properties than conventional batch cooking. Figure (6) shows the effect of batch cooking method on strength delivery.

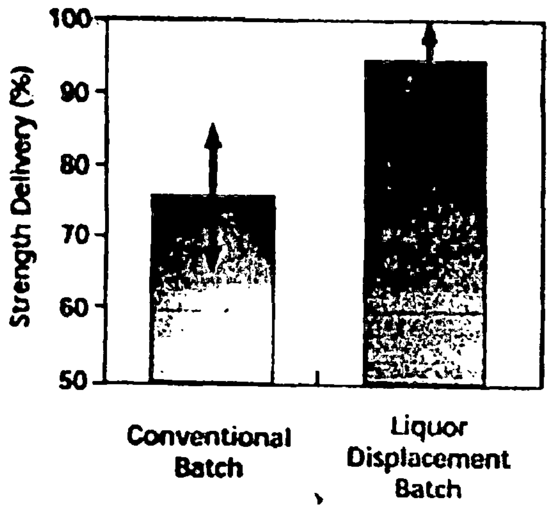


Figure (6) Strength delivery of pulp from conventional and displacement batch digesters

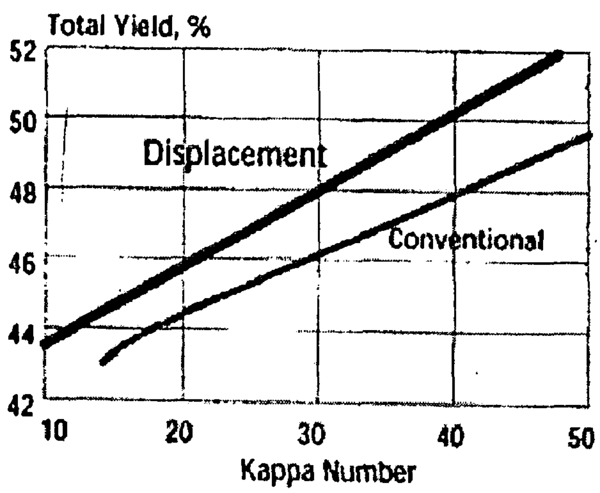


Figure (7) Yield Vs Kappa number for conventional and liquor displacement batch cooking of softwood Kraft pulps.

Where due to the uniform cooking and modified cooking chemistry, liquor displacement cooking system improve yield of pulping as shown in **Figure** (7).

2.2.4. An overview comparison of RDH system with Conventional system

Table (2): Comparative study of RDH Vs Conventional batch cooking processes

RDH Cooking System	Conventional Cooking system
▫ Initially more black liquor and less white liquor charge, and high at the end of hot fill.	▫ Both black liquor and white liquor charged together.
▫ Preference given to SH^- ions	▫ No preference given to SH^- as both black liquor and white liquor charged together.
▫ Pulp strength is high due to SH^- ions impregnation.	▫ Pulp strength is low compared to RDH process because of stated reason as above.
▫ Cold blowing at the low temperature results in minimum fiber damage. Reduced sulphur emission and no need of blow heat recovery system.	▫ Hot blowing results in fiber damage mechanically.
▫ Uniform cooking as the high bath ratio keeps liquor circulation rate constant. Due to good heat and mass transfer provided by displacement. (Bath ratio of 4:1).	▫ Non-uniformity of pulp quality.
▫ Low dilution factor and high black liquor concentration.	▫ High dilution factor and low black liquor concentration.
▫ Strength delivery ratio is more uniform and higher	▫ Strength delivery ratio is less than RDH cooking.
▫ Kappa number variation is ±0.5	▫ Kappa number variation is high.

2.3. Technique for controlling water-soluble scales (9)

In the Kraft pulping process. The evaporator, which concentrate the weak black from 15% to 65- 80% total solids for firing in the recovery boiler, routinely foul with inorganic and /or organic scale on heat transfer surfaces. This is a widespread problem that results in lost capacity and lost time for scale removal. It can also increase the overall evaporation and/or environmental load, limit pulp mill production, and reduce the steam supplied by the recovery boiler. If the mill's evaporation capacity is already limited, these impacts could be immediate.

In response to the industry need, the evaporator performance audit to help mill evaluate the nature and extent of their evaporator fouling problems and recommend a strategy to address and minimize them.

One common type of scale is soluble scale, which consists of water-soluble Na_2CO_3 and Na_2SO_4 in the form of burkeite ($2Na_2SO_4$:Na_2CO_3) and mixture of burkeite and Na_2CO_3. The other is insoluble or "hard" scale, which is $CaCO_3$, and mixtures of soluble scale, $CaCO_3$, and fiber. Soluble scale can be removed by boiling out the evaporator with water or weak black liquor, but hard scale must be removed by acid cleaning or hydro blasting.

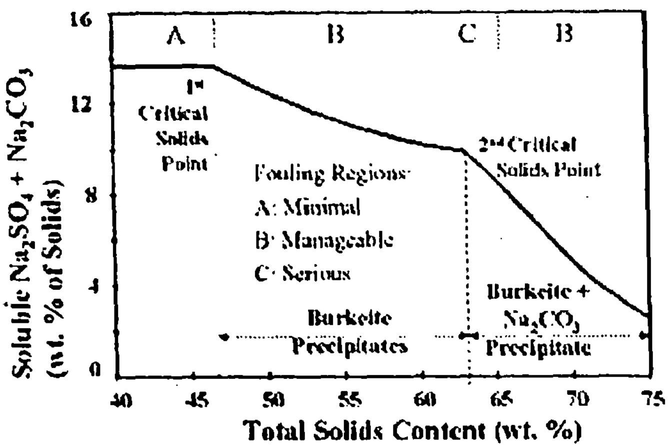


Figure (8): Simulated solubility behavior of Na_2SO_4 + Na_2CO_3 in black liquor.

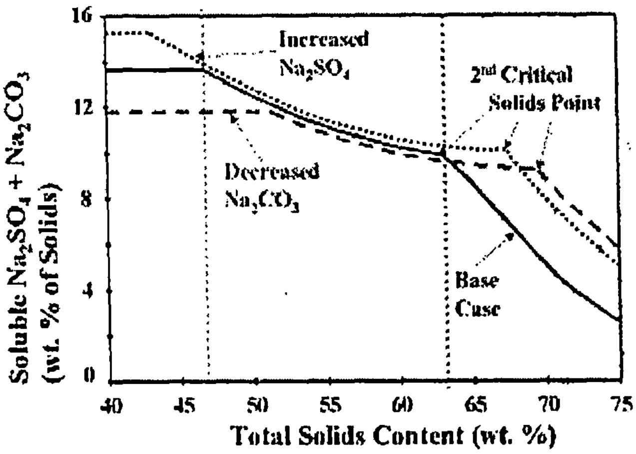


Figure (9): Critical solids points can be Controlled by varying the black liquor Chemistry (9).

2.3.1. Solubility behavior of Na_2CO_3 + Na_2SO_4 in black liquor (10)

Fouling occurs when Na_2SO_4 and Na_2CO_3 precipitate from solution as illustrated in **Figure (8)**. The two critical solids points occur as the liquor solids content is raised. The 1st critical solids point occurs when the liquor becomes saturated with respect to Na_2SO_4 + Na_2CO_3 and burkeite starts to precipitate. At the 2nd critical solids point, Na_2SO_4 has been depleted and the solution becomes saturated with the remaining Na_2CO_3, and Na_2CO_3 precipitates along with any remaining burkeite. The steeper slope above the 2nd critical solids point indicates that the extra Na_2CO_3 precipitate at a significantly higher rate than the burkeite did above the 2nd critical solids point. This increase in precipitation rate is an important characteristic of the 2nd critical solids point.

The observed evaporator fouling behavior can be described in terms of specific fouling regions that have been identified in **Figure (8)**. Fouling is "minimal" in region A. in region B, where burkeite is gradually precipitating; the fouling is "manageable". The fouling is "serious" in region C, a small solids range around the 2nd critical solids point, due to the sudden increase in precipitation rate. Finally, above the 2nd critical solids point, the fouling behavior once again becomes "manageable" (region B) as the additional precipitate has had the opportunity to grow on existing crystals.

The relative composition of Na_2SO_4 and Na_2CO_3 in the black liquor determines the locations of the 1st and 2nd critical solids points. And these can be moved as illustrated in **Figure (9)**. Adding Na_2SO_4 to the system (e.g., makeup salt cake) will increase the 2nd critical solids point. This is a desirable effect. However, because of the extra salt in the system, the 1st critical solids point will decrease, which is not desirable. To prevent this decrease in 1st critical solids, sulfate should be added after this point. Reducing the Na_2CO_3 content (e.g., by improving the causticizing conversion) will increase both critical solids points, which is desirable.

The precipitation rate is higher just above the 2nd critical solids point than just below it, and the 2nd critical solids point can be controlled by adjusting the Na_2CO_3 and/or Na_2SO_4 content of the liquor. The 2nd critical solids point also correlates with soluble-scale problems in falling-film evaporators.

The solubility of the $2Na_2SO_4$, Na_2CO_3 is independent of temperature in range of 100-140^{0}C and is assumed to be unaffected by liquor organics. The major components affecting burkeite solubility are,

- Effective sodium
- Na_2CO_3 + Na_2SO_4 mass-concentration ratio
- Black liquor solids content.

To develop mathematical equation (13) to predict BLS from a known Na_2CO_3 + Na_2SO_4 weight % on solids and known level of effective sodium. Both values can be obtained from a WL analysis of Na_2CO_3, Na_2SO_4, Na_2S, and NaOH.

Effective sodium is defined as,

➢ Effective sodium, g Na/L = Total sodium – 23(Na_2CO_3/53 + Na_2SO_4/71) ------------ {1}

The general solubility relationship at a Na_2CO_3/Na_2SO_4 ratio of 80:20, from data developed by T.M. Grace is,

➢ $BLS = 85.47 - 1.47Z - 0.95W$ -------------- {2}

The correction factors for the other Na_2CO_3/Na_2SO_4 ratio, from the data developed by T.M. Grace is,

➢ $f = r(5 * 10^{-5} x^2 - 7.7*10^{-4}x - 2*10^{-5}) - (1.013*10^{-3}x^2 + 0.01568x + 0.9993)$ -------------- {3}

The relations are referred to the following data developed by T.M. Grace (13),

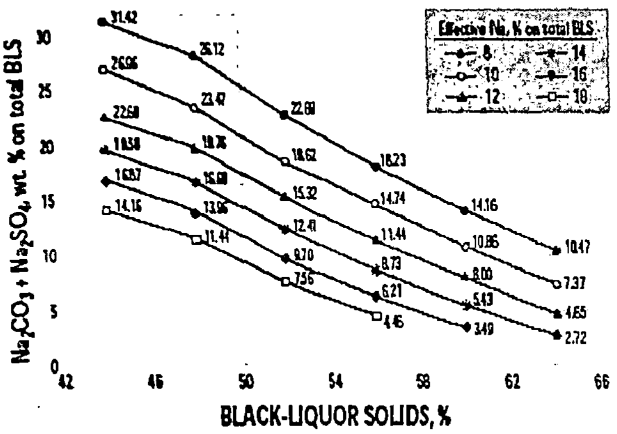


Figure (10) Observed solubility of burkeite in black liquor (at Na_2CO_3 / Na_2SO_4 ratio of 80:20) at six levels of effective sodium

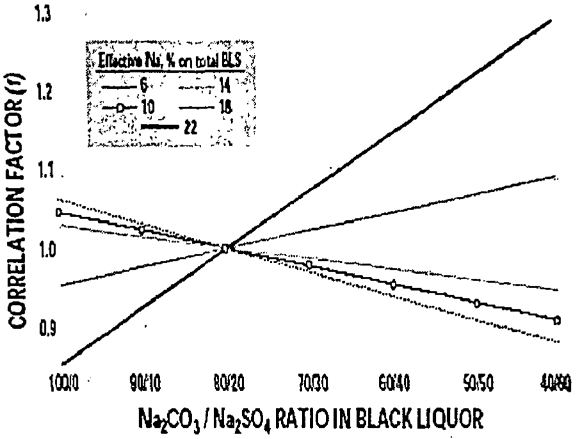


Figure (11) Correction factor for determine solubility of burkeite in black liquor (relative to Na_2CO_3 / Na_2SO_4 ratio of 80:20) at five level of effective sodium)

Finally, the critical BLS- the black liquor solids content where burkeite precipitation will begin- can be calculated from following equation using the correction factor (eq., 3) and liquor solids content (eq., 2),

➢ Critical BLS, % = [f(BLS/100)]/{1-[(1-f)(BLS/100)]} --------- {4}

Where,

Z – Effective sodium, % on total BLS

W – Na_2CO_3 + Na_2SO_4, % wt on total BLS

BLS – Black liquor solids, %

f – Correction factor

r – Percent Na_2SO_4 from ratio of Na_2CO_3 / Na_2SO_4.

2.3.2. Factors in the Kraft recovery cycle affecting burkeite solubility

There are several factors in the Kraft recovery cycle that influence the solubility of burkeite in black liquor:

- Effective-alkali/wood ratio
- Causticizing efficiency
- Spent-acid addition
- Makeup chemical addition.

2.3.3. Implications to scale control (10)

In practice there are two ways of doing,

- One is to minimizing the Na_2CO_3 in the soda cycle in the soda cycle by maintaining a high recuast conversion. And adding Na_2SO_4 to the black liquor, as salt cake or pent acid. Maintaining a high recaust conversion is preferred because it decreases the amount of material that can deposit as scale, and it leaves the possibility open for using the second option as well. Adding Na_2SO_4 in combination with poor recaust conversion can decrease the solubility limit of burkeite.

- Second is to maintain constant EA / Wood ratio in pulp mill, change in EA / Wood will change the critical solid point of black liquor.

2.4. Operating experience with indirect heating of strong black liquor (7)

The use of indirect heating of strong black liquor to achieve desired firing temperatures has two beneficial effects. Safer firing conditions results because dilution of solids concentrations is eliminated as experienced when using direct steam injection, and, overall recovery unit thermal efficiency is increased due to firing the higher solids concentration.

2.4.1. Potential benefits of indirect heating

The normal preparation of strong black liquor for firing in the Kraft recovery furnace includes: 1) evaporation of the liquor to an acceptable safe solids concentration; 2) mixing of salt cake (makeup and/or recovered chemical from the flue gas); and 3) heating the liquor to the boiler manufacturers recommended firing temperature. The heating to proper firing temperature has been obtained by mixing steam with the liquor immediately prior to firing. This method presents two drawbacks.

First, the solids concentration of the liquor, is diluted from 1 to 2 percentage points depending on the manufacturer firing temperature requirements by the condensed steam. This dilution may be detrimental to maintaining safe firing conditions in a Kraft recovery furnace, where low solids concentration can lead to blackouts or smelt/water explosions.

The second drawback is that the additional water content must be evaporated during the combustion process, and therefore decreases the net efficiency to steam. This, in turn, reduces the thermal efficiency at a time when energy conservation is of major importance.

Indirect heating of the liquor eliminates the solids dilution occurring with direct steam mixing. Thus, safer firing conditions are obtained. Indirect heating, by eliminating solids dilution due to the added condensed steam, also provides for higher thermal efficiency. A one-percentage point drop in solids concentration adds over 70 kJ/kg solids fired loss to the flue gas stream (assuming a 177^0C flue gas exit temperature). With indirect heating, condensate recovery to the feed water system can add even more improvement.

- A case study of mill (Weyerhaeuser Company, Tacoma, Washington), its performance by going for indirect heating of black liquor before firing as follows (7):

- Comparison of heat requirements for indirect and direct steam heating of strong black liquor. Reference temperature = 26.7^{0}C. Indirect heating uses 2.4 bar steam. Direct heating uses 9.66 bar steam. Condensate returns conditions — atm. pressure/83.3^{0}C. Liquor conditions entering heater - *63%* solids @93^{0}C and 0.70 specific heat.

Table (3): Comparison of heat requirement for indirect and direct steam heating of strong black liquor

Particulars	Indirect Heating kJ/kg solids	Direct Heating, kJ/kg solids
Evaporation Requirements (Assume Same)	-	-
Heating Requirements (93^{0}C – 107^{0}C)	74.09	74.09
Added Loss to Flue Gas (177^{0}C)		
□ Sensible	0.00	7.85
□ Latent	0.00	67.17
Condensate Return Credit	6.59	0.00
Total	67.50	149.11

System arrangement of the Mills as,

The indirect heating system consists of three parallel shell and tube heat exchangers (Model 22-16) provided by B&W through the Patterson-Kelly Company, Inc. The exchangers are of fixed tube sheet design, and each contains 71.7 square meters of surface area. The tubes and tube sheets are 304 SS; the shells carbon steel. The tube side design consists of four passes, while the shell side is one pass. The tubes are 25.4 mm OD * 16 BWG. The system is designed such that two exchangers are in service, while the third is washed and put on stand-by.

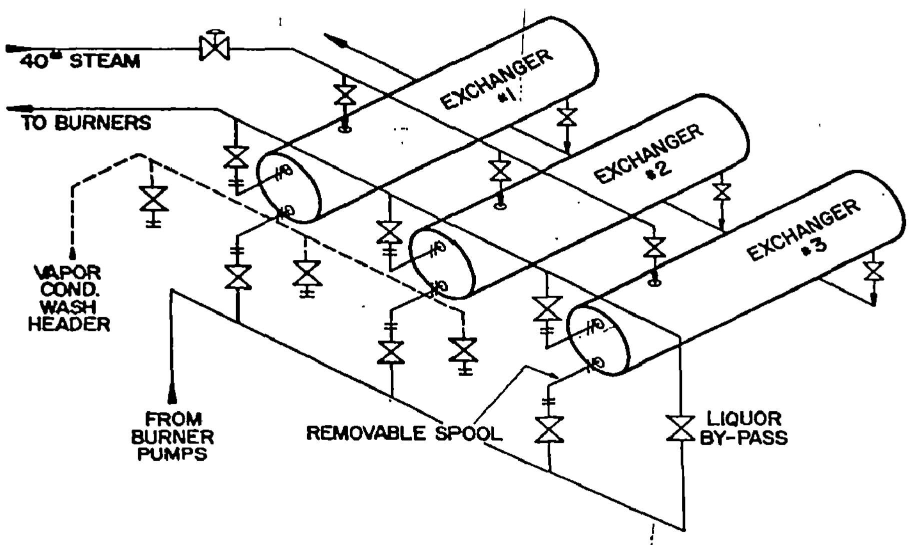


Figure (12) Arrangement of Indirect Heating of Strong Black Liquor

The indirect heating system is located after the salt cake mix tank and prior to the on line solids measurement devices. Each exchanger is designed to heat 54934 kg/hr of 62% black liquor from 90^0C to 107^0C. The pressure drop through each unit at full capacity is designed at 1.72 bar on the liquor side (tube side). The steam requirements were designed at 1019 kg/hr of 2.4 bar steam at 138^0C saturation temperature. The total heat duty for each unit is 2189 kJ/hr at full load. This implies a design overall heat transfer co-efficient of around 810 kJ/m^2 ^{0}C hr.

2.5. High solids evaporation of Kraft black liquor using heat treatment (14)

To increase the dry solids content further, to 70% - 80% range does present a small problem. The problem does not lie in the evaporator equipment but in the handling of the concentrated black liquor after the concentrator. Specially, the problem is related to the typical properties of the black liquor and more precisely the viscosity. Black liquor viscosity changes rapidly with the dry solids content in this range, results in difficulties in handling.

2.5.1. Black liquor viscosity (8)

The viscosity is an important rheological property of black liquor. Viscosity, which changes with the composition and temperature of black liquor as well as pulp mill operation practices, is of interest because of its on evaporation rate, heat transfer rate and liquor spray size. High viscosity limits the heat transfer in multiple effect evaporator system.

The viscosity of black liquor for most hardwoods is 4 to 5 times more than that of softwood spent liquors. A sharp rise in viscosity of hardwood black liquors is noted at concentration of 40-45% solids. Substantial drop in viscosity is observed with increase in residual active alkali. RAA levels of 6-7 gpl and above prevent slow condensation of reactions during evaporation, which leads to increase in viscosity. Black liquor is considered non Newtonian in character with time dependency in nature

Table (4): Kraft hardwood black liquor viscosity at 80^0C in mPaS

Samples	45%	55%
Mixed hardwood	14	123
T Tomentosa	16	145
E Tereticornis	33	370
N Latifolia	9	79

The reduction in viscosity has become a major necessity for handling particularly non-wood black liquors. The techniques available for reduction in viscosity from handling point of view can be broadly discussed,

Discussion	Effect
❑ **Increasing temperature of the black liquor**	The viscosity falls down significantly at elevated temperatures. This concentrated black liquor is usually handled in high temperature bodies of the multiple effect evaporator system. Strong black liquor is handled at 80-90^{0}C while it is sprayed in recovery furnace around 110-120^{0}C.
❑ **Decrease of concentration**	The economy of processing black liquor in recovery system is strongly dependent on the increasing solid concentration in final black liquor. However increasing viscosity makes the handling difficult. Therefore attempts are being made to burn the black liquor of lower concentration even at the cost of reduced thermal efficiency.
❑ **Increasing residual alkali concentration**	Increasing RAA in black liquor reduces viscosity significantly. This means increasing black liquor RAA to suitable level (around 6 gpl) by adding NaOH externally at appropriate concentration to improve the handling of black liquor.
❑ **Removal of silica**	Usually removal of silica (particularly for non wood black liquor) has the advantage of reducing viscosity at same levels of RAA this has the advantage of reducing the scaling tendency as well.
❑ **Addition of Selective chemicals (like Busperse-47 or Chemosperse-47)**	It is noted that Kraft black liquor have lower viscosity than soda black liquors. This leads to believe that some surfactants (like lingo sulfonates) may be able to reduce viscosity of soda black liquors. Some non ionic chemicals like busperse-47 at 25-40 ppm level act as good penetrating and dispersing agents and reduce black liquor viscosity.

2.5.2. Heat treatment system (14)

The thermal heat treatment system was developed based in a theory of thermal deploymerization at the large linguine molecule fraction in the black liquor. The system where the black liquor is heated up to a temperature above the digester temperature and kept at this temperature for about 20 to 40 minutes. The effect of this treatment is a permanent reduction of the liquor viscosity.

The treatment taken place at around 40-50% dry solids and could be an integral part of the evaporator plant. Liquor from the evaporation effect is taken out at about 40- 45% dry solids. The liquor is heated in consecutive heat exchanger stages, up to a temperature of 180-210^{0}C and flows into a reactor vessel. The retention time in the vessel varies, depending on the production but the normal time is around 30 minutes. After the retention vessel the liquor flows through a series of flash vessel where the heat treated liquor is flash cooled down to a temperature suitable for liquor feed to the further effect.

As a result of thermal of black liquor, remarkable black liquor viscosity reduction with achievable higher dry solids in remaining effect of the evaporator plant. Reduction in viscosity after thermal treatment could help in handling the black liquor by keeping at less viscous even at atmospheric pressure.

The viscosity results of pilot plant scale during thermal treatment of mill black liquor (15% eucalyptus, 85% bagasse) is given in **Figure (13).**

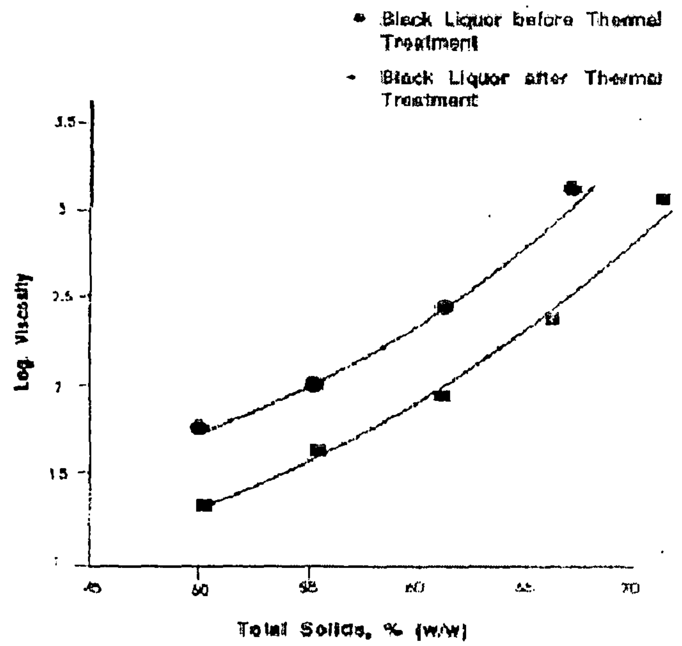


Figure (13): Viscosity results of pilot plant trial during thermal treatment of mill's black liquor (Bagasses 85% + Eucalyptus 15%)

Case study of the plant as (14),

The evaporator plant of Rosco type 7 effect free flow falling film evaporator designed for 68% dry solids. After the installation of thermal treatment plant, started getting the final heavy black liquor at about 75-77% dry solids. The direct benefit of the plant is 5% capacity increase in production rate. The SO_2 emission from the recovery boiler has reduced from a level of about 150-200 ppm down to 0 ppm.

2.6. Condensate recovery and return

Steam is one of the most important of utilities and also the most expensive. Recovering and recycling of condensate is key factor in any efficient steam system.

2.6.1. Flash steam (19)

Condensate is discharged through steam traps from high pressure to low pressure. As a result of this drop in pressure, some of the condensate re-evaporates. This referred to as flash steam. About 50% of the total energy supplied could be lost through flash steam. Flash steam recovery is an essential part of achieving an energy efficient system.

An effective condensate recovery system collects the hot condensate from the steam using equipment and returns it to the boiler feed system. It can rapidly pay for itself in reduced fuel costs alone.

Fig. (14) Approximate Amount of Flash Steam In Condensate

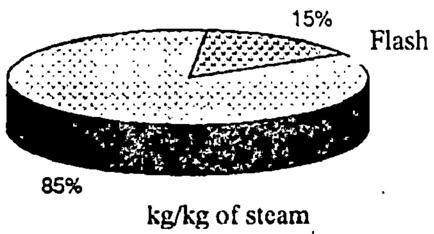
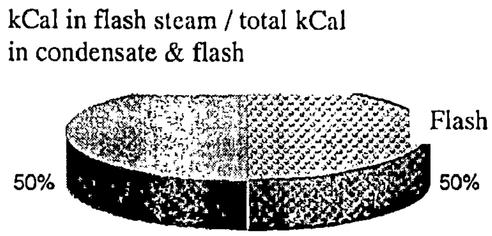


Fig. (15) Approximate Amount If Energy In Flash Steam



Why return condensate?

- Monetary Value: Condensate is such a valuable resource that the recovery of even relatively small quantities is economically justifiable. Even the discharge from a single steam trap is often recovering.
- Water charges: Any condensate not returned needs to be replaced by make up water. If condensate is recovered, water charge is reduced.
- Effluent restrictions: Draining of hot condensate is increasingly restricted in global standard plants.
- No boiler dearating: Boiler output is maximized.
- Reduced water treatment costs: Condensate is ideal boiler feed water.

Why recover flash steam?

The best reason to use a flash tank is to reclaim the energy from the flashing steam. When a Flash Tank is used and flash steam is returned to the system it not only recovers the full kJ of the steam but also has other benefits.

1. The steam is actually soft water that would need to be replaced if vented from the system.
2. The condensate at low-pressure is easier to return to the system.
3. It eliminates transferring a high temperature mixed flow media.
4. Provides an area for air separation.
5. Reduces condensate line sizing due to excessive steam loads.

6. Creates a cushion in return lines.

7. Reduces cooling water requirements.

An example (19) of the flash calculation is 6.89 bar condensate flashing to a 0.34 bar low-pressure steam line.

$$\frac{309 - 196}{960} = 11.8\% \text{ Flashing to Steam.}$$

For a condensate load of 450 kg/hr. X 11.8 % = 53.1 kg/hr. would flash to steam. This leaves 396.9 kg/hr. of condensate at 0.34 bar. This flash steam of 53.1 kg/hr X 2568.8 kJ/kg = 136,408 kJ per hour recovery. Using a Rs. 225 / million kJ fuel cost and a boiler efficiency of 80%. The savings equal Rs. 38.25 per hour. For a boiler operating 24 hrs. A day 365 days a year this would save Rs. 336,087.

2.6.2. Recovering heat from boiler blow down (19)

Blow down contains a lot of heat. Most boiler manufactures recommend automatic TDS control. This is done by continuously taking boiler water from 4-6" below the surface of the boiler water where TDS is at its highest level. Continuous blow down is the only acceptable way to control boiler TDS. Even so blow down losses can be tremendous and heat recovery is required.

The first stage of continuous blow down heat recovery uses a flash tank to flash part of the blow down water into steam. Most of the time a deaerator or feed water heater operating at 0.34 –1.03 bar and can use all the additional steam. The heat recovered from the flashing continuous blow down, is 50-55% of the heat recovered. Returning the steam to the system also saves make-up water that would otherwise need replaced.

A second stage can be added to the flash tank and heat recovery system that includes a heat exchanger that transfer the remaining heat into the boiler feed water. This stage of the heat recovery can provide 3 5-40% of the heat recovery. Besides transferring heat to make up water the blow down is cooled to 37-43^{0}C and will be acceptable in most drain systems.

2.7. Application of magnetic flux technology in furnace oil handling system (18)

The most important factors in the flux product are the magnetic field intensity and the collimation of the magnetic lines of flux. The intensity of the magnetic field by flux product is far superior to that generated by regular permanent magnets and the collimation of the magnetic fields renders the magnetic lines of flux exactly parallel to each other at extremely high densities (to the order of millions of lines of flux per sq. cm.).

The advantages of the flux product as follows:

- Increase kJ output.
- Increase thermal efficiency of the fuel.
- Reduces maintenance on the fuel side.
- Reduce carbon buildup.
- Increase the combustion efficiency, reducing fuel consumption.
- Reduces stack emission levels.
- Lead to a cleaner environment.

Flux Hi Tech India Pvt. Ltd. has conducted exhaustive research into the utilization of permanent magnetic fields in alleviating these problems currently associated with hydrocarbon fuel combustion. These studies led them in invention of the Flux maxiox, with its rare earth metallurgy and the unique patented flux-collimator-pervader.

2.7.1. Principles of flux maxiox (18)

The hydrogen atom has one positive charge (proton) and one negative charge (electron), i.e. it possesses a dipole moment. It can be either diamagnetic or paramagnetic (weaker or stronger response to the magnetic flux) depending on the relative orientation of its nucleus spins. Hence, it occurs in two distinct isomeric forms – para and ortho, characterized by the different opposite nucleus spins. In para H_2 molecule, which occupies the even rotation levels (quantum number), the spin state of one atom relative to another is in the opposite direction rendering it diamagnetic. In the ortho molecule, which occupies the odd rotational levels, the spin are parallel with the same orientation for the two atoms, and therefore is paramagnetic and a catalyst for may reactions.

The spin orientation has a pronounced effect on physical properties (specific heat, vapor pressure), as well as behavior of the gas molecule. The coincident spins render ortho-hydrogen exceedingly unstable and more reactive than its para-hydrogen counterpart. To secure conversion of para to ortho state, it is necessary to change the energy of interaction between the spin states of the H_2 molecule.

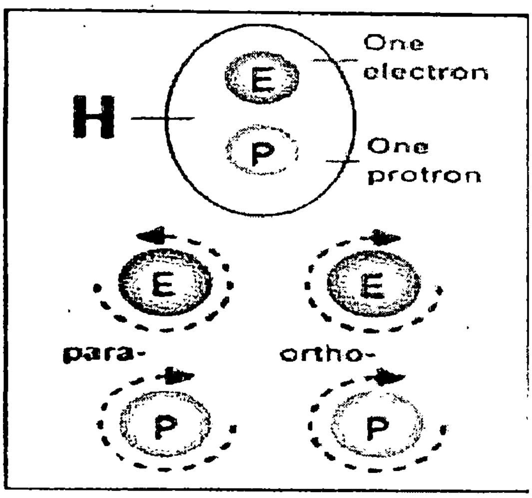


Figure (16): Changes of hydrogen in hydrocarbon molecules.

At 20^0C, 75% of hydrogen is in the para form. It is only when we drop the liquid hydrogen temperature to -235^0C that 99% of the hydrogen is in the ortho state. In the fifties an American rocket scientist, Simon Ruskin, realized that para hydrogen could converted to higher energized ortho hydrogen through magnetic simulation, i.e. the application of the proper magnetic field to change the spin state of the hydrogen molecule. This greatly enhances the energy of the atom and the general fuel reactivity, i.e. the combustion efficiency.

The same principle is utilized and the same effect is achieved by the action of the Flux maxiox, where a strong enough flux is developed to substantially change the hydrogen in the hydrocarbon molecule from its para state to the higher energized ortho state. The combustion of a molecule energized as such will give a better kJ output than the same molecule in para state. This results in higher energy output for the same quantum of fuel burnt.

2.8. Summary

The literature survey helped to bring out possible approaches of energy conservation in the area of which study made. With support of literature, the case studies are outcome with suggestion of improvements.

CHAPTER 3

OVERVIEW OF PULP AND PAPER INDUSTRY & PERFORMANCE STATUS OF PLANT

3.1. INTRODUCTION

The Pulp and Paper Industry has made steady progress in this country. At present there are about 424 plants manufacturing Paper, Paperboards and Newsprint. Total installed production capacity of 5.5 million tonne per annum as compared to 323 million tonne per annum of total world capacity. The state wise breakup capacity is given in **Table (5)**

Table (5): Working capacity of mills in India (2)

State	No of Mills	Working Capacity, tonnes	% Initial Working Capacity
Maharashtra	73	1029	18.7
Uttar Pradesh	83	919	16.7
Gujarat	68	758	13.8
Andhra Pradesh	20	513	9.3
Tamil Nadu	32	423	7.7
Punjab	35	360	6.5
Orissa	9	299	5.4
Karnataka	13	257	4.7
Others	91	939	17.2
Total	**424**	**5497**	**100**

The Pulp and Paper Industry is highly energy intensive. The main fuel used in the paper industry is coal. The other fuel used are Furnace oil, Diesel, Black liquor etc. Large mills generate most of power through Co-generation, while smaller mills depends exclusively on purchased power. Energy cost accounts for 25% of the total cost of manufacturing and it is rising by leaps and bounds every year. The breakup of cost of production under various heads is shown in **Figure (17)**.

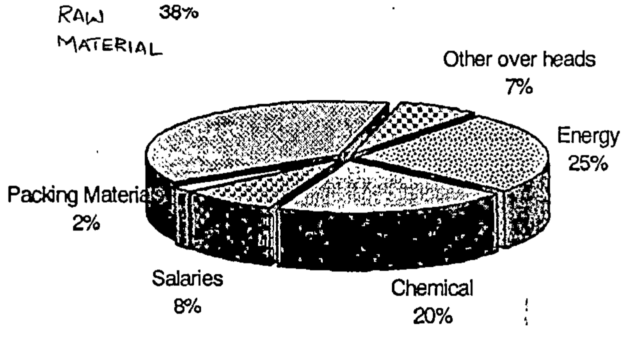


Figure (17): Breakup Cost of Production per tonne of paper

The consumption of paper and paperboard has increased from 1.1 million tonne in 1980 to 4.0 million tonne in 2001 indicating an average growth rate of 6-7% per annum and improving the per capita consumption in the country from 2.19 to 4.2 kg. The consumption per capita of paper in India is still one of the lowest in Asia, where the average annual consumption is around 20 kg.

3.2. AN OVERVIEW ENERGY CONSUMPTION PATTERN

Pulp and Paper Industry- essentially a chemical process and energy intensive industry – consumption about 7% of the country's coal and approximately 3% of the electrical energy requirements of the whole manufacturing sector. The industry needs only about 15-25% of high-grade energy (electrical) while the rest is constituted by low-grade energy.

The **Table (6,7)** shows a comparative Energy Consumption pattern between paper mill in India and the developed countries.

Table (6): Steam consumption of developed countries and India (3)

Particulars	Developed countries	India
Digester	1.9-2.3	2.3-3.9
Evaporator	1.5-2.2	2.5-4.0
Paper Machine	1.9-2.0	3.0-4.0
SR Plant	0.3-0.5	0.5-1.1
Bleach Plant	0.2-0.25	0.35-0.40
Total Specific steam consumption, t/t of paper	**5.5-7.5**	**9-13**

Table (7): Electricity consumption of developed countries and India (3)

Particulars	Developed countries	India
Chipper	92-98	112-118
Digester	43-46	58-62
Washing and Screening	116-122	145-155
Bleaching Plant	66-69	88-92
Stock Preparation	164-175	275-286
Paper Machine	410-415	465-475
SR Plant	127-135	170-190
Utilities and Others	160-165	246-254
Total consumption, kWh/t of paper	**1150-1250**	**1200-1700**

The specific energy consumption of Indian paper mills is between 1200 to 1700 kWh/tonne of paper as compared to 1150-1250 kWh/tonne of paper. Specific steam consumption is of 9 to 13 tonne/tonne of paper as compared to 5.5 to 7.5 tonne/tonne of paper in developed countries. There is enormous scope for energy conservation in this sector.

3.3. PERFORMANCE STATUS OF THE PLANT

3.3.1. Profile of the plant

ITC entered the field of paperboards in 1975 when it incorporated **Bhadrachalam Paperboards Ltd**. The new company was setup as an integrated paperboard manufacturing facility and commenced operations at Bhadrachalam in Andhra Pradesh, 300 km east of Hyderabad. The Bhadrachalam mill today produces 210,000 TPY of papers & boards and is the largest single location mill in India. The mill is focused on producing paperboards for packaging and graphics segments and product range includes **Cyber XLPac (folding box boards), Pearl/Saphire Graphik (solid bleached boards)** high value boards apart from the Ecoviron range of recycled boards. The mill also makes **liquid packaging boards** for Tetrapak in India.

Very recently the mill also commissioned India's only ECF pulp mill with a capacity of 100,000 tonnes a year. This location will also see the commissioning of an 80,000 TPY board machine from Voith by June 2004. The Bhadrachalam location today has two board machines and two smaller paper machines. The unit is ISO 9002:2000 series accredited. The unit is also ISO 14001 certified for EMS.

In the track of Energy Conservation also, the industry leads in its own way by adopting several conservation measures and has been presented with many awards.

The mill had been chosen for energy conservation study because of their higher motivation and encouragement in energy conservation activities.

3.3.2. Technological status of plant

The paper mills, which are in existence today, have been installed over a span of more than 100 years, and hence practice technologies falling in a wide spectrum ranging from oldest to very modern. Most of the machinery installed is imported. The new technologies like,

- New fiber line with Pressure Screen, and Double drum press washer, reporting fiber loss of 1.5-2.0%, and soda loss of 7.5 – 8.0 kg Na_2SO_4/ kg BD pulp.
- Two stage Oxygen delignification helping in reducing the Kappa number of the unbleached by 45%.
- Chlorine dioxide is used in bleaching & eliminating chlorine and hypo chlorine. The bleach sequence followed is D1 – Eop – D2. Helping in reduction of pollution load (AOX load reduced to 0.65 kg/T of pulp).
- Secondary Fiber Treatment Plant, Maximum capacity of 70 TPD, reducing the pollution load.

- Modern Recovery Boiler with firing capacity of 625 T BLS/day, and steam generation of 3.5 t/t of BLS.
- Modern system of Black liquor evaporation from 18% conc. to 70% conc. with 7 effects FFFF evaporator.
- Sophisticated instrumentation (DCS) in Digester house, and SRP.

The modernized technology helping the plant toward the growth of energy efficient and environmental friendly.

3.3.3. Process flow description (19)

The schematic flow process description of pulp and chemical recovery section of mill is given in **Appendix.**

Appendix-I/1: Process flow description of digester house.

Appendix-I/2: Process flow description of new fiber line (System of washing, screening, & Oxygen delignification)

Appendix-I/3: Process flow description of ECF bleaching sequence

Appendix-I/4: Process flow description of black liquor evaporation by 7 effects FFFF Evaporator.

Appendix-I/5: Process flow description of recovery furnace.

Appendix-I/6: Process flow description of recausticizing section.

3.3.4. Steam generation and power generation units

3.3.4.1. Coal Fired Boilers (CFB)

The plant has five coal fired boilers. The specification of the operating boilers as follows,

Table (8): Capacity description of cogeneration boiler of mill

Type	Make	Capacity, MT/ hr	Fuel	Steam Pr. / Temp, bar / ^0C
CFB-1	ABL make, Spreader stoker	27.5	Coal	42/480
CFB-2	ABL make, Spreader stoker	27.5	Coal	42/480
CFB-3	Spearing & Co. AFBC	20	Coal	12.5/300
CFB-4	ABL make, AFBC boiler	50	Coal	62/480
CFB-5	ABL make, AFBC boiler	80	Coal	62/480

The steam generation and coal consumption for the period April 2003 to Feb 2004 of these boilers are as follows:

Boilers	Steam Generation, MT	Coal Consumption, MT
CFB-1	50051	46393
CFB-2	80591	
CFB-3	103895	
CFB-4	338239	59404
CFB-5	600333	105016
Total	1173109	212813

3.3.4.2. Turbo Generator

The plant has three Turbo Generators. The specifications of the TG's are as follows:

Table (9): Specification of the Turbo Generator of the mill

Particulars	TG-1	TG-2	TG-3
Type	Back Pressure Cum extraction	Back Pressure	Extraction Cum Condensing
Make	BHEL	BHEL	BHEL
Capacity, MW	5	7.5	21
Inlet steam parameters Pr. in bar (P)/Temp. ^{0}C(T)/Quantity tph (Q)	42 / 405 / 50	62 / 480 / 56	62 / 480 / 80
Extract Steam P/T/Q	11.5 / 240 / 10	-	-
Back pressure steam parameters P/T/Q	4.6 / 220 / 40	4.5 / 220 / 56	5 / 220 / 40
Condensing Steam P/T/Q	-	-	-0.91 / 50 / 80

The TG-1 & 3 are of controlled extractions. TG-2 & 3 are synchronized with the grid and TG-1 is in island mode and catering to the catering to the emergency loads such as auxiliaries of the boilers, etc.

Against 33.5 MW installed capacity of co-generation, about 26.3 to 28.5 MW of power being generated and balance 1.5 to 3.0 MW being imported from grid.

When steam demand in the plant increases, the demands get supplied from CFB-2. Where the CFB-2 is a stand-by boiler. The present system energy balance of the mill is given in **Appendix-I/7.**

3.3.5 Operational observation of pulp and recovery Section

The following **Table (10)** shows the operational observation in pulp and recovery section of the mill, which is used in case studies for bring out the various recommended practice.

Table (10): Operational observations of pulp and recovery sections

Composition of Raw Material:

Raw Material	% Consumption	Ash %	Lignin %	Hemi-cellulose %	Cellulose %	Average length, mm	Average diameter, mm
Eucalyptus	50	1.0	24.7	14.1	50.6	0.73	12
Bamboo	20	1.7-4.8	27.8	15.1	59.9	1.65	12
Subabul	30	NA	NA	NA	NA	NA	NA

Composition of cooking Liquor (as Na_2O):

NaOH	:	68 gpl
Na_2S	:	21 gpl
Na_2CO_3	:	17 gpl
Na_2SO_4	:	2.3 gpl
Active Alkali	:	89 gpl
Sulphidity	:	23.59%
TTA	:	106 gpl
Causticizing Efficiency	:	80%

Cooking Schedule of Digester:

Particulars	Time, minutes	
	80 m^3 digester	100 m^3 digester
Dry loading 1. Chips fill 2. White liquor fill (1st fill) 3. Chips packaging and Black liquor addition	 45 5 10	 45 5 10
MP steaming and 2nd fill of white liquor	135	135
Time at temperature (cooking)	45	45
Blow	15	20
Total time	255	260

Weak Black Liquor Analysis from Pulp Mill:

Particulars	Conditions
pH	12.6
RAA, gpl as Na_2O	10.79
Fiber content, ppm	22
Solids %TS	17.94
Organics, % by Wt	50.63
Inorganic % by Wt	49.37
Acid insoluble, % by Wt	3.47
Silica, % by Wt	1.89
GCV, Kcal per kg of BlS	3356.5
Sodium sulphate Na_2SO_4, % by wt as such	5.57
Sodium Carbonate, % by wt as such	13.5
TTA, Na2O, % as such	29.15

Elemental composition of Firing Black Liquor to SRP:

Particulars	Conditions
Elemental analysis of black liquor from HBL tank	

1. Carbon, % 2. Hydrogen, % 3. Oxygen, % 4. Sodium, % 5. Sulphur, % 6. Chloride, % 7. Potassium, % 8. Inert, %	35.90 2.30 38.04 13.70 2.42 3.94 3.43 0.27
% Conc. of liquor to firing	66.45
Make-up salt, TPD	10.00
Spent acid from ClO_2, TPD	20.00
Ash addition, TPD	1.23
Organics, % on BLS	57.22
Inorganic, % on BLS	42.78
GCV, kJ/kg of BLS	12852

3.3.6. Energy consumption profile of the plant

The ITC Ltd., Bhadrachalam plant uses Coal, Furnace oil and Black liquor (by-product) as the main source of energy for generation of steam and electricity. The electricity also imported from APSEB grid to meet the balance requirement.

The **Table (11)** shows Annual production, Coal consumption, Steam generation, Power generation, and Grid power utilization for the year of 2000 to 2004 (to date) are as,

Table (11): Utility bill of plant from 2001 to 2004

Particulars	Units	2000-01	2001-02	2002-03	2003-04
Gross production	Lakh MT	2.31	2.36	2.38	2.27
Coal consumption	Lakh MT	2.30	2.36	2.43	2.13
Steam generation (Coal)	Lakh MT	12.52	13.49	13.35	14.73
Steam generation (BL)	Lakh MT	2.77	3.03	3.86	5.26
TG power generation	GWh	213.865	216.847	228.136	214.914
DG power generation	GWh	2.813	2.105	1.049	1.224
Grid power	GWh	8.749	12.367	12.633	15.573
Furnace oil	kl	382	410	344	360

3.3.7. Energy demand of process sections

Steam demand of ~~130~~ 151 tph at 4.5 bar, 200^{0}C and ~~27.5~~ 32.2 tph at 11 bar, 300^{0}C is met by six boiler using coal and black liquor as fuel. The steam balance diagram of the mill is given in **Appendix-I/7.**

Table (12): Steam demand of process section of plant

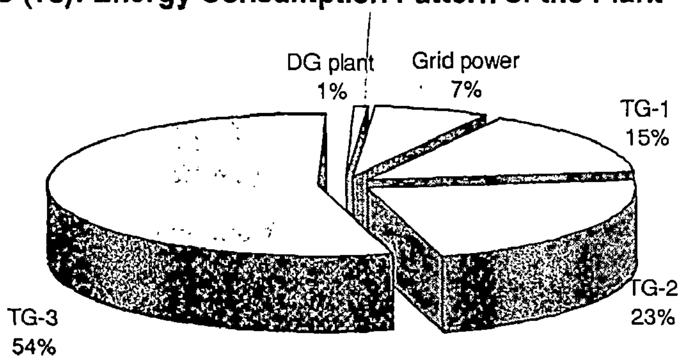
Area of Consumption	Steam Consumption TPH	
	MP steam @ 11 bar	**LP steam @ 4.5 bar**
Paper M/C –1	0.6	23
Paper M/C – 2	-	8.8
Paper M/C – 3	-	2.6
Paper M/C –4	2.6	42
Dearator –1		6.04
Dearator –2	-	7.03
Dearator – SRP	-	6.01
Digester House (include BHR)	19.0	5.75
Fiber Line	1.7	0.86
Bleach plant	-	14.0
ClO_2 Plant	3.5	1.7
Evaporator		20
Causticizer		2.8
SFT		5.0
SRB	4.8	5.0
Coating Plant	-	0.5
Total	**32.2**	**151.08**

The electricity demand is about 231.711 GWh is met by 3 TG Plant, DG Plant and by APSEB.

Table (13): Electricity demand of the Plant

Area	Electricity generation, GWh
TG-1	33.958
TG-2	52.224
TG-3	128.732
DG plant	1.224
Grid Power	15.573
Total	**231.711**

Figure (18): Energy Consumption Pattern of the Plant



Specific energy consumption of the plant section wise,

Table (14): Specific energy consumption pattern of the plant (19).

Sl.No	Particulars	Sp. Power Consumption, kWh	Sp. Steam Consumption, MT
1.	Chipper, per tonne of chips made	10.97	-
2.	Pulp Mill, per tonne of unbleached pulp	7.00	1.62
3.	Bleach plant, per tonne of bleach pulp	372.56	1.30
4.	SRP, per tonne of AA recovered	531	6.93
5.	Co-generation plant, per tonne of steam production	18.12	0.12
6.	PM-1, per tonne of machine production	566	3.04
7.	PM-2, per tonne of machine production	851	2.60
8.	PM-3, per tonne of machine production	773	2.39
9.	PM-4, per tonne of machine production	515	2.24
10.	Others, per tonne of m/c production (gross)	102.18	NA
11.	Total, per tonne of paper production (gross)	**1020.75**	**7.35**

As the specific energy consumption is high compared to the developed countries, there is more scope for energy conservation.

3.3.8. Energy cost pattern of the plant

Table (15): Energy cost pattern of the plant

Particulars	Annual consumption (2003- 04)	Unit Cost Rs.	Annual Cost Rs. Crores	% On Total Annual Cost.
Coal consumption	213,000 MT	1.5 / kg	31.95	83.42
DG power generation	1224,000 kWh	7.08 / kWh	0.86	2.24
Grid Power	15573,000 kWh	3.25 / kWh	5.06	13.21
Furnace oil	360 kl	12 / Lit	0.43	1.12

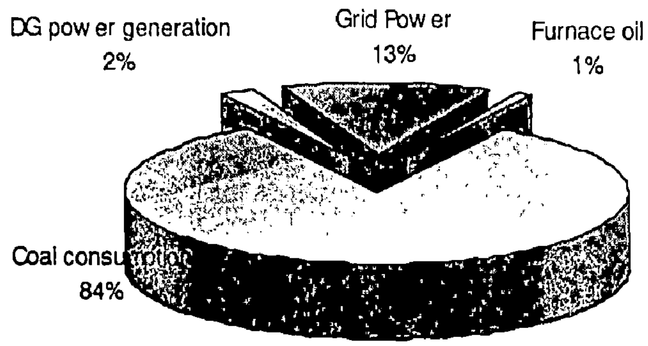


Figure (19): Energy Cost Pattern

From the **Figure (18 & 19)** it is informed that,

- Out of the total energy demand, 92% is supplied by coal and by product of the process (Black liquor). The remaining energy demand is supplied by grid of about 7%.
- Although only 7% of the plant energy requirement comes from grid, the cost of energy is as high as 13%. This clearly indicates the need to accrued top priority to energy conservation.

Energy cost used in the process calculation,

Table (16): Energy Cost of the Plant

1. Cost of Steam Generation.(Rs. Per tonne)		
Water	3.16	
DM water	17.05	
Coal	267.33	
Power	5.14	
Repair & Maintance	8.28	
Total	**300.96**	
2.Cost of TG generated power (Rs. Lakh Per kWh)		
Power generation (from Apr 03 to Nov 03)	Lakh kWh	1558.72
Coal value	700.69	
Cooling tower chemicals	7.91	
Repair & Maintance	68.58	
Power for steam production	163.86	
Total cost	941.03	
Unit cost	**Rs. / kWh**	1.27
3. Cost of Grid Power		
Unit cost	**Rs. / kWh**	**3.25**

3.4. Summary

The discussions with referred to performance status and utility bill of the plant, helped to know energy status of the plant. For the discussion regarding the economic feasibility of each case study, cost of steam generation of Rs. 300 per tonne, condensate of Rs. 20 per m^3, and unit charge of power Rs. 3.25 per kWh had been used (19). The average annual working hours of 330 days is considered.

CHAPTER 4

ENERGY CONSERVATION APPROACHES – CASE STUDIES

4.1. Two stage batch digester steaming

Present practice:

After charge of white liquor and black liquor into digester according to liquor to wood ratio, the liquor get circulated through indirect heat exchanger to attain cooking pressure of 7.0 ~~9.5~~ bar, 165^{0}C. The heating to this cooking temperature is carried out by MP steam supply to heat exchanger at pressure 11 bar. Since the heating is carried out by single stage MP steam. The detail calculation for steam requirement in digester house is given in **Appendix-II/1.**

Recommended practice:

In two stage steaming, the digester is initially heated and pressurized by low-pressure steam. After the digester has been partially brought to cooking pressure and temperature, it is switched over to 11 bar pressure steam, which takes the digester to full pressure (7 bar) and completes the cooking cycle.

By increasing demand for low-pressure steam, this method allows a change in the extraction ratios of mill turbine generators. The amount of 4 bar extraction may be increased while 11 bar extraction is decreased. This measure would boost the turbine's electrical output while using the same amount of input steam. Savings would be realized through the replacement of purchased electricity with less expensive co generated power. The MP steam extraction from TG plant (from TG-1) is 10 tph and remaining demand of plant is taken care by CFB-3. The details calculation of steam requirement and savings given in **Appendix- II/2.**

Table (17): Improvement in performance of cogeneration plant by applying two stage of steaming in digester house

Parameters	Performance value for existing process	New performance value by retrofitting the two stage steaming	Remarks
MP steam consumption for cooking, tph	20.33	11.56	Reduction in MP steam extraction from turbine.
LP steam consumption for cooking, tph	-	8.21	Reduction in MP steam extraction in system steam balance, equivalent increase in LP steam extraction.
Power generation of TG-1 (MP steam extraction turbine), MW	4.2	4.34	Reduction of power consumption from Grid by 6%.

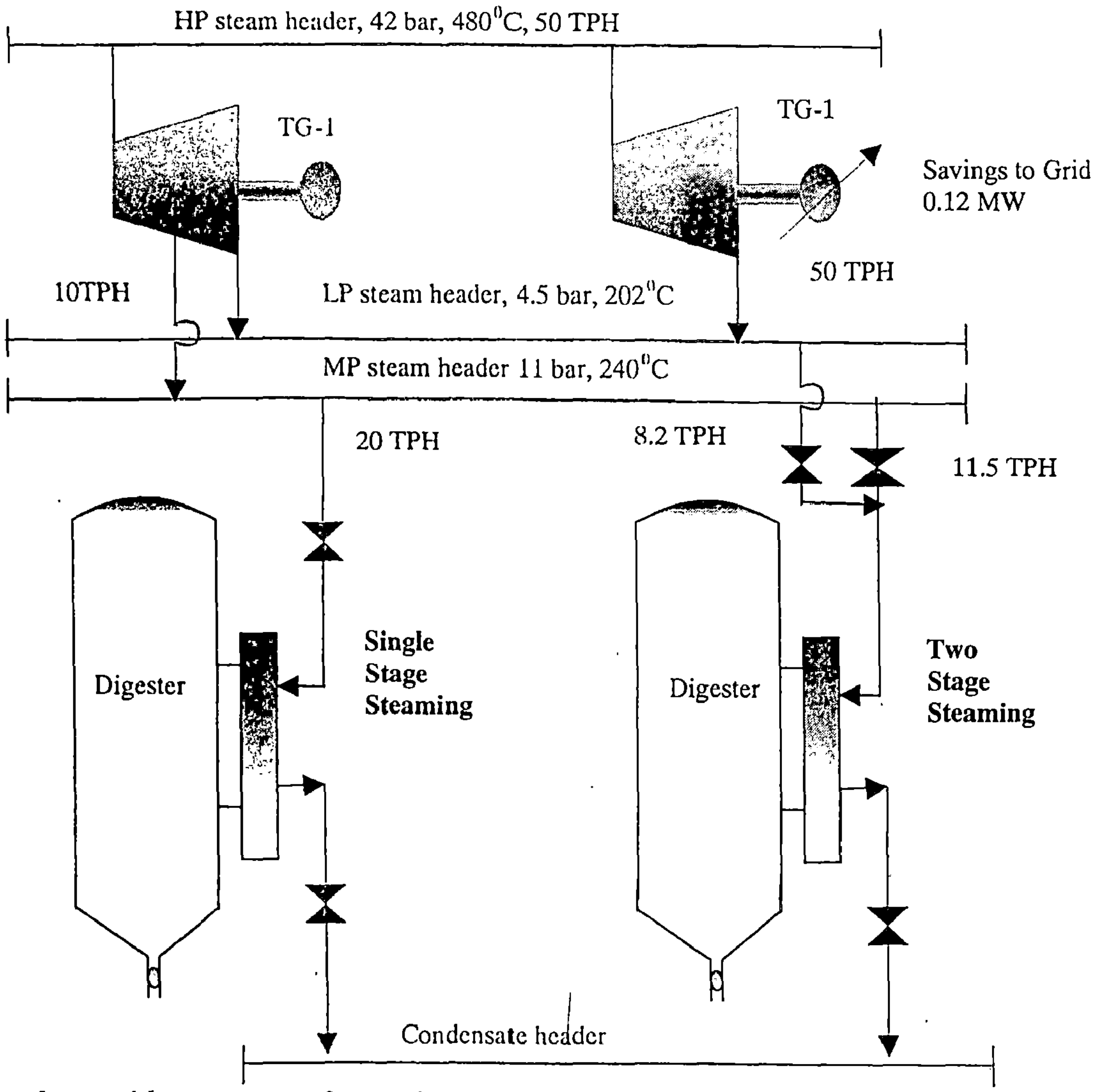
Possible savings by improvement,

Annual savings = Rs. 30.37 lakh

Investment = Rs. 50 lakh

Simple pay back Period = 23 months

Figure (20): Modified co-generation by two stage of steaming in digester house



Measures along with two stage of steaming,

By going for two stage of steaming, possibilities in increasing of cooking time of digester, this will affect the mill productivity. Along with two stage of steaming the following measures (Sec. 2.2.1) has to be taken care in step of improvement in productivity and energy savings.

- Presteaming of chips with steam packers – increases the bulk density of digester.
- Impregnation of chips with black liquor and cooking it up to first stage of steaming and displacing it with white liquor fill (i.e. taking part of advantage of RDH process) – increases yield of pulp with possibility reduction in alkali consumption.

- Spilt alkali dosage – increase yield of pulp with improved alkali profile.

4.2. Improvement of blow heat recovery system of pulp mill

Present practice:

In a day of maximum 39 blows are blown, each of 15 to 20 minutes duration. The total quantity of pulp blown is about 2332 MT/day for 80 m^3 digester and 884 MT/day for 100 m^3digester.

It was observed the temperature of the foul condensate at the outlet of the heat exchanger is 66^0C. The foul condensate temperature inlet to the PC and SC is going as high 66^oCwith flow of 570 m^3/hr; this is resulting in in-adequate condensation of flash steam.

It was observed that the temperature of hot water generated is about 65.5^0C for a flow rate of 230 m^3/hr, while the required hot water temperature in new fiber line is 80^0C. The low temperature of hot water is due to in-sufficient heat transfer in the plate heat exchanger. This has resulted in direct steam usage in the hot water tank to maintain the temperature of 80^0C. On an average the LP steam usage for hot water generation is about 5.74 MT/hr.

The existing plate heat exchanger has a total heat transfer area of 83 m^2 and this has got adequate provision for providing additional plates to increase the heat transfer area. The calculation for the amount of flash vapor generation and efficiency of the present BHR system is given in **Appendix-II/3.**

Recommended practice:

Making the availability of the foul condensate from the plate heat exchanger at temperature of 55^0C, by increasing the heat transfer area of plate heat exchanger. The flow of foul condensate to the PC and SC required is about 791 m^3/hr at 55^0C, the flow is respect to peak blow. The PC and SC condensate pump is to be replace with higher capacity.

The following measure needs to be taken to increase the BHRS efficiency,

- Replace the existing foul condensate pump with the higher capacity pump. The capacities of the existing and proposed pumps are given in the following **Table (18).**

Table (18): Blow heat recovery system, recommended pump capacity

Particulars	Existing pump capacity		Proposed pump capacity	
	PC	SC	PC	SC*
Flow rate, m^3/hr	380	190	632	159
Head, m	15	15	Supplier recommendation	-do-
* Taking the secondary pump handling capacity of 20% of total flow.				

- Increase the heat transfer area in the plate heat exchanger to 174 m^2 from the present area of 83 m^2.

The performance value for existing process and performance value after retrofitting are listed in the **Table (19)**:

Table (19): Improvement in performance of blow heat recovery system by applying proper system accessories

Parameters	Performance value for existing process	New performance value by retrofitting	Remarks
Temperature of foul condensate at inlet of PC and SC, ^{0}C	66	55	Decreased temperature will enhance the performance of the condensers.
Hot water temperature at outlet of heat exchanger, ^{0}C	65.5	75	Achieving the desired hot water temperature.
Heat transfer area of the plate heat exchanger, m^2	83	174	Increased heat transfer area to attain hot water temperature of 75^{0}C.
Additional LP steam usage in the hot water tank, TPH	5.74	2	Reduction in steam consumption at the hot water tank.

Possible savings by improvement,

Annual saving = Rs. 91.23 lakh

Investment = Rs. 15 lakh.

'Simple pay back period = 2 months.

By implementing these measures, the LP steam consumption in the hot water tank can be minimized to a great extent. The temperature profile in proposed heat is given in schematic below:

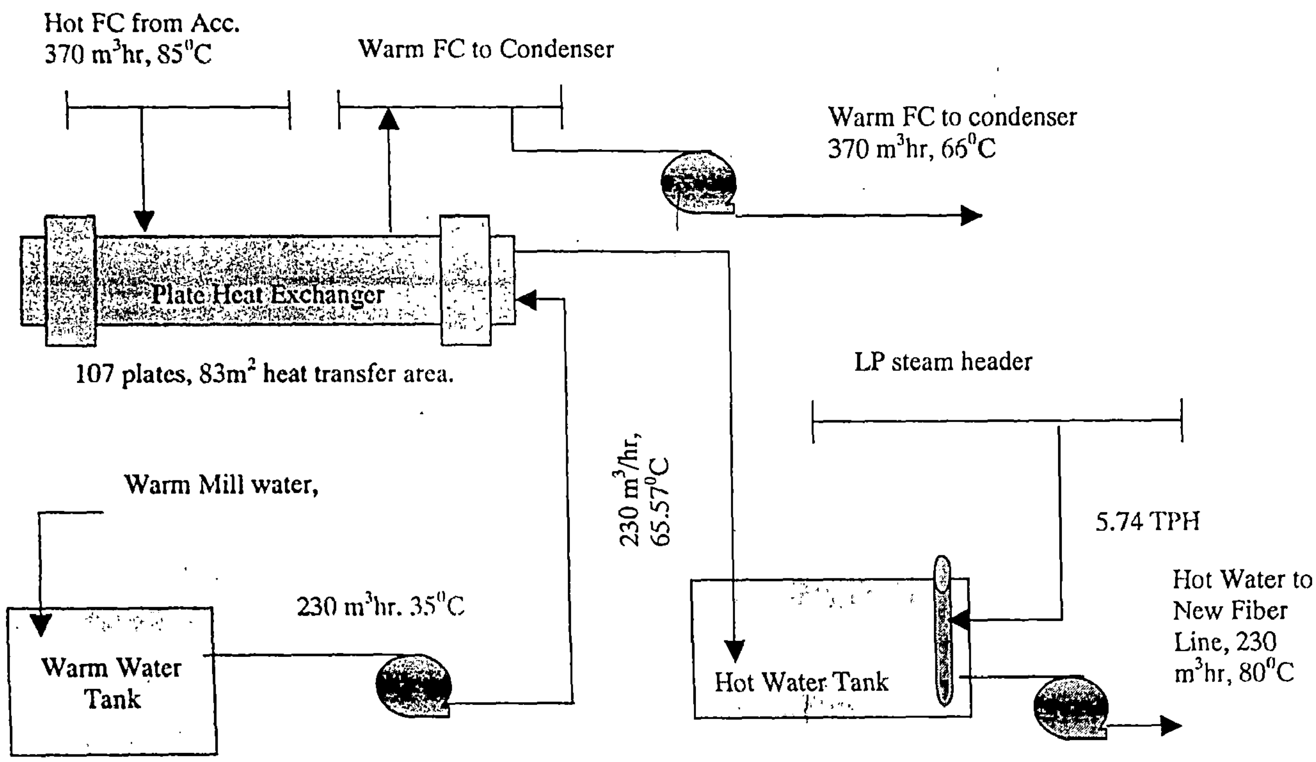


Figure (21): Present practice of hot water generation in BHR system

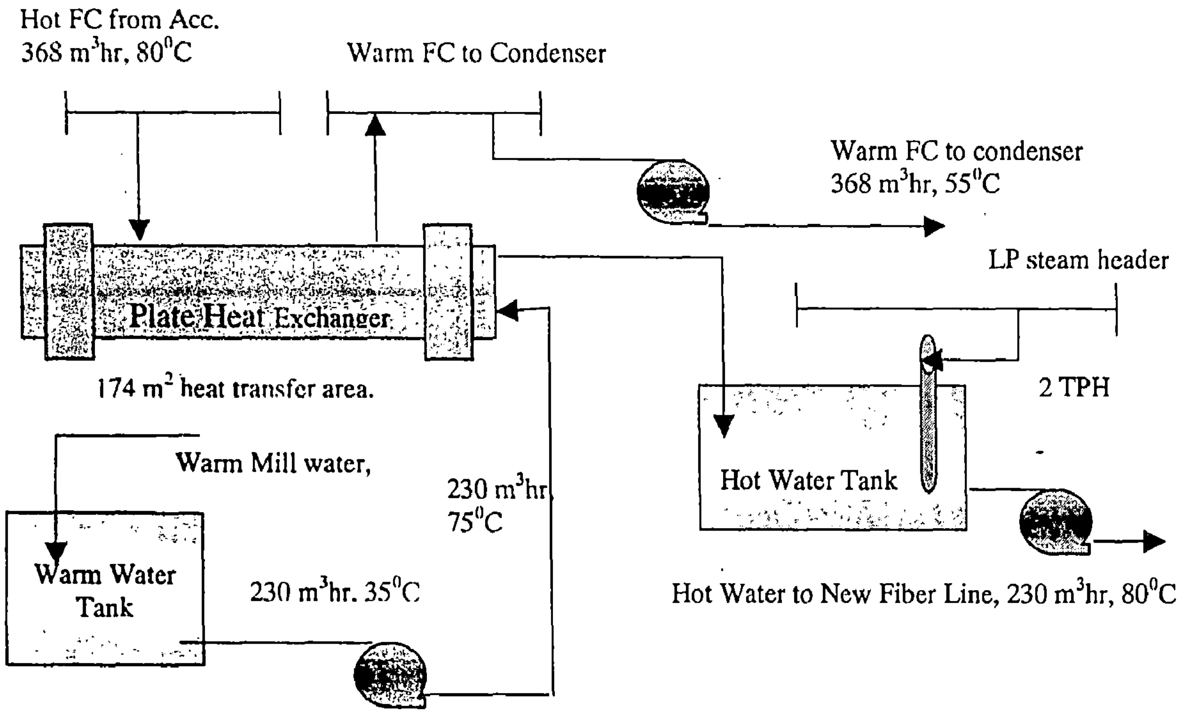


Figure (22): Recommended practice of hot water generation in BHR system

4.3. Proper scheduling of digesters

Digester cycle consists of a series of operations in sequence. The time of duration of each is dependent upon the pulping process followed. The various operations are carried out as follows:

I. Chip cooking and liquor filling.
II. Steaming for attaining the cooking temperature.
III. Holding at the cooking temperature.
IV. Blowing of pulp at the end of cooking.

The mill has eight digesters out of which six digesters are 80 m^3 capacities and two are 100 m^3 capacities, working with the indirectly heated system. The total pulp production is 345 TPD. The purpose of study is to find out the operational bottleneck and possible way of reducing it. This is likely to increase the productivity of digester house with reduced specific energy consumption. For finding out the possibility of operational bottleneck, the study is carried out for 24 hours.

A systematic study of the cooking cycle of the eight digesters is given the **Table (20)** using the analysis shown in **Figure (23)**.

Figure (23): Scheduling of Digesters

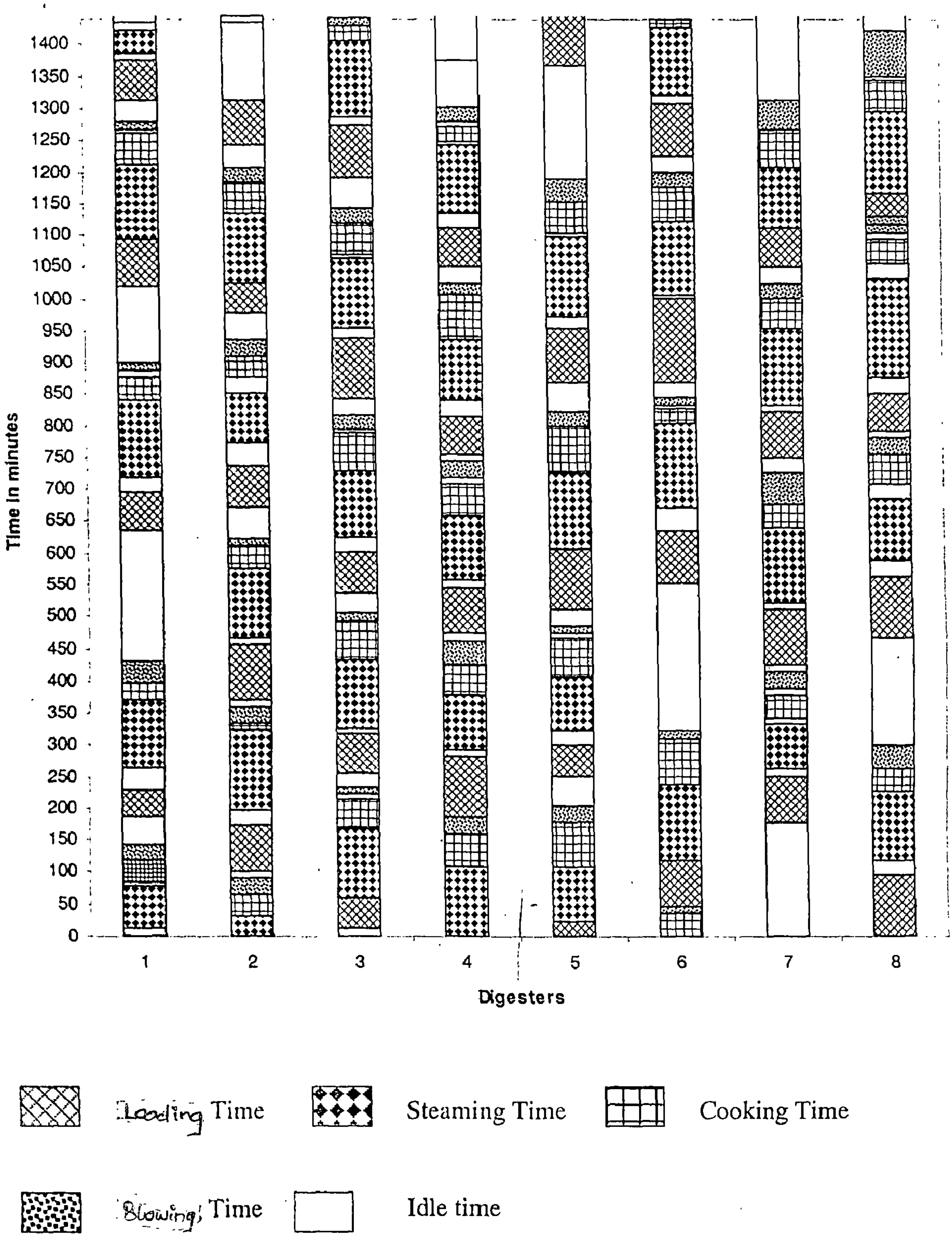


Table (20): Analysis figures of cooking schedule of eight digesters

For 80 m3 digester (6 nos.)				
	Time in Minutes			
Particulars	Min	Max	Standard Deviation	Average
Loading Time	42	132	20.97	74.37
Steaming Time	66	126	16.37	106.85
Cooking Time	24	72	16.30	46.61
Blowing Time	12	36	7.75	21.6
Idle Time	6	228	51.39	42.51
Cycle Time			-	291.94
No. of cycle per day			-	29.60
For 100 m3 digester (2 nos.)				
	Time in Minutes			
Particulars	Min	Max	Standard Deviation	Average
Loading Time	36	96	20.28	72
Steaming Time	72	132	25.61	112.5
Cooking Time	36	60	8.93	43.5
Blowing Time	24	72	17.49	37.5
Idle Time	12	180	60.63	45.88
Cycle Time			-	309.88
No. of cycle per day			-	9.29

From the above analysis figure it has observed that mill average cooking cycle is about 4.8 hours for 80 m^3 digesters and 5.16 hours for 100 m^3 digester. The average number of batches is about 38 batches per day.

The above analysis in **Table (20)** clearly shows that the major reasons for these bottle neck appears to be the following:

1. Loading Time:

- In both digesters (80 and 100 m^3) the maximum time observed to be ~3 times the minimum time, this is too high. Unless there is a bottleneck. It is also observed that average loading time of 100 m^3 digester is less than 80 m^3 digester.
- It appears that before and after chips loading, there is idle time in most batches. Of 39 batches, this is time for 24 batches. In the other two batches there is no idle time after chips loading. This appears to improve feeding arrangement of digester with proper supervision management.
- By doing so, it likely bring down the loading time. Taking 50% more than minimum as an ~54 minutes for loading in place of 74.37 minutes and 72 minutes. The possible time savings per day is about 603 minutes for 80 m^3 digester and 195 minutes for 100

m^3 digester. In term of production capacity is about 17.65 TPD for 80 m^3 digester and 6.79 TPD for 100 m^3 digester.

2. Steaming Time:

- In both digesters the maximum time observed to be 2 times the minimum time. The bottle neck appears to be non availability of full pressure steam, when more than one digester is waiting for steaming, the steam pressure falls down when the concerned digester are steamed, resulting increase in steaming time.
- This appears to improve steam feeding arrangement of digester. Installations of steam accumulator for digesters to act as surge tank, and thereby permit steaming of two digesters without increase in steaming time.
- By doing so, steaming time comes down. Taking 50% more than minimum as an ~99 minutes for steaming in place of 106.85 minutes and 112.5 minutes. The possible time savings per day is about 227 minutes for 80 m^3 digester and 121 minutes for 100 m^3 digester. Possible increase in production capacity is about 6.6 TPD for 80 m^3 digester and 4.2 TPD for 100 m^3 digester.

3. Cooking Time:

- The average time taken for cooking is about 46 minutes for 80 m3 and 43.5 minutes for 100 m^3 digesters. The cooking time for 80 m^3 digester is higher than higher capacity digesters. And time variation for cooking is about 16.30 minutes for 80 m^3 digesters and 17.49 minutes for 100 m^3 digesters.
- This appears to have proper H-factor control with proper supervision management.
- By doing so, cooking time comes down. Taking 50% more than minimum as an ~36 minutes for steaming in place of 46.61 and 43.5 minutes. The possible time savings per day is about 307.69 minutes for 80 m^3 digester and 67.5 minutes for 100 m^3 digester. Possible increase in production capacity is about 9 TPD for 80 m^3 digester and 2 TPD for 100 m^3 digester.

4. Blowing Time:

- In both digesters the maximum time observed to be 3 times the minimum time. The bottleneck appears to be an improper supervision management, waiting for blow due to overlapping of proceeding batch, and longer blow duration.
- The mill has two blow tanks, blow tank 1 is available only for digester 1 to 4 and blow tank 2 is available only for digester 5 to 8. It had appeared that when blow tank

2 is having its availability for blowing, where digester 1 to 4 get overlapped to blow in blow tank 1.

- This appears to have changes in blow lines (i.e. making available of blow tank to all digester) and increase in-size of blow line. Along with good supervision management.
- By doing so, it likely bring down the blowing time. Taking 50% more than minimum as an ~18 minutes and ~36 minutes for blowing in place of 21.6 minutes and 36 minutes. The possible time savings per day is about 104.4 minutes for 80 m^3 digester and 13.5 minutes for 100 m^3 digester. Possible increase in production capacity is about 3.05 TPD for 80 m^3 digester and 0.5 TPD for 100 m^3 digester.

5. Idle Time:

- From the schedule it is observed that there are long hours of gap between two successive operations due to no apparent reason. The average idle time per batch is observed to be 42.51 minutes for 80 m^3 digester and 45.88 minutes for 100 m^3 digester.
- With better supervision, this can be bringing about a significant saving.
- This would mean savings of about 986 minutes for 80 m^3 digester and 252 minutes for 100 m^3 digester. Possible increase in production capacity is about 28.87 TPD for 80 m^3 digester and 8.78 TPD for 100 m^3 digester.

An approximate impact of these improvements on scheduling is likely to give another 20% additional production or the present production being achieved from 8 digesters can be achieved in 4 digesters of 80 m^3 and 2 digesters of 100 m^3. The increase in capacity utilization will have its favorable impact in bring down the specific energy consumption.

The above study clearly shows the possibility in increasing of digester productivity; the mill management has to conduct this study for 72 hours for bring out further improvement possibilities.

4.4. Improvement of steam economy in MEE system:

Present practice:

The WBL from pulp mill at 18% concentration is feeding to 5th effect of 7th effect FFFF evaporator, the product liquor comes out at 67% concentration. The liquor flow sequence is 5-6-7-4-3-2-1B-1A. The designed WBL flow temperature to 5th effect is 80^{0}C, where now days the WBL temperature feed to 5th effect is higher than design temperature. The system economy is about 5.98; detail calculation of present system of evaporation is given in **Appendix-III/1.**

Recommended practice:

Feeding arrangement for 7-effect evaporator is mixed feed arrangement, the dilute feed enters an intermediate effect and flows to the next higher effect. On this section liquid flow occurs in the forward feed mode. Partly concentrated liquor is then pumped to the effect before the one to which the feed is introduced. It then flows towards the first effect in the backward feed mode. Thick liquor is withdrawn from the first effect.

In forward feed if the feed liquor is at higher temperature than the saturation temperature of the first effect, some evaporation will occur automatically as vapor flashing. Since the saturation temperature of the boiling solution in each effect is lower than the temperature of effect preceding it, there is flashing free evaporation in each succeeding effect, which reduces overall steam requirement.

The higher incoming temperature of WBL get utilized by feeding it to 4th effect, just by increasing the forward flow by one effect and reducing the backward flow by one effect. By this way steam economy of the system get improves.

Table (21): Benefits of high temperature black liquor feed to 4th effect of 7 effect FFFF evaporator

Temperature of feed liquor, ^{0}C	Feed to 5th body		Feed to 4th body	
	S.E	Final % conc. of liquor	S.E	Final % conc. of liquor
80	5.88	64.85	5.84	63.66
82	5.93	66.13	5.914	65.62
84	5.97	67.47	5.98	67.70
86	6.02	68.86	6.05	69.92
88	6.06	70.32	6.12	72.29
90	6.11	71.84	6.18	74.82
Comparison made on 20000 kg/hr of steam consumption.				

Figure (24): Improvement in Steam Economy of MEE system by feed to 4th effect instead of 5th effect

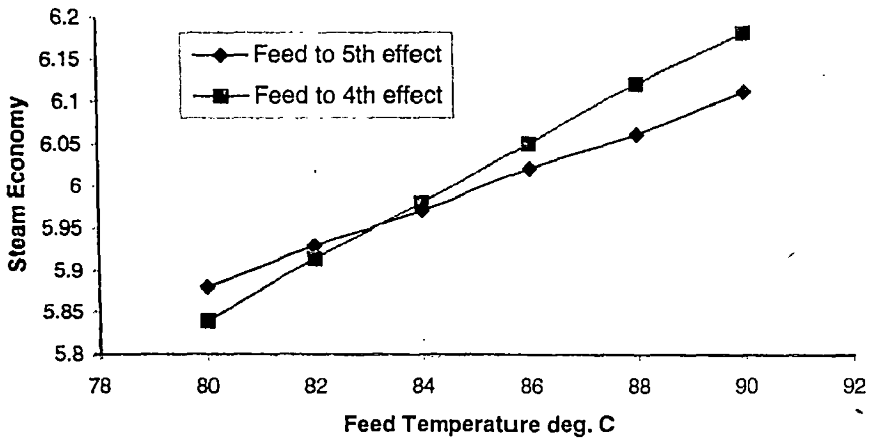
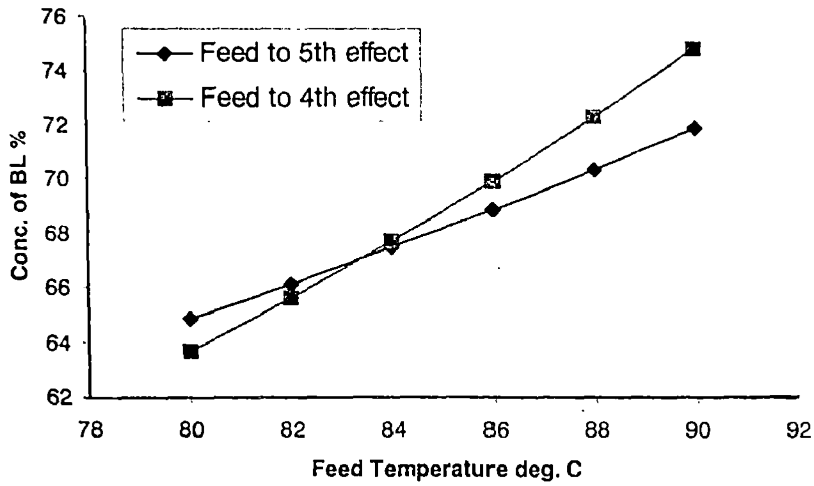


Figure (25): Improvement level in BL Conc. by feed to 4th effect instead of 5th effect



Based on analysis of **Figure** (24 & 25), the following conclusion drawn as,

- Where up to feed temperature of 83.75^{0}C, the system of evaporation is not providing any gain in steam economy as well as in concentration level of black liquor by feeding it to 4th effect.
- Where as feed temperature from 84 to 90^{0}C, the system performance is higher than present feeding arrangement.

Benefit in steam generation rate in boiler as,

Table (22): Energy saving cost by improvement in steam economy in MEE

Temperature of feed liquor ^0C	% Conc. of liquor from evaporator		Steam generation, TPH		Increase in steam generation, TPH	Net saving @ 300 per Ton of steam generation Rs. lakhs	Investment, Rs. lakhs	Simple pay back period, months
	Feed 5	Feed 4	Feed 5	Feed 4				
86	68.86	69.92	83.5	84.19	0.69	16.4	Nil	Margin
88	70.32	72.29	84.25	85.76	1.51	35.8	Nil	Margin

4.5. Controlling the crystallization of salts in black liquor and increasing the system availability:

Present practice

The evaporation of WBL from 18%TS to 67%TS is taken care by 7 effects FFFF evaporator, with the steam economy of 5.98. The first effect is subdivided in to three bodies out of which two are working and one is available for cleaning with WBL and process condensate. Similarly, the second effect is subdivided in to two bodies out of which one is working and other is available for cleaning. The change over of bodies takes place after each 5 hours of running. The body change over sequence is given in **Appendix-III/3.**

The system is supplied by manufacturer for 8 hrs running cycle, due to the problem of scaling; the running hours came down to 5hrs. The analysis is carried to find out the nature of scale formation. The samples of scale had been collected during the washing period of the effects; the results are given in **Table (23),**

Table (23): Analysis of Scales from 2nd effect of MEE

Particulars	1st Hour of washing	4th hour of washing
Calcium (as $CaCo_3$) % by Wt	1.2	0.68
Magnesium (as $MgCO3$) % by Wt	0.46	0.09
Sodium sulphate (as such) % by Wt	8.71	1.3034
Acid insoluble % by Wt	3.08	2.06
Silica % by Wt	2.07	-
Loss of ignition % by Wt	35.31	31.28

The discussion with referred to above table; it is found that natures of scales are water soluble. Where the hard scales due to $CaCO_3$ is less in percentage compared to water soluble scales. The system behavior at beginning of each cycle is observed (i.e. after each boil outs) and compared with behavior after periodic cleaning (i.e. once in two months pressurized water cleaning), the deviation is less. So study is carried on controlling the water soluble scales.

The first critical solid point of black liquor evaporation of present system using the white liquor analysis in **Table (10)** is calculated and results has shown with effect of variation in causticizing efficiency and EA / Wood ratio on critical black liquor solids, is shown in **Figure (26,27)**. And assuming the brown stock yield to be steady of 46%.

The solubility of sodium salts from black liquor is also predicted using white liquor analysis in **Table (10)**, the results are shown in **Figure (28,29)**.

Figure (26): Critical Solids %TS Vs EA/Wood Ratio

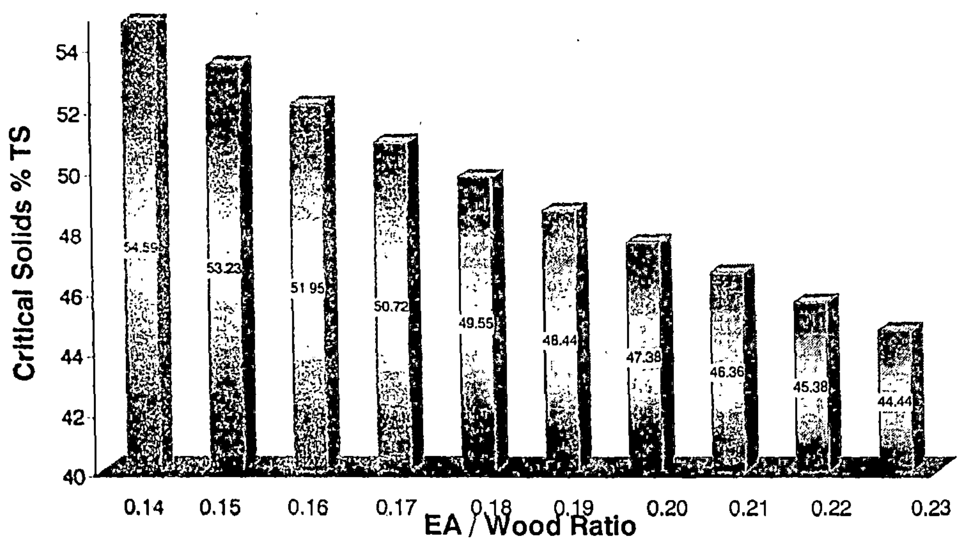


Figure (26) shows predicted relationship between critical solids and the EA / wood ratio. The EA / wood ratio is a pulping parameter driven on bases of pulp quality not on black liquor evaporation. The present EA / wood ratio is 0.165, the burkeite begins to precipitate at 50% BLS. At particular case EA / wood ratio of 0.2, the burkeite begins to precipitate at about 2.62% lower than the present condition.

Figure (27): Causticizing Efficiency, % Vs Critical solids % TS

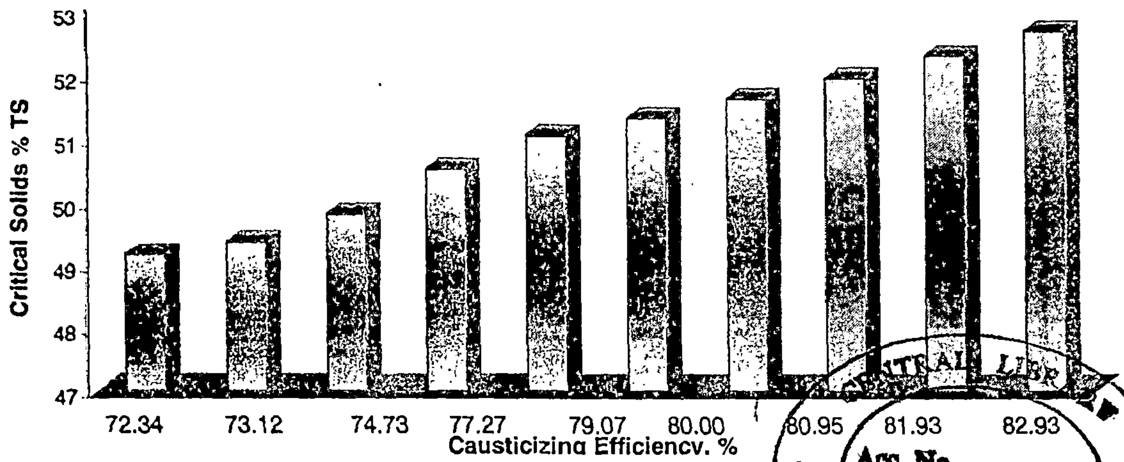


Figure (27) shows predicted relationship between critical solids and the causticizing efficiency. The relation is predicted at present EA / wood ratio of 0.165. The present causticizing efficiency is 80%. At particular case of 77% causticizing efficiency, the burkeite begins to precipitate at about 0.51% lower than present condition.

Figure (28): Solubility of Baurkeite in black liquor

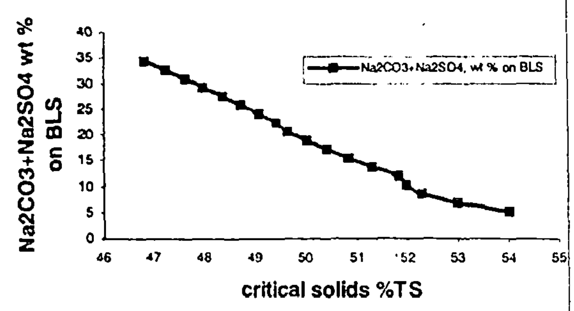
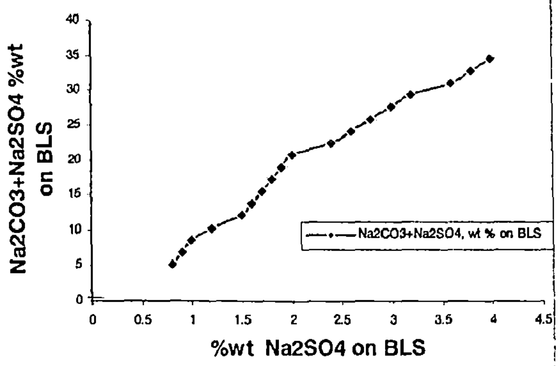


Figure (29): Effect of Na2SO4 % in black liquor solids



Based on the predicted results shown in **Figure (28,29)**, the conclusion drawn as,

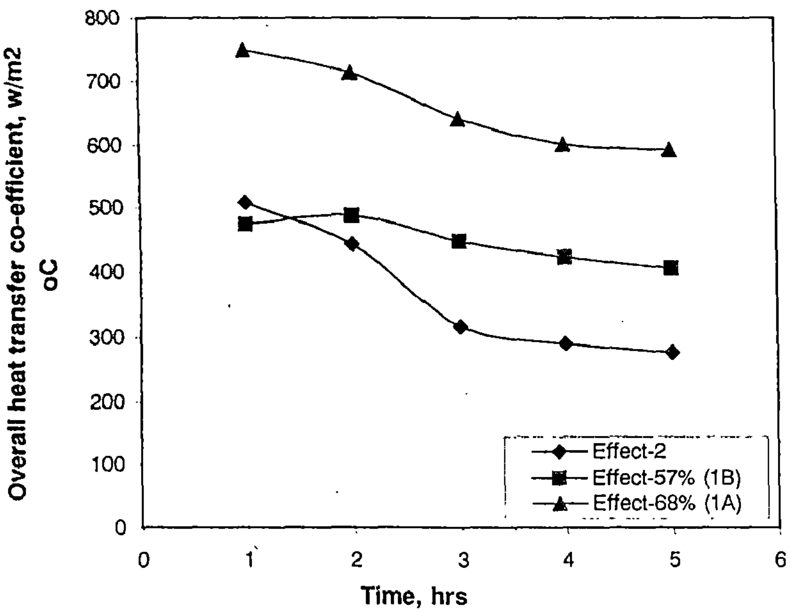
- At critical solids point of 49-52% the fall of Na_2CO_3 + Na_2SO_4 % wt on BLS is higher than 47-49% solids point.
- Where the settlement rate of Baurkeite is higher at critical solids point of 49 to 52% is predicted.

The samples of black liquor, feed to 2nd, feed to 1st effect, and feed to HBL tank is collected from the time of body change over to next change over time. At the same, the performance of the body is noted; calculations are shown in **Appendix-III/2,** the system behavior is shown in **Figure (30)**. The Na_2CO_3 and Na_2SO_4 are tested from the collected samples, to know the rate of precipitation of sodium salts from the black liquor. The results are such,

Table (24): Composition of Na_2CO_3 and Na_2SO_4 in concentrated black liquor from effect 1 & 2.

Particulars		Effect-2			Effect-1		
Time in hrs		1	2	3	1	2	3
Solids, %TS	In	40.13	40.31	42.77	49.48	48.83	50.65
	Out	49.48	48.83	50.65	66.30	64.20	63.00
Na_2CO_3, % wt on BLS	In	20.54	21.81	18.08	18.96	16.55	15.78
	Out	18.95	16.56	15.78	11.73	11.38	14.96
Na_2SO_4, % wt on BLS	In	6.35	7.05	5.75	6.35	3.17	3.81
	Out	6.11	3.17	3.81	3.57	2.62	2.31
Na_2CO_3 / Na_2SO_4, % wt on BLS	In	3.36	3.09	3.14	2.99	5.22	4.14
	Out	2.98	5.22	4.14	3.29	4.34	6.48

Figure (30): Overall heat tansfer co-efficient for the present 5hrs running cycle, w/m2 deg. C



Based on analysis shown in **Table (24) & Figure (30)**, the following conclusion drawn as,

- The samples concentration levels are in the range of predicted critical solids points.
- The ratio of Na_2CO_3 to Na_2SO_4 feed to 2nd effect is 3.5. It is on higher side.
- The system overall heat transfer co-efficient fall with respect to precipitation rate of sodium salts from black liquor evaporation is uniform.
- The sodium salts precipitation is higher in mid hours of 2nd effect and in beginning hours of 1st effect.(i.e.),

  In effect 2 – Na_2CO_3 is 5.25 % wt on BLS and Na_2SO_4 is 3.88 % wt on BLS.

  In effect 1 – Na_2CO_3 is 7.23 % wt on BLS and Na_2SO_4 is 2.78% wt on BLS.
- Reason of baurkeite formation is due to,
  - Lower causticizing efficiency
  - Higher Na_2SO_4 in black liquor composition.

Recommended practice

The most important process in keeping the heating surfaces clean by controlling the crystallization of sodium salts in black liquor. Recommendations to control soluble scale deposition,

- Maintaining low levels of Na_2CO_3 and Na_2SO_4 in the black liquor at 5% and 15% wt on BLS.

- Maintaining the Na_2CO_3 to Na_2SO_4 ratio less than three.

Control by addition of recycled chemical streams, e.g., makeup sulfate streams such as salt cake, spent acid from ClO_2 plant. Even fly ash from SRB can be added. By controlling the crystallization of black liquor, compared to present evaporation cycle time of 5 hours it get improves to 6 to 8 hours cycle time. The present and recommended cycle time of evaporation details are given in **Appendix-III/3.**

Figure (31): Predicted overall heat transfer co-efficient for 6hrs operating cycle

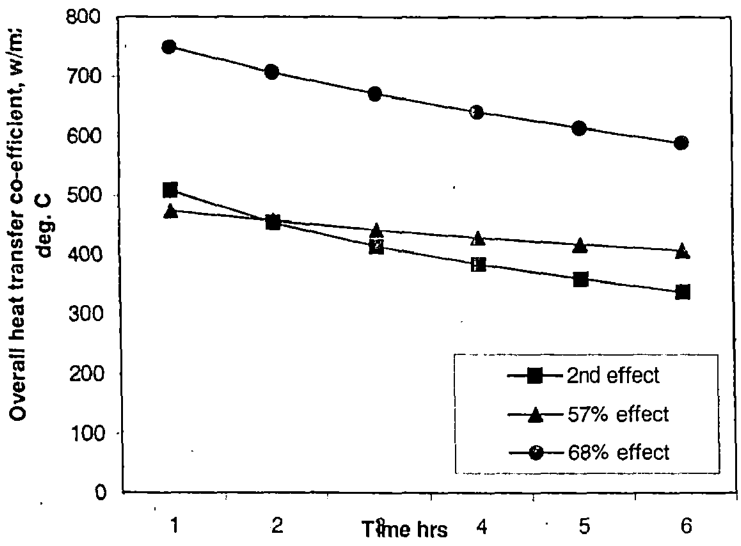
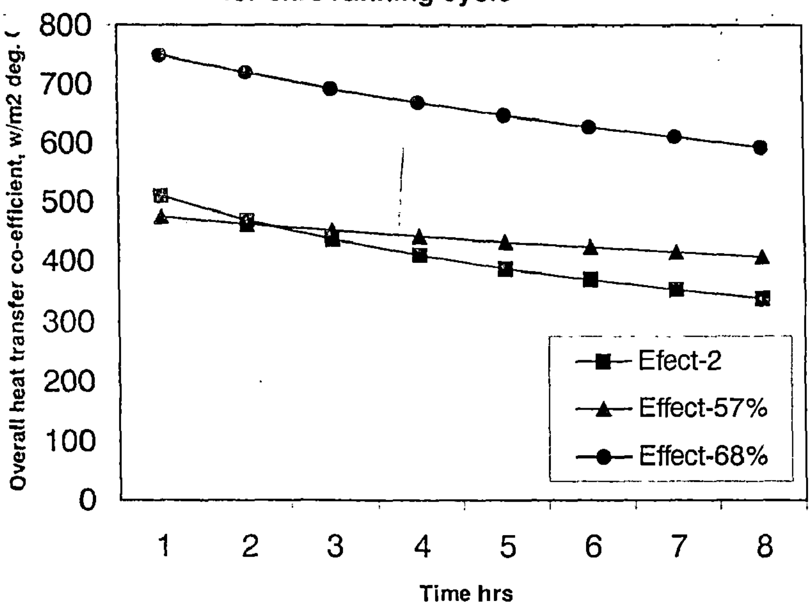


Figure (32): Perdicted overall heat transfer co-efficient for 8hrs running cycle



The system overall heat transfer values are predicted by using straight line equation (15) $1/U^2 = a\theta_b + d$, where a and d are constants for any given operation and U is the overall heat transfer co-efficient at any operating time θ_b, for possible improvement in running hours of system to 6 and 8 hrs. Assumption made during the prediction is by considering the present behavior of system. The predicted overall heat transfer co-efficient of the system at 6hrs and 8hrs running cycle is shown in **Figure (31,32)**.

The present parameters and anticipated parameters after implementing the proposal are listed in the table below.

Table (25): Improvement in performance of MEE by controlling the water soluble scales

Particulars	Performance value for existing process	New performance values by retrofitting	Remarks
Na_2CO_3 / Na_2SO_4 % wt on BLS, at beginning of cycle	3.36 (Enter to 2nd effect)	2.84 (**By addition of spent acid from ClO_2 plant and sodium sulphate as make up)	Decreasing in deposition rate of sodium salt led to increasing in running hours.
Operation time of evaporation per cycle	5 hours	6 to 8 hours	Increasing in running hours - decreasing in down time and evaporation cost of condensate.
Time to attain product liquor after change over of body	1.25 hrs/day	0.75 to 1 hrs/day	Decreasing in down time, increasing in product liquor production.
** Mill is adding 10 TPD as make up salts before firing as Na_2SO_4. Instead of adding there, adding in system of evaporation to control soluble scales.			

The net savings possible by implementation of this proposal,

Table (26): Energy saving cost by increasing the system availability of evaporation system of black liquor

Particulars	Cycle time of 5hrs	Cycle time of 6hrs	Cycle time of 8hrs
Wash Process condensate addition in WBL tank, m^3/day	250	200	150
Steam requirement to evaporate this water from WBL, TPD	41.12	32.89	24.67
Equivalent cost of thermal energy, Rs. Lakhs per annum	41	32.5	24.4
Time to attain the product liquor after change over of body, Hrs / day	1.25	1	0.75
Equivalent cost of down time in term of steam generation in boiler, Rs. Lakhs per annum	99	79.20	59.40
Total cost per annum, Rs lakhs	140	111.70	83.80
Net saving possible per annum, Rs. Lakhs	-	27.95	55.89

4.6. Indirect heating of heavy black liquor up to firing temperature:

Present practice:

The heavy black liquor (HBL) at 100.5^{0}C, 67% TS from the ash mixing tank get fired in the recovery boiler through the black liquor heater, where the black liquor heated to required firing temperature recommended by manufacturer (120^{0}C). The heating is carried out by direct heater with MP steam at pressure of 10.5 bar.

Recommended practice:

By indirect heating of heavy black liquor, eliminates the solids dilution occurring with direct steam mixing, thus safer firing condition exist (the low solids concentration can lead to blackouts or smelt/water explosions). Indirect heating also avoids additional water evaporation during the combustion process, and therefore increases the net efficiency to steam. The condensate after heating the black liquor also gets recovered. The detail calculation is shown in **Appendix-IV/1.**

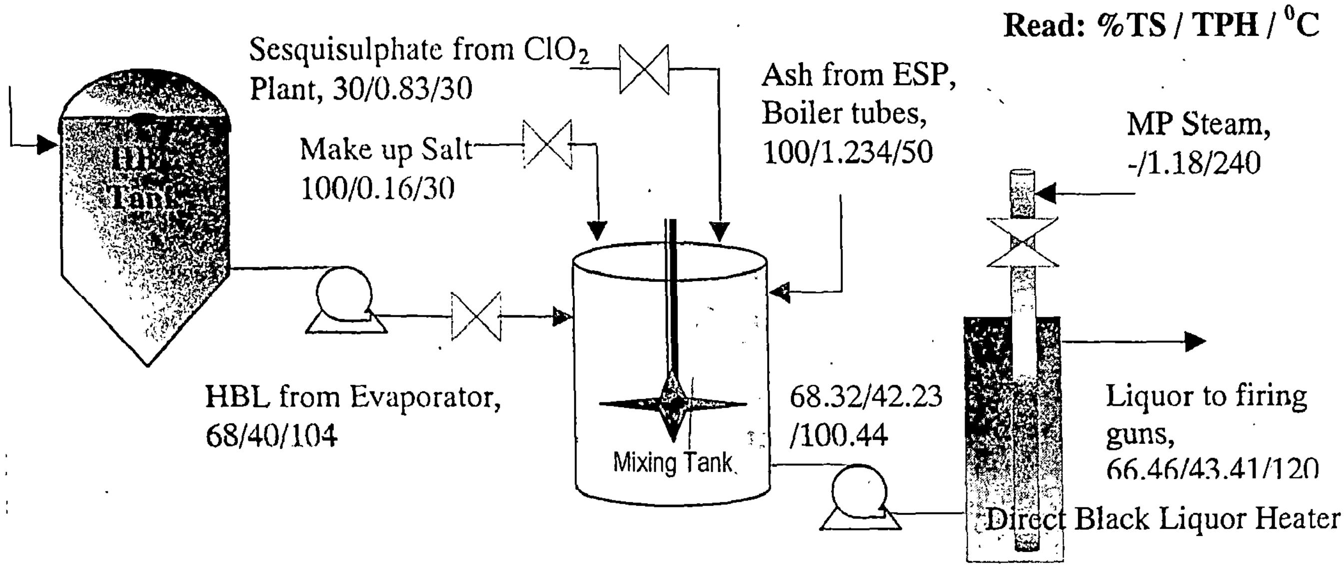


Figure (33): Present Practice of heating black liquor up to firing temperature

The performance improvement by indirect black liquor heating up to firing temperature compared to existing performance is given in **Table (27)**

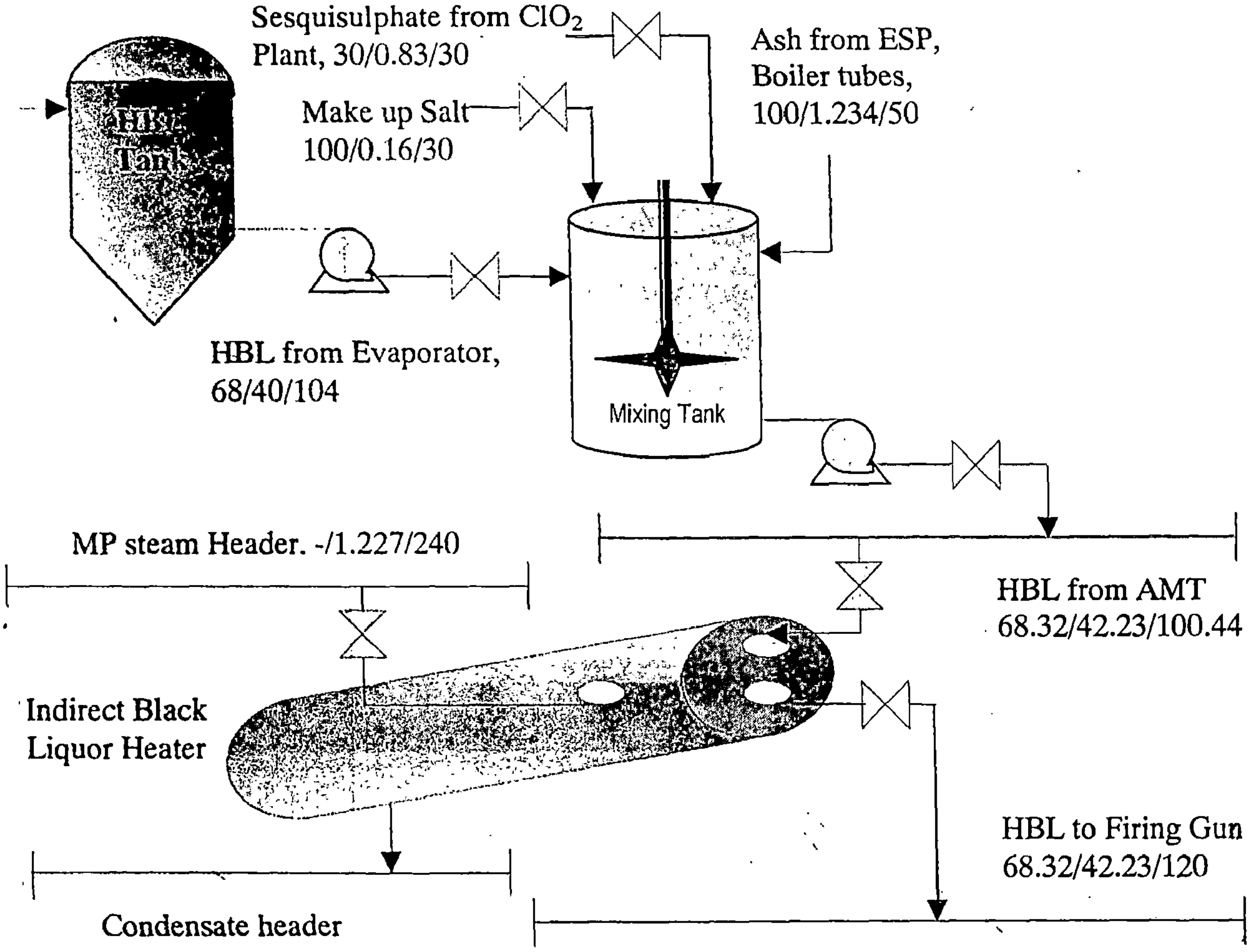


Figure (34): Recommended practice for heating black liquor up to firing temperature

Table (27): Improvement in performance of chemical recovery process by applying indirect heating of black liquor up to firing temperature

Parameters	Performance value for existing process	New performance value by retrofitting the indirect heating of black liquor	Remarks
Concentration of black liquor to firing, %TS	66.46	68.32	Black liquor dilution reduces by 1.8%.
Steam generation in recovery boiler, tph	78	79.19	Heat loss to flue gas reduced by 8.13%.
MP steam required for heating to 120^{0}C, tph	1.18	1.227	3.8% increase in steam consumption for heating BL up to firing temp. used steam recovered as condensate.

Possible savings by improvement,

Net saving = Rs. 29.16 lakh per annum

Investment = Rs. 12 lakh per annum

Simple pay back period = 5 months

4.7. Effect of adding neutralized sesqui sulphate crystal (i.e. Spent acid from ClO_2 plant) in black liquor as make up sodium sulphate and increasing in steam generation.

Present practice:

The WBL from pulp mill at 18% TS get concentrated in evaporator to 66% TS is get fired in recovery boiler after addition of make up chemicals in ash missing tank. Here the make up chemical addition is partially with sodium salts and spent acid from ClO_2 plant (sodium sesqui sulphate $Na_3H(SO_4)_2$). The sodium salt addition is about 6 TPD and sodium sesqui sulphate is about 4 TPD at 30% TS.

The spent acid from the ClO_2 plant gets filtered from the generator liquor as a moist crystal and is subsequently redissolved in hot water, the solution of spent acid at pH of 1-2 is sent to SRP as make up salt. Where the black liquor get diluted to 0.55 TPH. Since the pH level is low, the handling system of the spent acid getting corroded. It has observed that the pH level of black liquor before and after addition is remaining same.

Recommended practice:

The moist sesqui sulphate crystal (@10% moisture) is mixed with sodium carbonate, the residual acid gets react with sodium carbonate to form sodium sulphate. The crystal sodium sulphates get feed as make up salt to SRP.

The laboratory test has conducted at different weight addition of sodium carbonate, to check the pH level and for any residual sodium carbonate level. The test results as such, Details of sesqui sulphate crystal from ClO_2 plant,

Particulars	Sesqui sulphate from ClO_2 plant
Solids, %	92.85
pH @ 10% of solution	1.20
Na_2SO_4 % wt as such	55.5
NaCl % wt as such	0.088

Laboratory test result at different weight addition of sodium carbonate,

Table (28): Laboratory results by addition of sodium carbonate with spent acid from ClO_2 plant

Particulars	Sample-1		Sample-2		Sample -3	
	Na_2CO_3, gms	$Na_3H(SO_4)_2$, gms	Na_2CO_3, gms	$Na_3H(SO_4)_2$, gms	Na_2CO_3, gms	$Na_3H(SO_4)_2$, gms
Mix of sesqui sulphate and sodium carbonate	20	100	22.5	100	25	100
pH @ 10% solution	4.3		9.46		9.7	
Solids %	86.18		89.23		89.35	
Na_2SO_4, % wt as such	98.12		97.35		96.2	
Residual Na_2CO_3 % wt as such	-		2.56		3.71	

Formation of sodium sulphate from crystals of sodium sesqui sulphate occurs by means of following reaction,

$$2Na_3H(SO_4)_2 + Na_2CO_3 \longrightarrow 4Na_2SO_4 + CO_2 + H_2O$$

The optimized amount of Na_2CO_3 required is 0.225 tonne per tonne of sesqui sulphate production from ClO_2 plant to give sodium sulphate as make up salt.

System Balance, **Read: %TS / TPH / ^{0}C**

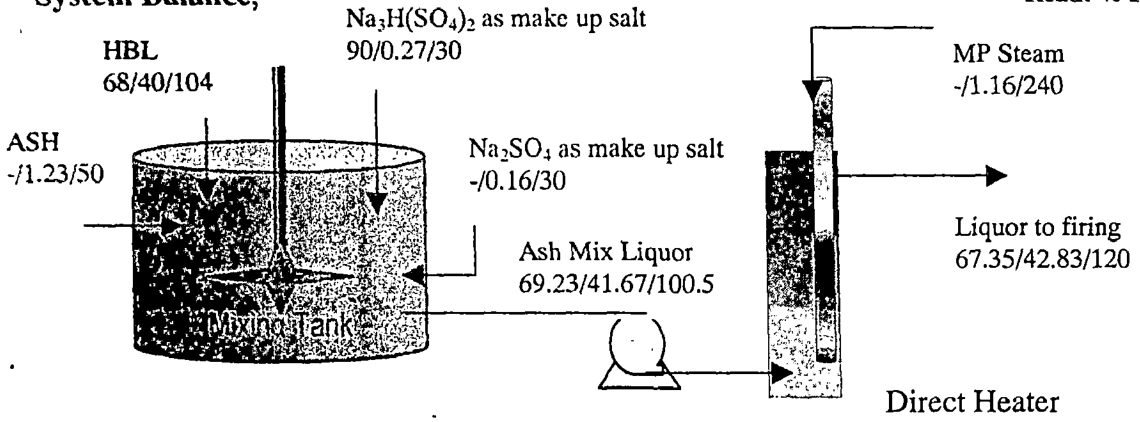


Figure (35): System balance of black liquor handling system in SRP by addition of $Na_3H(SO_4)_2$ crystal as Na_2SO_4 make up salt.

The mixing of sesqui sulphate crystal with sodium carbonate is to be carried out with concrete mixer to form sodium sulphate crystal. The present performance and retrofitted performance after implementing the proposal are listed in the **Table (29),**

Table (29): Process improvement by avoiding dilution of firing black liquor

Parameters	Performance values for existing process	New performance values by avoiding dilution of black liquor	Remarks
BL firing % conc. In boiler	66.46	67.35	Increase in black liquor firing concentration by 1.32% reduces the heat loss to flue gas by 4%.
Sesqui sulphate moisture %.	70	10	Decrease in moisture level, decreases the dilution of black liquor by 1.32%.
Steam generation in boiler, TPH	78	78.56	0.56 TPH increase in steam generation

Possible savings by improvement,

Annual savings in steam generation @ Rs. 300 per tonne of steam = Rs.13.3 lakh

Investment = Rs. 10 lakhs as per study taken in BILT Yamunanagar unit.

Simple pay back period = 9 months.

4.8. Optimization of cooling water requirement in surface condenser:

Present practice

The evaporated water vapor from the black liquor condensed in surface condenser. Present cool water consumption to condense the vapor at vacuum level of 650 mm of Hg is about 2050 m^3/hr, the vapor handling is about 23 MT/hr at 56^0C. For the above flow rate pump head is about 33 mwc and motor input power of 300 kW.

To measure the cooling water flow across the surface condenser, the orifice meter of 353 mm bore is placed in 600 mm cooling water pipe line, where it has observed that the pressure drop across the orifice meter is about 4 mwc, the equivalent motor input power to pump is about 32.3 kW at 2050 m^3/hr flow.

Recommended practice

The actual design flow recommended is 1600 m^3/hr, by this surface condenser vacuum is falling. The flow variation in system is carried and its effects are measured, the variations are shown in Table (30) at different flows. From the system behavior it has observed that 1700 m^3/hr is optimum without any changes in system vacuum level.

Table (30): Present system behavior at different flow rate across surface condenser

Flow, m_3/hr	Vacuum in SC, mm of Hg	Delta T across SC, ^{0}C	Cooling water Pump head, mwc	Vapor flow in SC, kg/hr	Motor input power, kW	Pump efficiency, %
1675	640	8.0	44	23532	295	61
1750	647	8.3	44	25887	290	65
2000	647	6.9	34	24614	300	55
2050	646	6.3	33	23000	300	56

It has been observed that 1700 m^3/hr is sufficient to condense the vapor. The requirement pump head for the above flow is calculated. In the calculation the head of the present pump delivery, is verified,

Table (31): Head loss calculation for different flow rate in surface condenser of black liquor evaporator plant.

Flow rate, m3/hr	Flow velocity, m/sec	Flow velocity, ft/sec	Head loss in Disch. Line of pump, m	Head loss in CW inlet line, m	Head loss in CW outlet line, m	Head loss in cond. Inlet line, m	Head loss in cond. outlet line, m	Head loss across condenser, m	Head loss across the orifice meter, mwc	Static head, m	Total CW pump head, m
1600	1.57	5.14	0.25	1.05	1.67	0.55	0.30	10	4	11	28.82
1650	1.62	5.30	0.26	1.11	1.77	0.55	0.31	10	4	11	29.00
1700	**1.67**	**5.46**	**0.28**	**1.17**	**1.87**	**0.56**	**0.31**	**12**	**4**	**11**	**31.19**
1750	1.72	5.62	0.30	1.24	1.97	0.56	0.32	13	4	11	32.38
1800	1.77	5.79	0.31	1.31	2.08	0.56	0.32	13	4	11	32.58

Flow rate, m^3/hr	Flow Velocity m/sec	Flow Velocity Ft/sec	Head loss In disch. Line of Pump, m	Head loss In CW inlet Line, m	Head loss In CW Outlet Line, m	Head loss In cond. Inlet Line, m	Head loss In cond. Outlet Line, m	Head loss across Condenser m	Head loss across the Orifice Meter, mwc	Static head, m	Total CW pump head, m
1850	1.82	5.95	0.33	1.37	2.19	0.57	0.32	13	4	11	32.78
1900	1.87	6.11	0.35	1.45	2.30	0.57	0.33	14	4	11	33.99
2000	**1.97**	**6.43**	**0.39**	**1.59**	**2.53**	**0.58**	**0.33**	**14**	**4**	**11**	34.42
2050	2.02	6.59	0.41	1.67	2.65	0.58	0.34	14	4	11	34.65

Actual required pumping energy for the above tabulated flow as follows:

Table (32): Pumping power requirement at different flow rate (Cooling water pump of surface condenser)

Flow rate, m3/hr	Total CW pump head, m	Pump output power, kW	Pump efficiency, %	Shaft power kW	Motor efficiency, %	Motor input power, kW
1600	28.82	125.6552	75	167.54	90	186.1559
1650	29.00	130.3913	75	173.86	90	193.1722
1700	**31.19**	**144.4877**	**75**	**192.65**	**90**	**214.0558**
1750	32.38	154.4121	75	205.88	90	228.7587
1800	32.58	159.8049	75	213.07	90	236.748
1850	32.78	165.2522	75	220.34	90	244.818
1900	33.99	175.9832	75	234.64	90	260.7159
2000	**34.42**	**187.589**	**75**	**250.12**	**90**	**277.9096**
2050	34.65	193.5636	75	258.08	90	286.7608

The economic analyses of the present and recommended flow are as follows:

Table (33): Energy saving cost with present running pump and with retrofitted pump (Cooling water pump of surface condenser)

Flow rate, m3/hr	Motor power with present pump, kW	Motor power with exact requirement, kW	Power saving, kW	Annual saving with exact capacity pump, Rs. lakh @ Rs.3.25 per kWh	Present power saving possible with recirculation rate of 1700 m^3/hr, kW	Annual saving with present pump, Rs. lakh @Rs. 3.25 per kWh
1700	290	214.1	75.9	19.53	10	2.57

The possible savings with 1700 m^3/hr flow by actual pump requirement as,

Annual savings = Rs. 19.53 lakh

Investment = Rs. 2.5 lakh

Simple payback period = 1.5 months.

The savings with 1700 m^3/hr flow with present running pump as, Annual savings = Rs. 2.57 lakh

Investment = Nil

Instead of measuring the flow through orifice, the flow can be controlled by delta T across the surface condenser. The set point for the cooling water flow auto valve is to set according to the delta T across the surface condenser.

Table (34): Energy Saving Cost by avoiding pressure drop by orifice meter

Particular	Condition
Pressure drop across orifice meter, mwc	4.0
Measured flow, m^3/hr	2000
Equivalent power consumption by pump @ 75% pump eff., kW	32.3
Equivalent power cost per annum @ Rs. 3.25 per kWh	8.31 lakhs

By avoiding the orifice meter across the surface condenser inlet flow and arranging the flow control by delta T across the surface condenser, the benefit as,

Annual savings = Rs. 8.31 lakhs

Investment = Nil

4.9. Smelt dissolving tank vapor recovery:

Present Practice:

After firing the black liquor in recovery boiler, the inorganic content flows out the boiler as smelt about 800^0C in to the smelt dissolving tank (SDT). To avoid dissolving tank explosions, shatter jets, MP steam jets, are used to break up the stream of smelt exiting the recovery boiler into small droplets. Where the temperature of smelt brought down to 95^0C by dissolving in weak white liquor available from causticizing section. Temperature drop of smelt in SDT results in large amount of vapor generation, which is venting out through stack. The generated vapor will contain both the condensable and non-condensable vapor (TRS, etc.,) and particulate matter.

Recommended Practice:

The generated vapor from SDT can recover by passing the vapor through jet type of condenser, then foul condensate at higher temperature from the condenser exchange heat to clean water by heat exchanger application, the hot clean water taken for process application (causticizing section). The foul condensate after the exchanging the heat, reused for further condensing the vapor in the condenser. The non-condensable vapors are vented out. The detail of calculation is given in **Appendix-IV/2.**

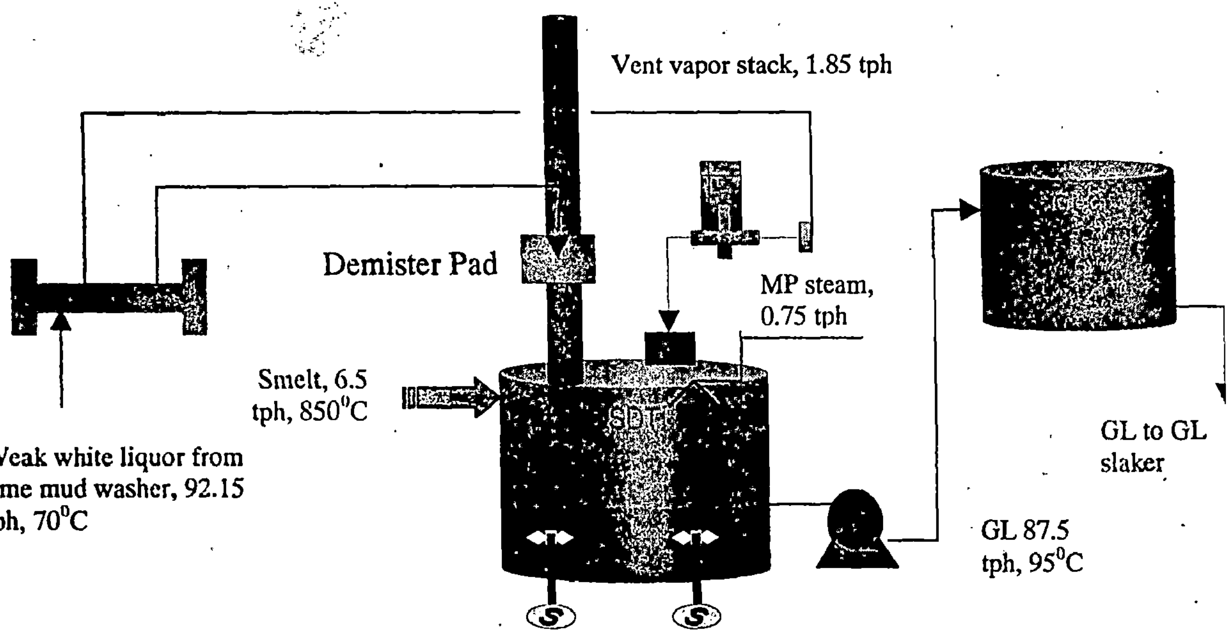


Figure (36): Present practice of SDT vent vapor handling

The other options to recovery the vapor by passing it through indirect condenser, where the vapor exchanges its heat to the process condensate water from evaporator. Benefit by this option, handling of the foul condensate is not required. But the care has to be taken in designing of heat exchanger, because of corrosive vapor handling. The detail of calculation is given in **Appendix-IV/2.**

Possible savings by improvement,

Option-1 (Vent vapor recovery by direct contact condenser)

Annual Savings= Rs. 56.12 lakh

Investment = Rs. 25 lakh

Simple Pay back period = 6 months

Option-2 (Vent vapor recovery by indirect condenser)

Annual Savings = Rs. 37.87 lakh

Investment = Rs. 20 lakh

Simple Pay back period = 7 months

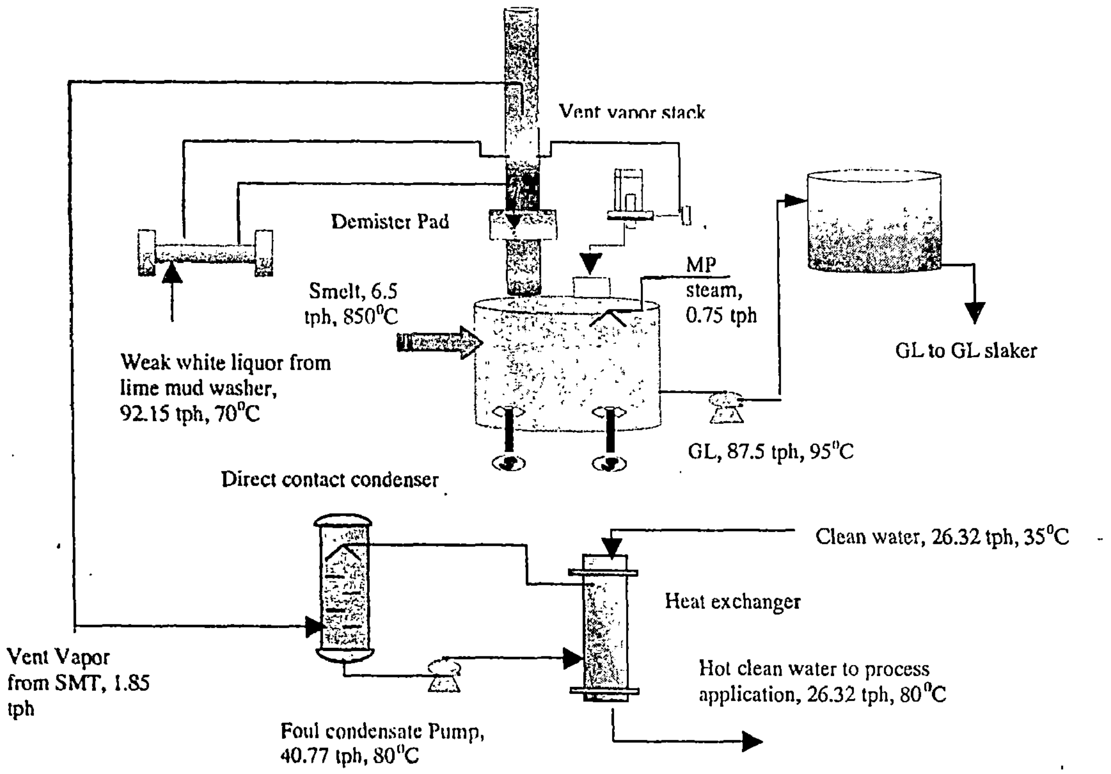


Figure (37): Recommended practice of SDT vapor handling (Option-1)

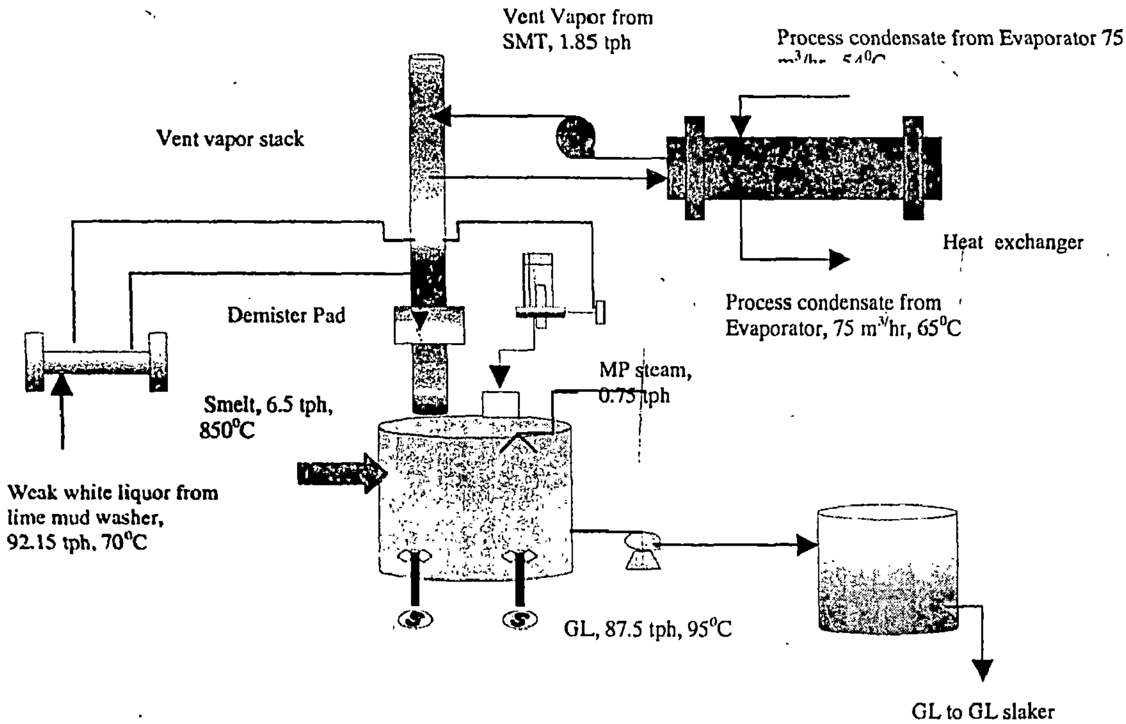


Figure (38): Recommended practice of SDT vapor handling (Option-2)

4.10. Boiler blow down and Air preheater condensate tank flash vapor recovery

Present Practice:

To maintain the quality (i.e. silica level, conductivity, and pH) of the boiler water, continuous blow down @ 3% of total steam production is in practice. This will be carried out at boiler drum pressure of 70 bar (abs), 275°C. The blow down water is get collected in blow down tank and get flashed from above pressure to atmosphere pressure, the vapor are vented out. The combustion air is admitted to primary and secondary port through the primary and secondary scaph, where the combustion air is preheated to 160°C across the primary scaph and 150°C across the secondary scaph. The preheating of air is carried partially with LP and MP steam. The condensate are collected in scaph condensate tank, where the condensate get flashed to atmosphere pressure, the vapors are vented out.

Recommended Practice:

The vent vapors from the blow down tank and scaph condensate tank can get conserve by preheating the feed water to deareator, where the LP steam consumption rate in deareator will come down. The other option is by compressing the vapor from atmospheric pressure to required LP steam pressure using the motive MP steam in thermo compressor (TC), where the pressurized steam get use back in air preheaters (scaph) and deareator. The detail calculation is given in **Appendix-IV/3.**

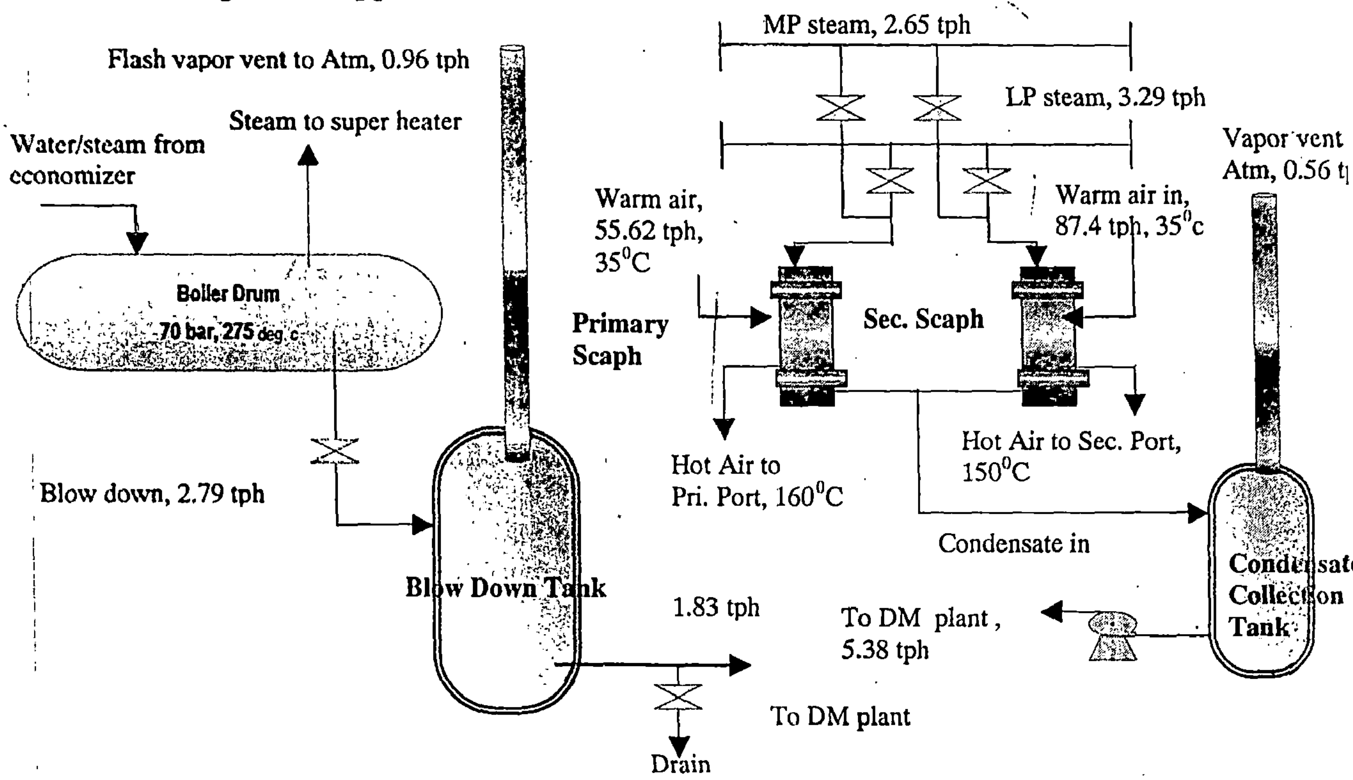


Figure (39): Present practice of handling of flash vapor from scaph and boiler blow down condensate tank

Option-1

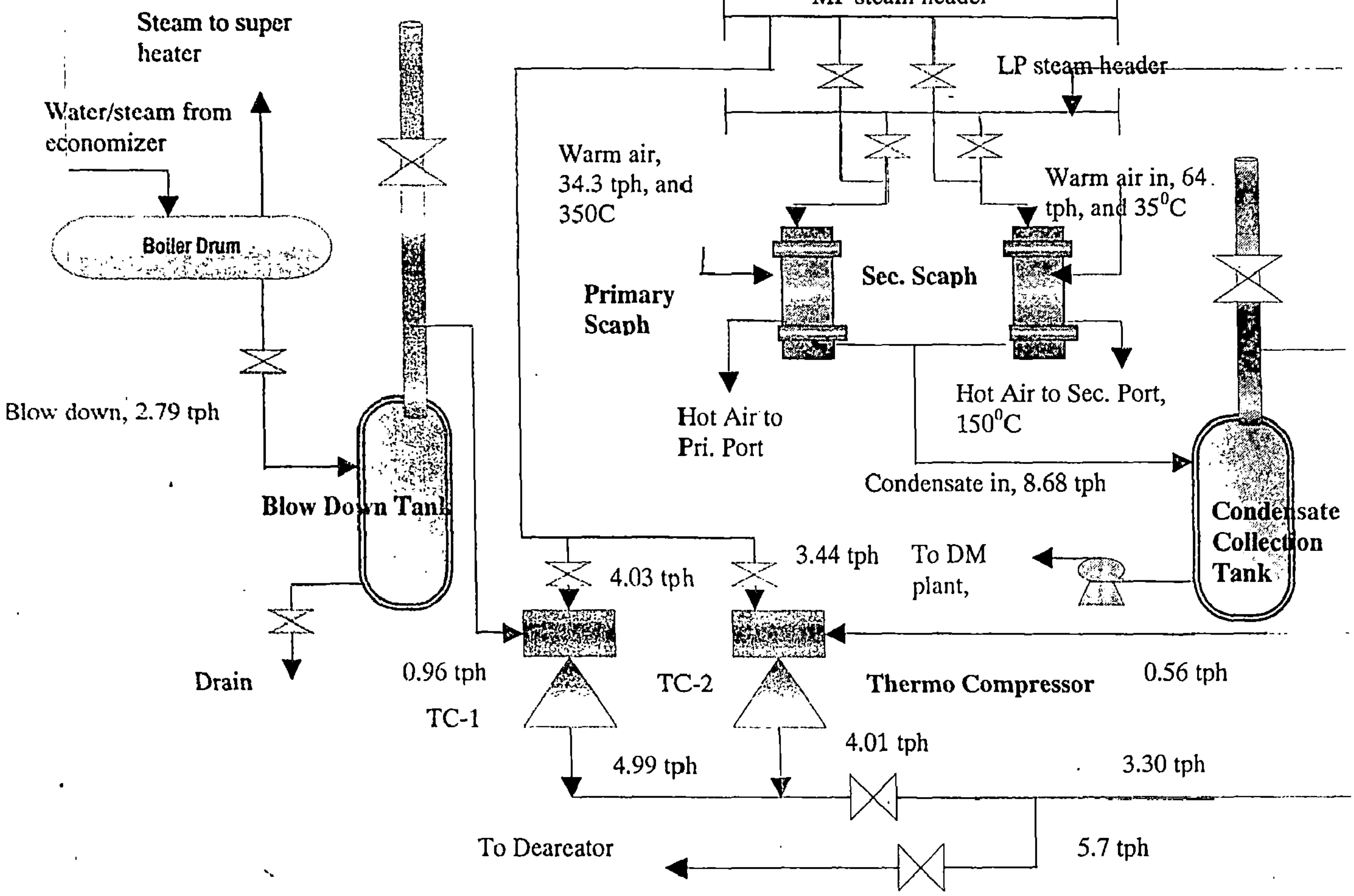


Figure (40): Recommended practice of handling of flash vapor from scaph and boiler blow down condensate tank (Option-1)

From the **Figure (40)** it has observed that, by compressing the flash vapor at atmospheric pressure to pressure of 4.5 bar reduce the LP steam consumption in air preheater by 3.3 tph and in deareator by 5.7 tph with motive steam consumption of 7.47 tph at 11 bar. Where the system will provide lower thermal efficiency of 15% and 23%.

From the other option shown in **Figure (41)**, the feed water flow to deareator get preheated in heat exchanger by flashed vapor reduce the LP steam consumption in deareator by 1.6 tph. Where this system will provide higher thermal efficiency of 81.35%.

Possible savings by improvement,

Option-1 (Reducing the LP steam consumption in scaph by recompressing the flash vapor)

Annual savings = Rs. 36.25 lakh

Investment = Rs. 10 lakh

Simple payback period = 4 months

Option-2 (Preheating the boiler feed water by flash vapor, reducing the LP steam in deareator)

Annual savings = Rs. 38.03 lakh

Investment = Rs. 12.5 lakh

Simple payback period = 5 months

Option-2

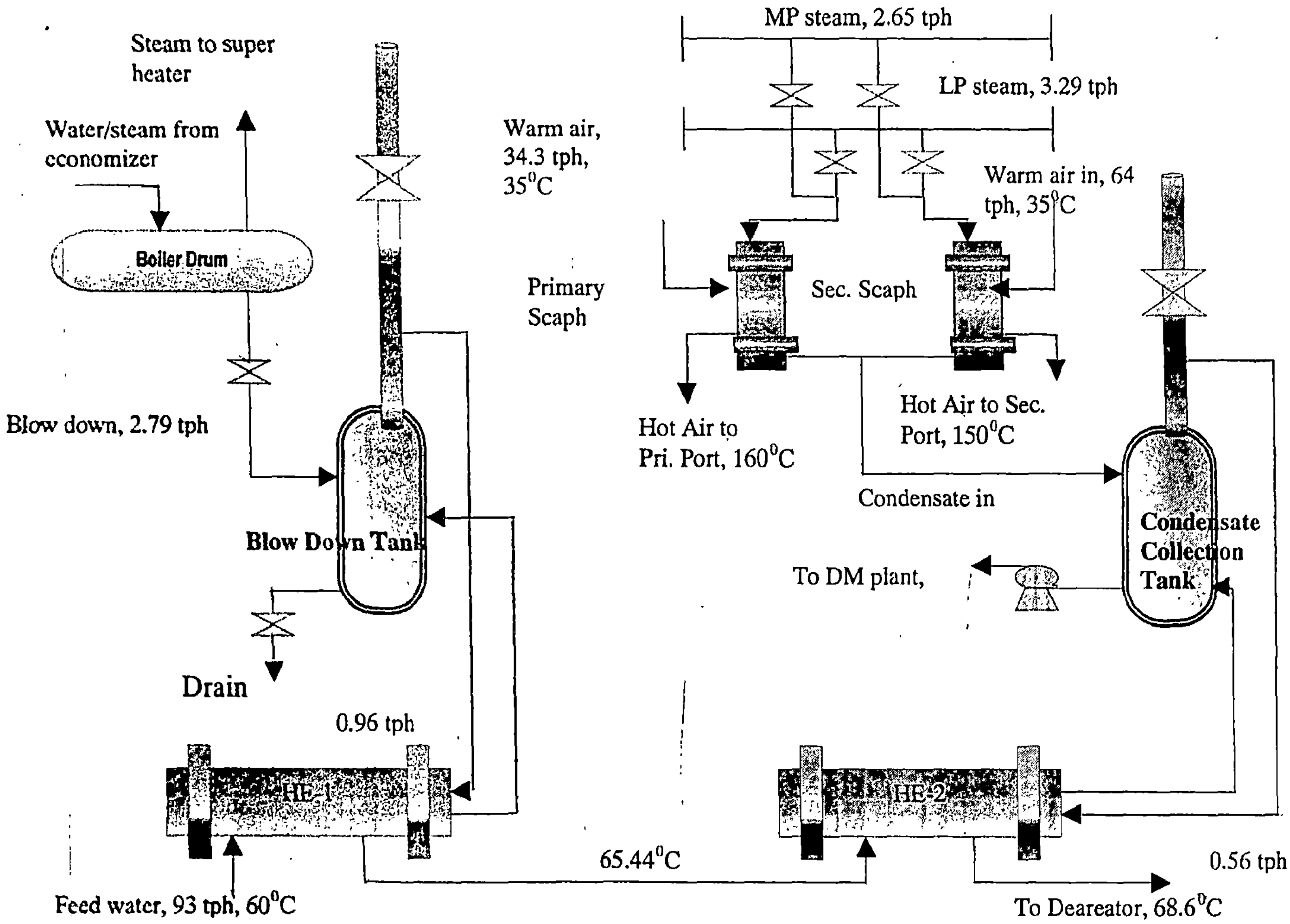


Figure (41): Recommended practice of handling of flash vapor from scaph and boiler blow down condensate tank (Option-2)

4.11. Utilization of condensate available from limekiln oil heater, SRP oil heater and Soot blowing in SRB.

Present practice

In the limekiln, heat is required to calcine $CaCO_3$ to CaO, the heat is supplied by furnace oil at burning zone. The furnace oil is fired at temperature of 128^0C, the furnace oil is preheated from temperature of 81^0C through steam-supplied heat exchanger. In the present practice, condensate from the heater gets drained.

The furnace oil firing is also required in soda recovery boiler, during start up and to maintain desired steam generation. The same practice is existing, as above case.

The soot blow is carried in recovery furnace to maintain clean heat transfer surface of superheaters, & economizers. The soot blowing is about 3.45 hrs in each shift. The arrangement is such that the condensate drain valve remains close during soot blowing. Due to condensate collection in condensate line of soot blowing arrangement, now a days the drain valve remained open during blowing period. The condensate is getting collected in boiler blow down tank, the flash vapor of condensate are venting out.

Recommended practice

The condensate from the both the heater can take it to near by condensate collection tank, where it gets flashed and steam gets conserved and condensate send back to DM plant. The detail calculation is given in **Appendix-IV/4.**

Possible savings by improvement,

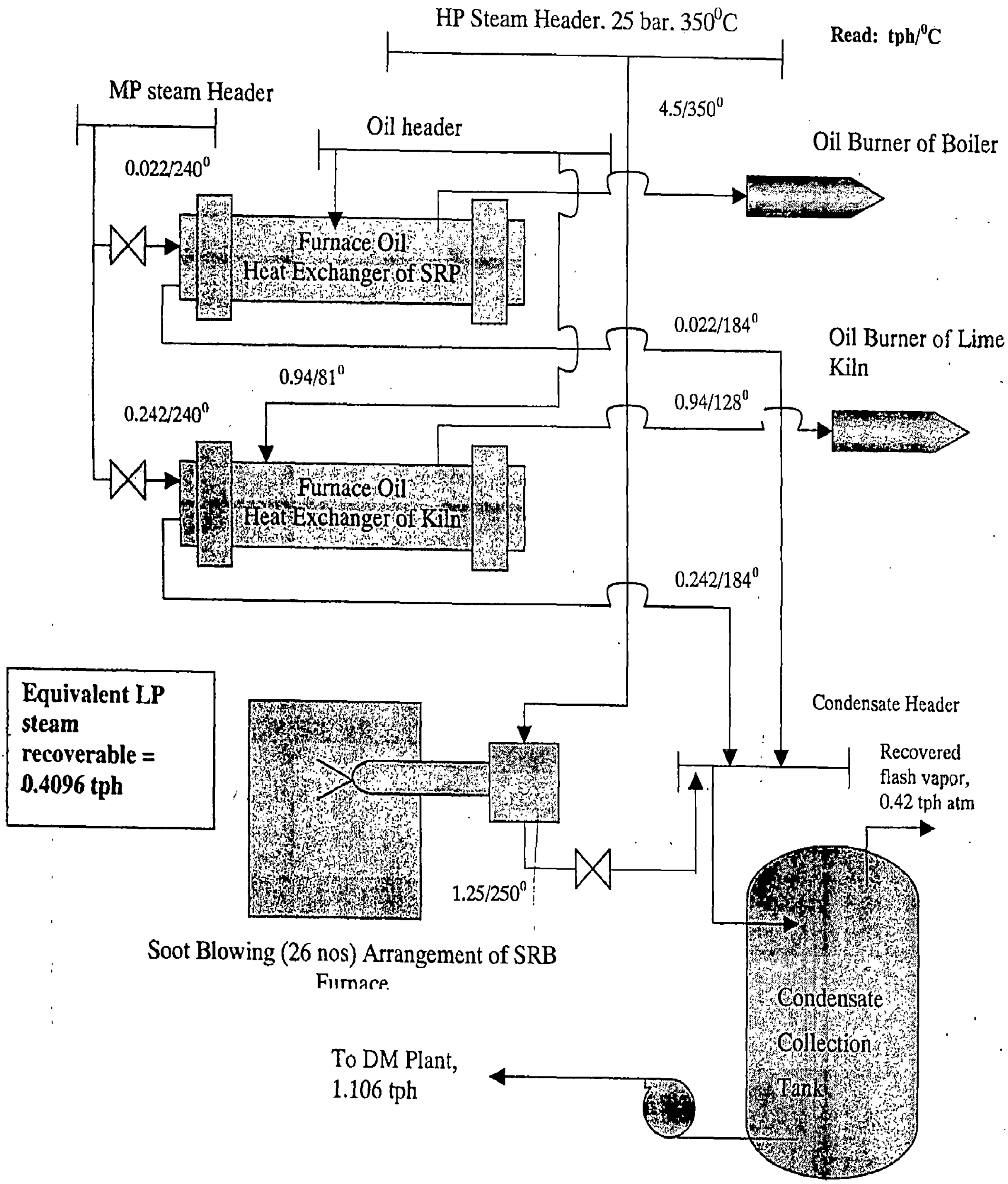
Annual savings = Rs. 10.58 lakh

Investment = Marginal

Simple payback period = Nil

From the **Figure (42)** it has observed that by proper condensate recovery, the possible flash vapor recoverable is about 0.42 tph at atmospheric pressure, its equivalent LP steam is about 0.4 tph.

Figure (42): Recovery of Condensate and Flash Vapor from Lime Kiln Furnace Oil Heater, SRP Furnace Oil Heater, & Soot Blowing Steam



4.12. Reduction of furnace oil consumption in limekiln section:

Present practice

The conversion of lime mud sludge to lime is an important step in the production of white liquor in the chemical recovery process. In essence, the process turns lime mud ($CaCO_3$) in to reburned lime (CaO) and carbon dioxide (CO_2). The lime mud sludge (at 44% moisture) and seashell (at 3.2% moisture) as a make up CaO, are burned in kiln by furnace oil firing at rate of 23 kl/day, to convert in to reburned CaO at purity level of 72%, the reburned lime is used back in green liquor slaker.

Recommended practice

Recommendation is by using the high intensity magnetic field, changing the spin state of the hydrogen molecule of furnace oil. Result in giving same energy level at lower amount of furnace oil consumption.

The working principle of magnetic flux is as,

- Hydrocarbons have basically a "cage like" structure, molecules treated with the magnetic energy tend to de-cluster, creating smaller particles more readily penetrate by oxygen, thus leading to better combustion. They become normalized and independent, distance from each other, having bigger surface available for binding (attraction) with more oxygen (better oxidation). De-clustering of hydrocarbon molecules, in the combustion chamber, increase saturation and reactivity of the fuel mixture with oxygen will achieve, results in a better oxidation of the primary hydrogen element and further oxidation of carbon, the secondary fuel element.
- Accordance to Van Der Waals theory of a weak clustering force, there is a very strong binding of hydrocarbons with oxygen in such magnetized fuel, which ensures optimal burning of the mixture in the combustion chamber.

The device installation is external online without cutting or modifying the fuel pipe line and the magnetic energy generated through the magnetic device is rendered concentric and exactly perpendicular to the flow of the fuel.

Trail run: The trail run made using FLUX MAXIOX (β series) V 2. The fuel consumption rate before and after the installation and the purity level of reburned lime is given in **Appendix-V.**

The present parameters and trial run parameters of the proposal are listed in the table below.

Table (35): Details of pre and post installation of flux maxiox in lime kiln furnace oil burner and its comparative benefits

Particulars	Pre-Installation of Trail	Post-Installation of Trail
Total no of days considered for evaluation, days	47	46
Total product lime production, MT	5883	5836
Total quantity of lime sludge handled at lime kiln, MT	7090	6754
Average moisture in lime sludge during trial, %	44	45
Total quantity of seashell handled at lime kiln, MT	2362	2614
Constant moisture considered in seashell, %	3.2	3.2
Furnace oil consumed during above period, kl	1087	1052
Specific oil consumption, liters per tonne	185	180
Steam consumption across the heater, TPD	5956.5	5764.72
Furnace oil savings, liters per tonne	5	
Steam consumption saving across heater, TPD	191.8	

Possible savings by improvement,

Annual Savings = Rs. 213 Lakhs

Investment = Rs. 2 Lakhs

Simple Pay back period = 0.11 months.

4.13. Improvement in power generation by modification of cogeneration system

Co-generation is widely used in paper mills around the world. Steam generated is used at different pressures and temperatures for cooking of chips in digesters in pulping process and for drying in paper machines. In addition, steam is used for concentration of black liquor in MEE.

Present practice

The present Co-generation system is shown in **Figure (43).** The plant has six boilers and three Turbo Generator with average power generation rate of 27.11 MW. In this Co-generation system CFB-3 is used for generating medium pressure steam at 11 bar and 240^{0}C. Either one of CFB-1 & 2 is used for generating steam at 42 bar, 408^{0}C and passed through extraction cum backpressure turbine for power generation.

Recommended practice

- The medium pressure steam generation boiler CFB-3 can be stopped from operation by installing extraction cum backpressure turbine to CFB-2. From SRB, 31 tph of steam at 42 bar get expanded to process requirement in pressure reducer, this can be taken into extraction cum back pressure turbine. The schematic of proposed measure is shown in the **Figure** (44).
- The CFB-3 is stopped from steam generation remains as stand by boiler.
- The steam balance of the system is balanced as per present system of Co-generation with extra power generation.

System Calculation,

Mass flow of steam from SRB to proposed turbine @ 42 bar = 31 TPH

Mass flow of steam from CFB-2 to proposed turbine @ 42 bar = 16 TPH

Extra coal required to run CFB-2 @ 80% boiler efficiency = 0.26 TPH

Extraction flow from proposed turbine @11 bar = 22 TPH

Back pressure flow from proposed turbine @ 4.5 bar = 25 TPH

Power generation by turbine @ 75% turbine efficiency = 3.09 MW

Annual saving in power generation @ Rs. 3.25 per kWh = Rs. 795.36 lakh

Extra amount in coal consumption per annum @ Rs. 1.5 per kg = Rs. 31.22 lakh

Possible savings by improvement,

Annual Savings = Rs. 764.14 lakh

Investment required = Rs. 600 lakh

Simple Pay back Period = 9 months.

Pay back period,

Depreciation per annum = Rs. 60 lakh

Enterprise tax per annum = Rs. 105.62 lakh

Increased cash flow = Rs. 658.52 lakh

Pay back period = 11 months

Figure (43): Existing Co-Generation System

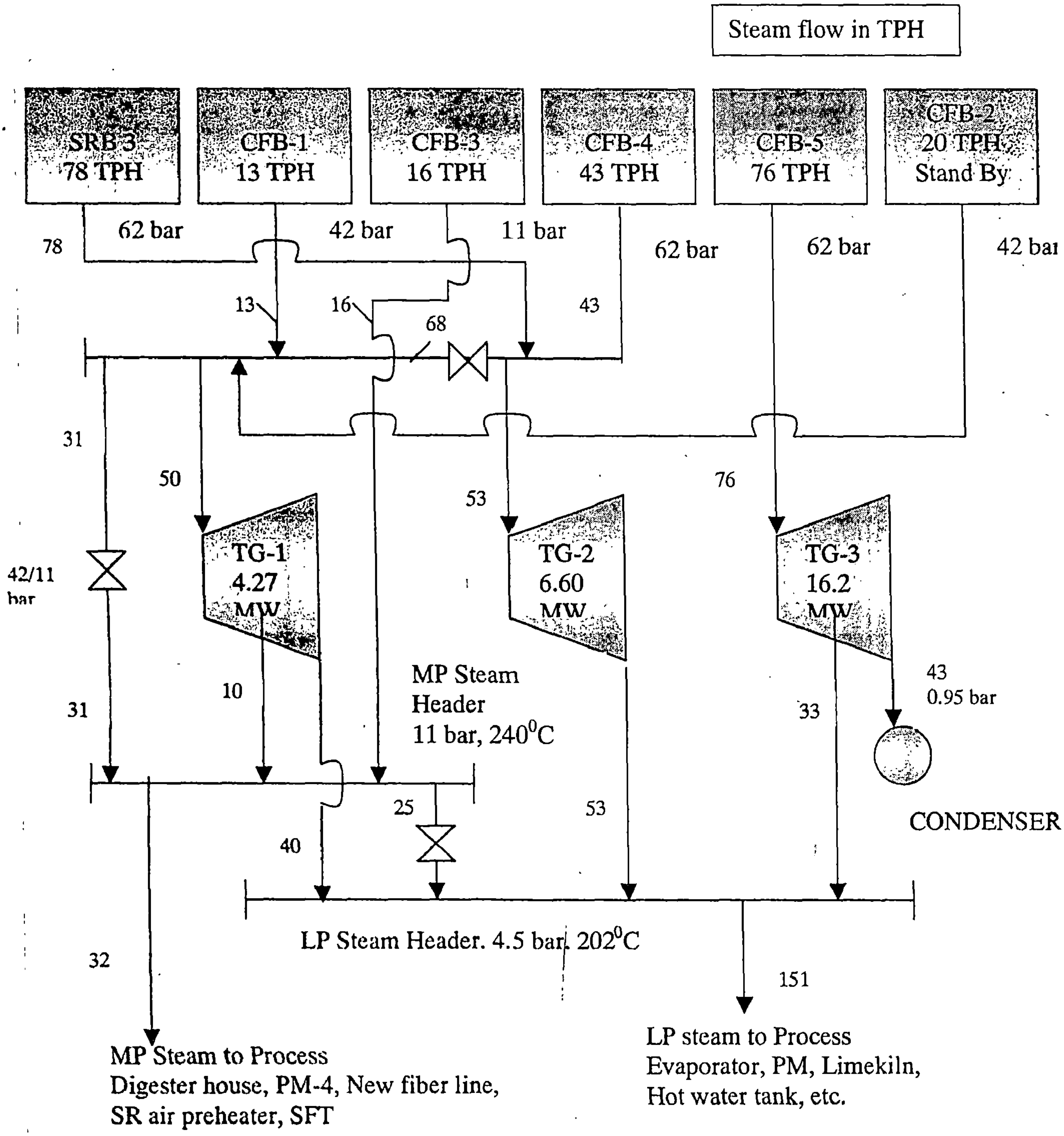
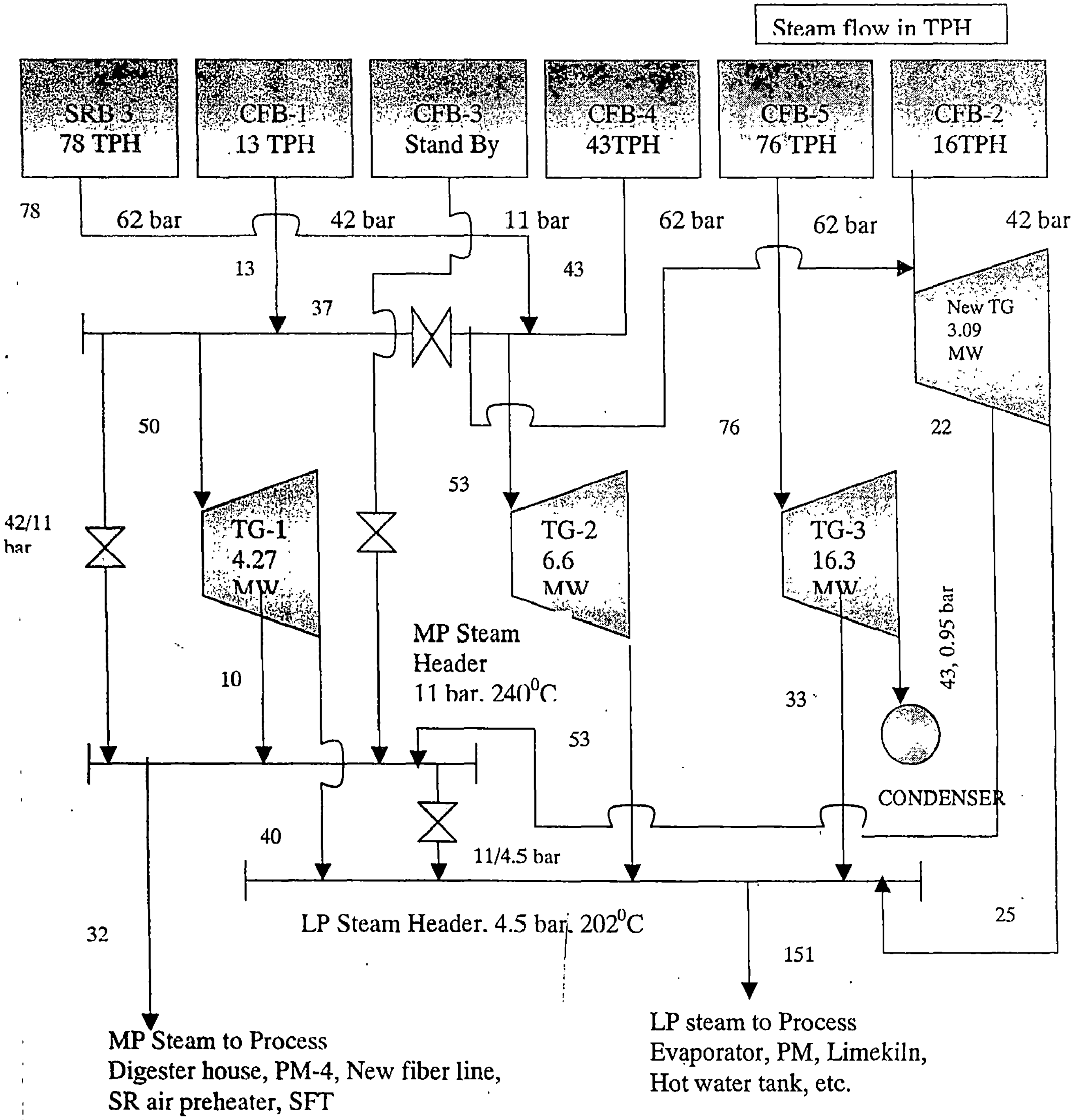


Figure (44): Modified Co-Generation System



4.14. Enhancing energy economy with new technological options.

4.14.1. Displacement batch cooking (RDH)

Background

Increased pressure to minimize environmental impact and effluent volumes, contributed to the development of batch cooking using modified cooking chemistry principles and extended cooking. Later in 1980s, the development of the next generation of liquor displacement batch cooking systems continued with the aim of providing greater optimization of pulp quality. Large, mill scale investigations of the influence of cooking chemistry and cooking equipment f pulp quality occurred. This development phase of liquor displacement Kraft batch cooking resulted in the Super Batch concept. This involved both energy efficiency and efficient use of residual and fresh cooking chemicals. The technique combines energy advantages with modified cooking chemistry.

How it so Energy Efficient?

The process calculation is carried with present system of description on displacement cooking (Detail calculation given in **Appendix-VI/1**). The following **Table (36)** shows the improvement in performance of pulping process by applying RDH process.

Table (36): Improvement in performance of pulping process by applying RDH process.

Particulars	Performance values of present process	New performance values by retrofitting the RDH process	Remarks
LP steam consumption in cooking, kg/cook	-	461	For presteaming of chips before cooking
MP steam consumption in cooking, kg/cook ❑ Direct heating ❑ Indirect heating	 - 11789.2	 3659 3037.05	Extraction of MP steam from cogeneration plant reduces by 43%
Bath Ratio, (liquor to Wood)	2.8	4.2	High bath ratio higher in uniformity of cooking
Hot water generation @ 80^0C, m^3/cook	142 (with LP steam of 3.5 MT/cook)	88	For attaining the present requirement of hot water, requires 4.8 MT of LP steam additional.
Condensate recovery, m^3/cook	10.610	2.551	Rate of condensate recovery is 38% due to system with direct and indirect heating.
Comparative made on present 80 m^3 digester productions per cook.			

Possible savings by improvement,

Annual savings = Rs. 26 lakh on grid power.

Pay back period = Long Term.

By above discussion on the RDH process, its performance improvement on energy consumption, compared to conventional batch process is observed. The other possible improvement measures had to be studied by mill on laboratory scale to know,

- Increase in yield of pulping.
- Improvement in alkali profile.
- Improvement in strength of pulp delivery.
- Reduction in chemical consumption in bleaching plant.

The process is not even energy efficient; it increases the productivity of the mill.

4.14.2. Thermal depolymerization of black liquor.

Background

Present black liquor from the washing section is at 14-18% dry solids average, which has to be concentrated above 65%, and eventually it is fired in the recovery furnace for recovering the inorganic as smelt and organics as steam. As the concentration of black liquor is increased, there is a steep rise in the viscosity of black liquor, which actually put a limitation on the further concentration of black liquor to a high solid concentration.

Attempt to reduce the viscosity of black liquor is either by increasing the temperature of black liquor or by controlling the RAA value of black liquor. The way of controlling RAA level for viscosity reduction presently exists in many mills, which indirectly increases the inorganic content on black liquor solids, results in reduction of calorific value of black liquor solids. Where as attempt of increasing the temperature of black liquor reduces the viscosity without any desire effect.

Reducing the viscosity of black liquor by heating it to the temperature above those normally encountered during pulping. Firing of black liquor at high dry solids concentration with proper combustion is one of the prerequisite for generation of high pressure steam and efficient cogeneration. Black liquor losses viscosity permanently when it is heat treated at 170-190^{0}C for approximately 30 minutes due to the degradation of the long chain lignin and carbonates. Therefore, even at high dry solids content, the black liquor can be pumped using centrifugal pumps.

For the comparative study with present and system with heat treatment

- Taking the liquor for heat treatment at 38.5% dry solids (from 3rd effect)
- Black liquor treatment temperature of 170^{0}C is assumed.

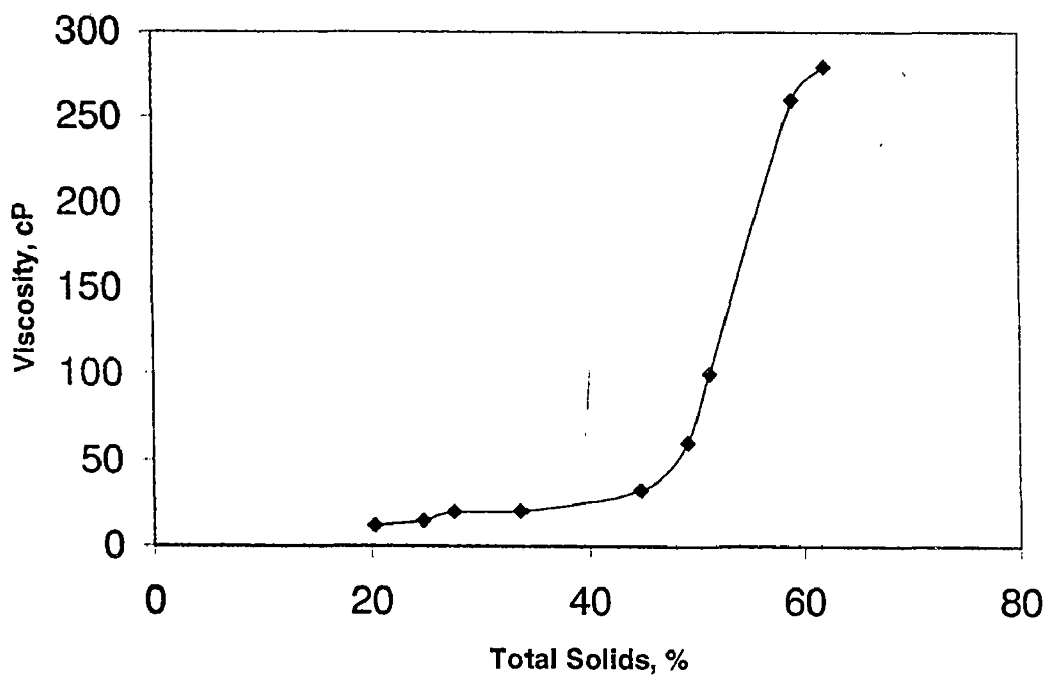
- Fixed final concentration of 78% dry solids.
- Temperature scales, pressure scales, and flow details are taken as per design values of present system.

The viscosity data at various concentration level of the present system of evaporation of black liquor is given in **Table (37).** The rise of viscosity level at different concentration points is shown in **Figure (45)**,

Table (37): Viscosity level at various Concentration points

Solids, %TS	pH	Viscosity, cP
20.31	12.8	12
24.78	12.8	15
27.62	12.9	20
33.75	12.9	21
44.78	12.9	33
49.23	12.8	60
51.19	12.9	100
59.09	13.3	260
62.10	13.0	280

Figure (45): Rise in viscosity level with increasing in total solids



The viscosity of black liquor shows a linear increasing in viscosity up to about 42% solids, the viscosity values shows a sharp rise after this concentration level. The point of inflection, the point where black liquor viscosity shows a sharp rise in concentration for the system appears to be in the range of 40-45% indicating that the attempt to decrease the viscosity at this point will have significant impact on transport and thermal property of black liquor.

The process calculation is carried with present system of description on heat treatment of black liquor evaporation (Detail calculation given in **Appendix-VI/2**). The following **Table (38)** shows the benefit of heat treatment with present conventional evaporation.

Table (38): Improvement in performance of chemical recovery process by applying heat treatment process

Particulars	Performance values for existing process	New performance by retrofitting the heat treatment process	Remarks
Steam economy	6.16	6.22	Increase in steam economy by 0.96%
Water evaporation, tph	125	130	
MP steam consumption for evaporation, tph	-	2.35	Thermal treatment to temperature of 170^{0}C is carried out by additional MP steam consumption.
LP steam consumption for evaporation, tph	20.3	18.55	Recovered flash vapor from MP condensate, utilizing back to system as LP steam.
Concentration of black liquor to HBL tank, % TS	70	78	Reduction of viscosity of black liquor improves concentration of black liquor with present system by 10.25%.
Steam generation in recovery boiler, tph	90	93.82	Steam generation in boiler increases by 4%.

Possible savings by improvement,

Annual Savings = Rs. 91 lakh on steam generation in SRB

Pay back period = long term

CHAPTER 5
CONCLUSION

5.1. Outcome

The detailed energy conservation approaches given in the report has been tabulated in **Table (39)**. The table shows the annual savings as well as investment required. The table also provides the payback period of the schemes. The pay back period accounts the effects on the investment of taxes and depreciation. The depreciation period is assumed to be 10 years and the tax rate is 15%.

Table (39): List of Energy Saving Opportunities

Sl. No	Recommendations	Energy Savings				Cost of Implementation Rs. lakh	Pay Back Period, months
		Steam TPH	Power, MW	Furnace Oil, kl/hr	Value, Rs. lakh		
1.	Two stage batch digester steaming	-	0.12	-	30	50	23
2.	Improvement of blow heat recovery system of pulp mill	3.84	-	-	91.24	15	2
3.	Proper scheduling of digesters	20% increase in production					
4.	Improvement of steam economy in MEE system	1.51	-	-	35.8	Margin	Nil
5.	Controlling the crystallization of salts in black liquor and increasing the system availability	1.37	-	-	112	Nil	Nil
6.	Indirect heating of heavy black liquor up to firing temperature	1.19	-	-	29.2	12	6
7.	Effect of addition of neutralized sesqui sulphate crystal (i.e. Spent acid from ClO_2 plant) in black liquor as sodium sulphate and increasing in steam generation	0.56	-	-	13.3	10	10
8.	Optimization of cooling water requirement in surface condenser.						
	a. Savings with exact requirement pump capacity	-	0.076	-	19.53	2.5	1.5
	b. Saving with existing pump capacity*	-	0.01	-	2.57	Nil	Nil
	c. Saving by avoiding orifice meter	-	0.032	-	8.31	Nil	Nil
	* Not considered for grand total						
9.	Smelt dissolving tank vapor recovery for process condensate water heating	1.59	-	-	37.87	20	7

SL. NO	Proposal	Energy Savings				Cost of Implementation Rs. Lakh	Pay Back Period, months
		Steam TPH	Power, MW	Furnace Oil, kl/hr	Value, Rs. lakh		
10.	Boiler blow down and Air preheater condensate tank flash vapor recovery	1.6	-	-	38.03	12.5	5
11.	Utilization of condensate available from lime kiln oil heater, SRP oil heater & soot blowing in SRB	0.41	-	-	10.58	Marginal	Nil
12.	Reduction of furnace oil consumption in lime kiln section	8	-	0.02	213	2	0.13
13.	Improvement in power generation by modification of cogeneration system	-	3.09	-	764.14	600	11
14.	Enhancing energy economy with new technological options.						
	a. RDH	-	0.102	-	26	NA	Long Term
	b. BL heat treatment	3.82	-	-	91	NA	Long Term
15.	**Grand total**	**23.89**	**3.43**	**0.02**	**1520.00**	-	-

5.2. Economic Feasibility and Discussion

Economic analysis of proposed case studies was done through pay back method by accounting the depreciation and tax on investment and all the Energy Saving Opportunities are economically valid since the payback period is less than two years. The case study came out with annual energy saving potential of Rs.1520 lakh, which works to be 40% of the annual energy cost (Annual energy cost is taken as Rs. 3830 lakh based on energy consumption during financial year 2003-2004). From the above outcome of proposed case studies some have started implementing.

Energy in the industry is still seen as an overhead cost and not as an important part of production as it should be. The industry should review the studies made and put forward to implement them. There should be regular reviews of the energy scene of the plant and the whole industry should function as a unit to optimize energy.

Awareness among the workers regarding energy conservation aspect should be increased. Various incentives in energy conservation have to be initiated.

REFERENCES

1. Mikko Tuhkanen, "Two stage batch digester steaming reduces mill energy consumption", Pulping process, Tappi press, 1990.
2. Brahmanand Mohanty, "Technology energy efficiency and environmental externalities in the pulp and paper industry", 1st edition, School of energy efficiency in Asia, Thailand, 1997.
3. Pradeep Chaturvedi, "Sustainable energy supply in Asia", Vol. 1 & 2, McGraw Hill International Ltd., 1997.
4. "Training manual of Displacement Pulping", Sunds Defibrator Pori Oy, Finland.
5. "Superbatch cooking: From innovation to experience", Paper Asia, March, 1997.
6. John C.W. Evans, "RDH process boosts batch digester efficiency at US and Finnish mills", Pulp and Paper, May, 1987.
7. Steven L. Erickson, "Operating experience with indirect heating of strong black liquor", Proceedings, Tappi press, 1978.
8. Rao, N J, "Design data and correlations of waste liquor / black liquor from pulp mills", IPPTA, Vol.4, September, 1992.
9. Wolfgang Schmidi, "Evaporator Fouling: How IPST's evaporator performance audits can help mills", IPST.
10. Wayne Adams, "A crystal growth system effectively reduces evaporator scaling, while providing the flexibility for future liquor chemistry changes", Kellogg brown and Root engineering, Houston, Texas.
11. Thakur Associates, A study on "Blow heat recovery system", ITC PSPD unit: Bhadrachalam, India.
12. Terry N. Adams, "Sodium salt scaling in black liquor evaporators and concentrators", Tappi Journal, June, 2001.
13. Mark A. Rosier, "Model to predict the precipitation of burkeite in the multiple effect evaporator and techniques for controlling scaling", Tappi Journal, April, 1997.
14. John Rauscher, "Liquor heat treatment at Pope & Talbot Inc.", Alhstrom Machinery Corporation.
15. Max S. Peters, "Plant design and economics for chemical engineering", McGraw Hill Book Company, New York, 1958.

16. Jain R K, "Improved energy efficiency through thermal treatment of black liquor", IPPTA, Vol.12, June, 2000.

17. A report on "Energy Audit of ITC Ltd., PSPD, Unit Bhadrachalam", TERI Project Report No. 2003IS01.

18. A case study report on "Flux technology for furnace oil combustion", Flux-Hi-Tech India Pvt Ltd., Mumbai, India.

19. A "Mill performance report", ITC PSPD, Unit Bhadrachalam, February, 2004.

20. **Web sites.**

20.1. www. Bee-india.com

20.2. www. Surf.com

20.3. www. Greenbusiness.org

20.4. www.tappi.org

21. **Referred Suppliers for investment details,**

21.1. Enmas Ahlstrom Pvt. Ltd., Chennai, India.

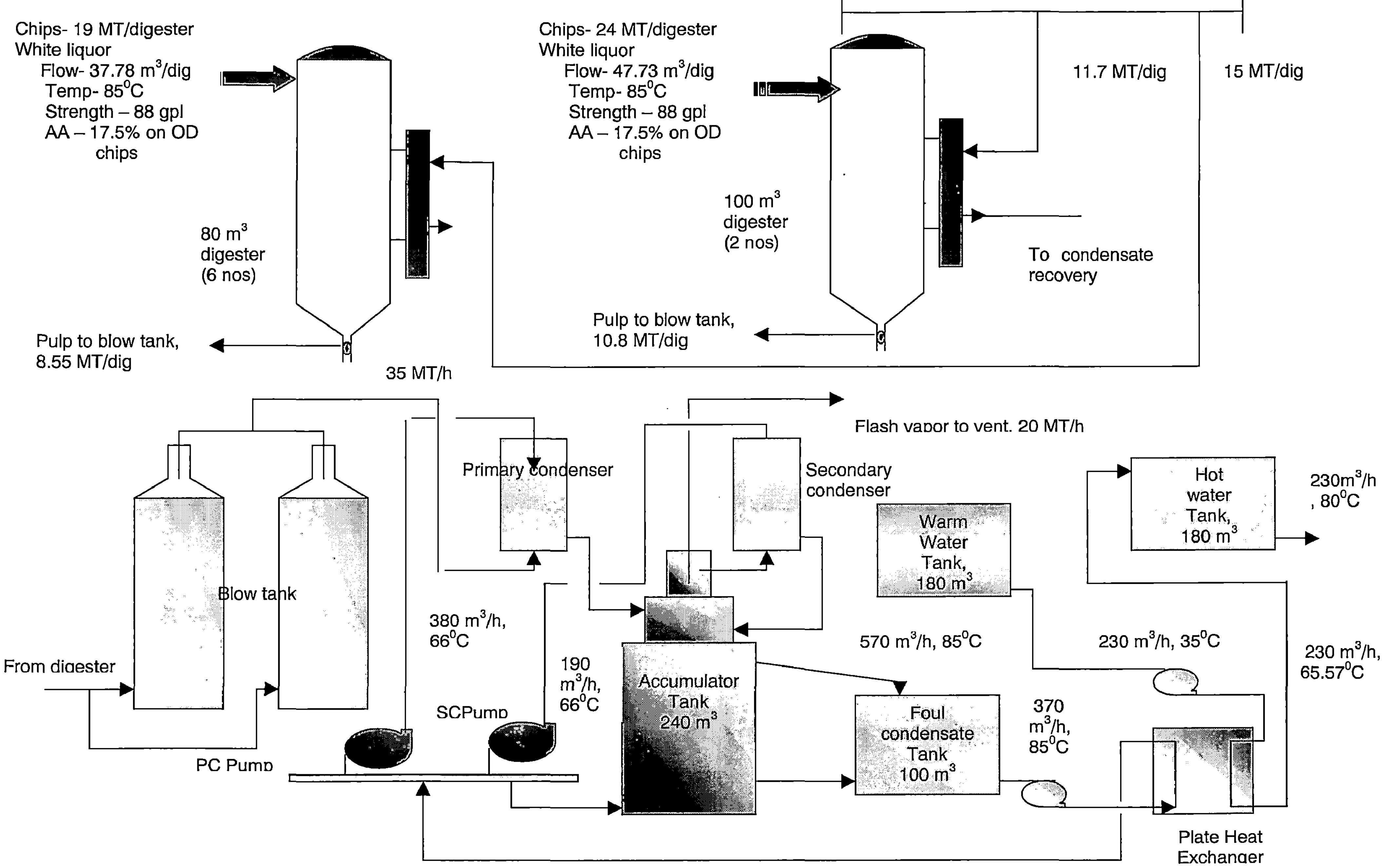
21.2. Forbes Marshall, Pune, India.

21.3. Alfa-Laval India Ltd., Pune, India.

APPENDICES

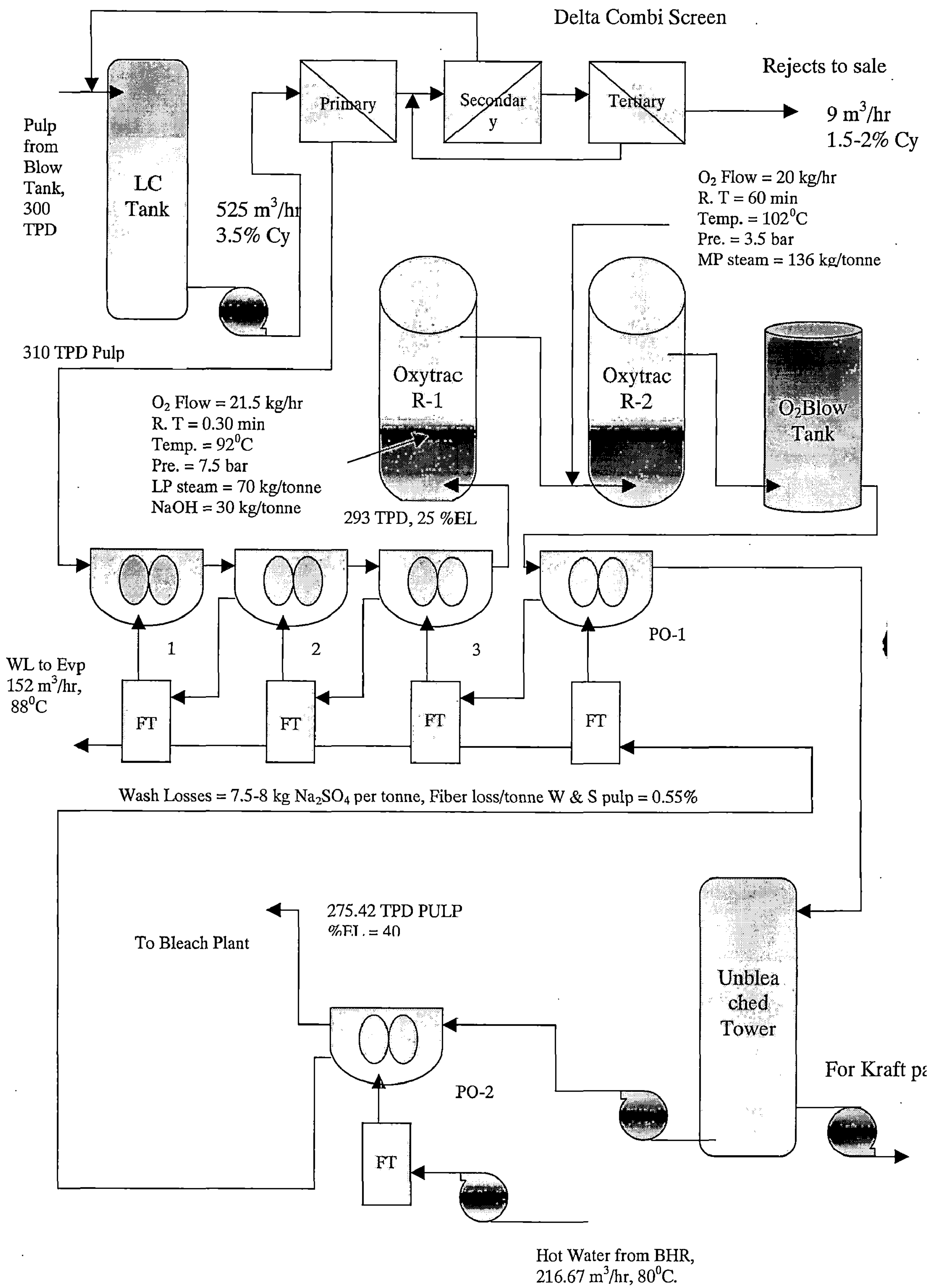
Appendix- I/1

Proce[illegible] of Digester House



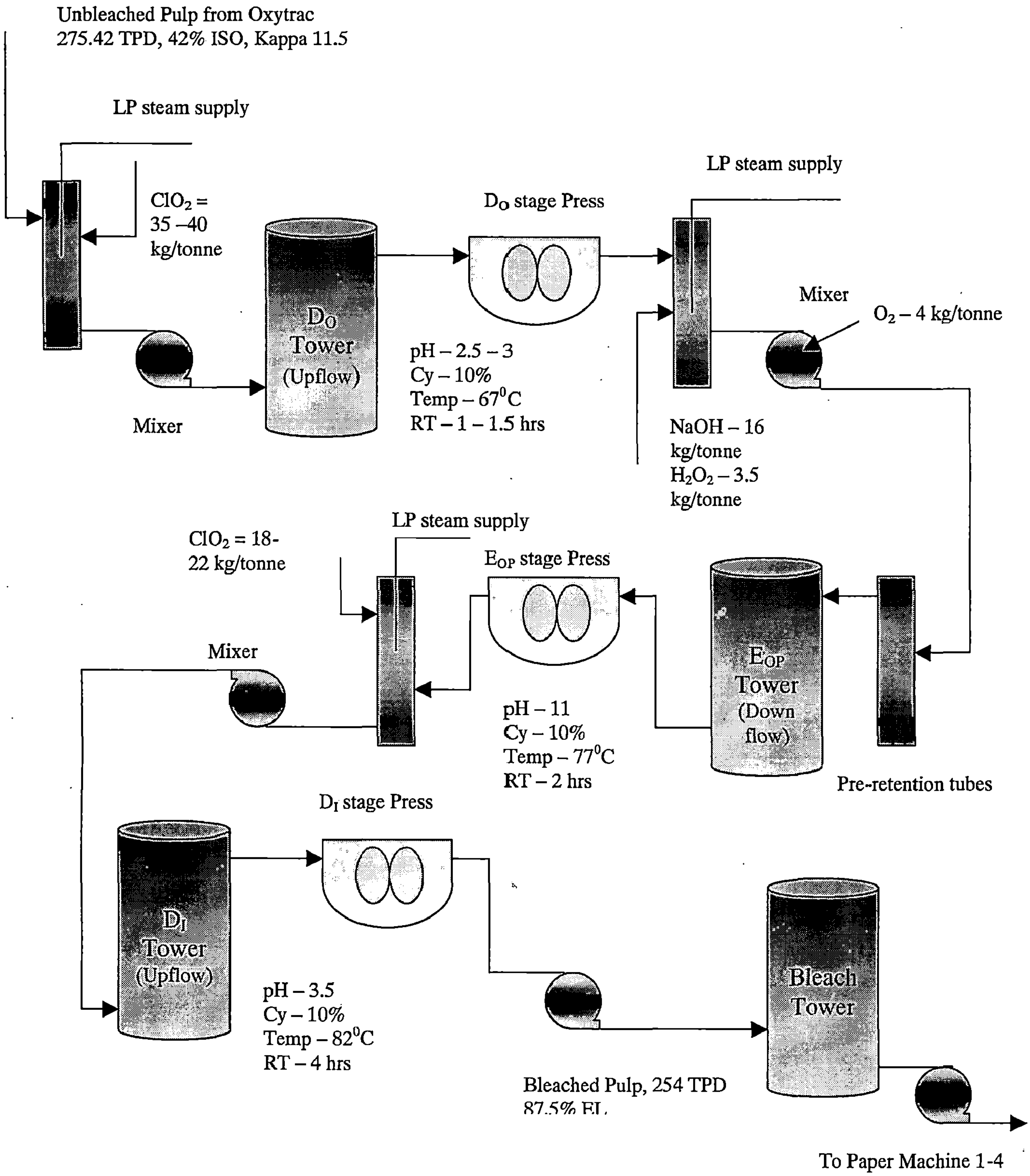
ɹɒpendix- I / 2

New Fiber Line (System of Screening, Washing, & Oxygen Deliginification)



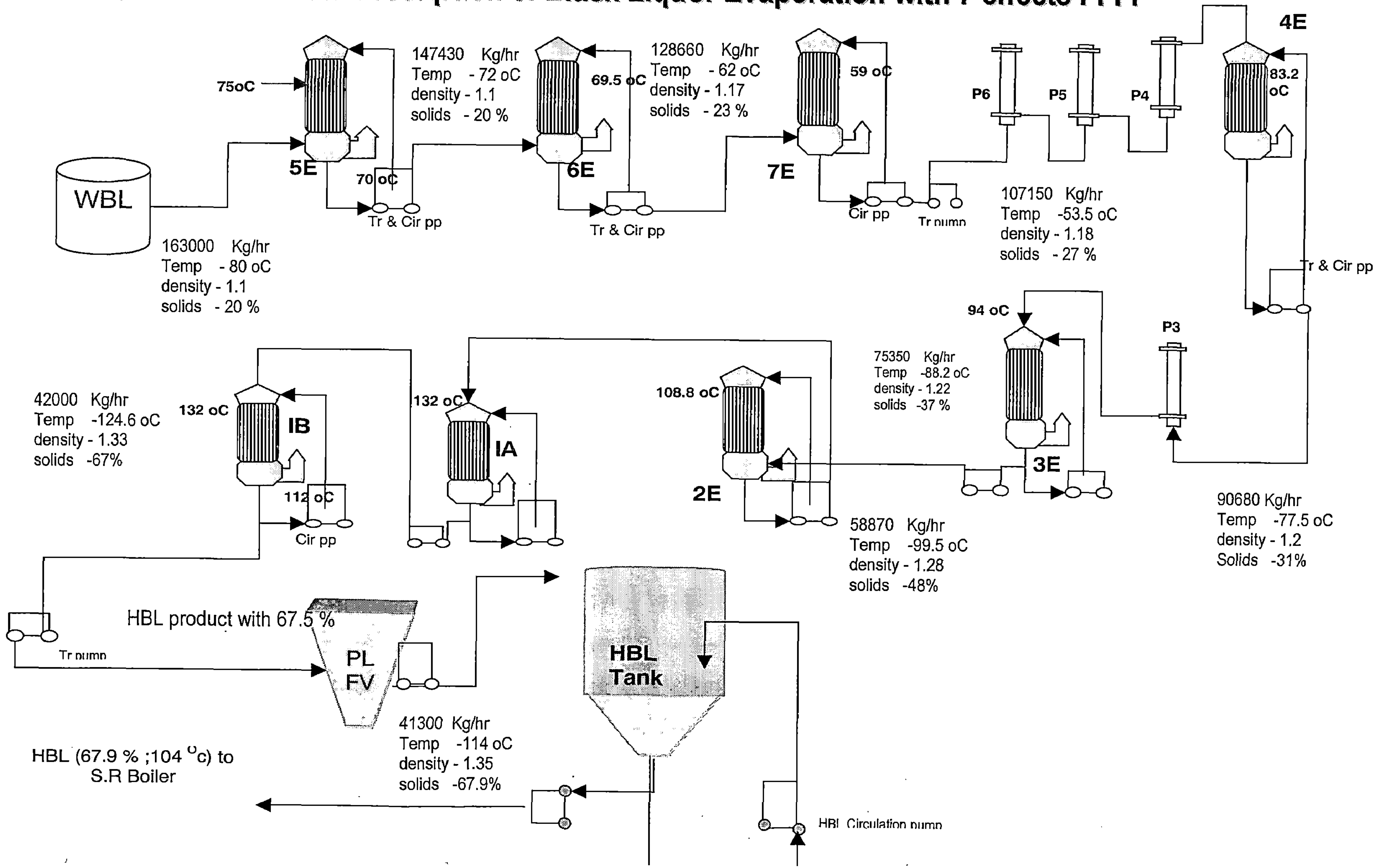
Appendix- I/3

Bleach Plant of ECF Sequence $D_OE_{OP}D_I$



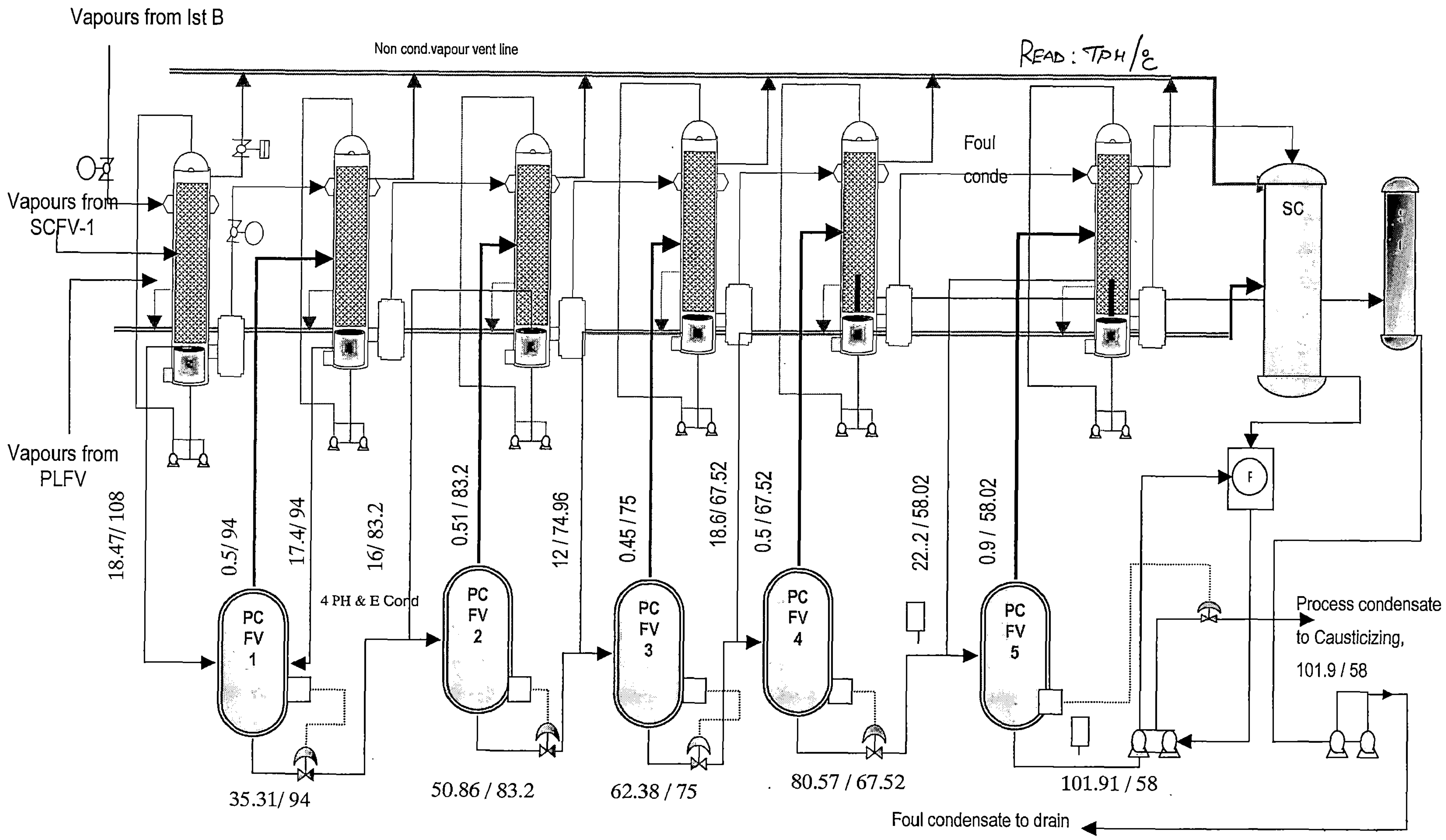
Appendix – I/4

Process Flow Description of Black Liquor Evaporation with 7 effects FFFF



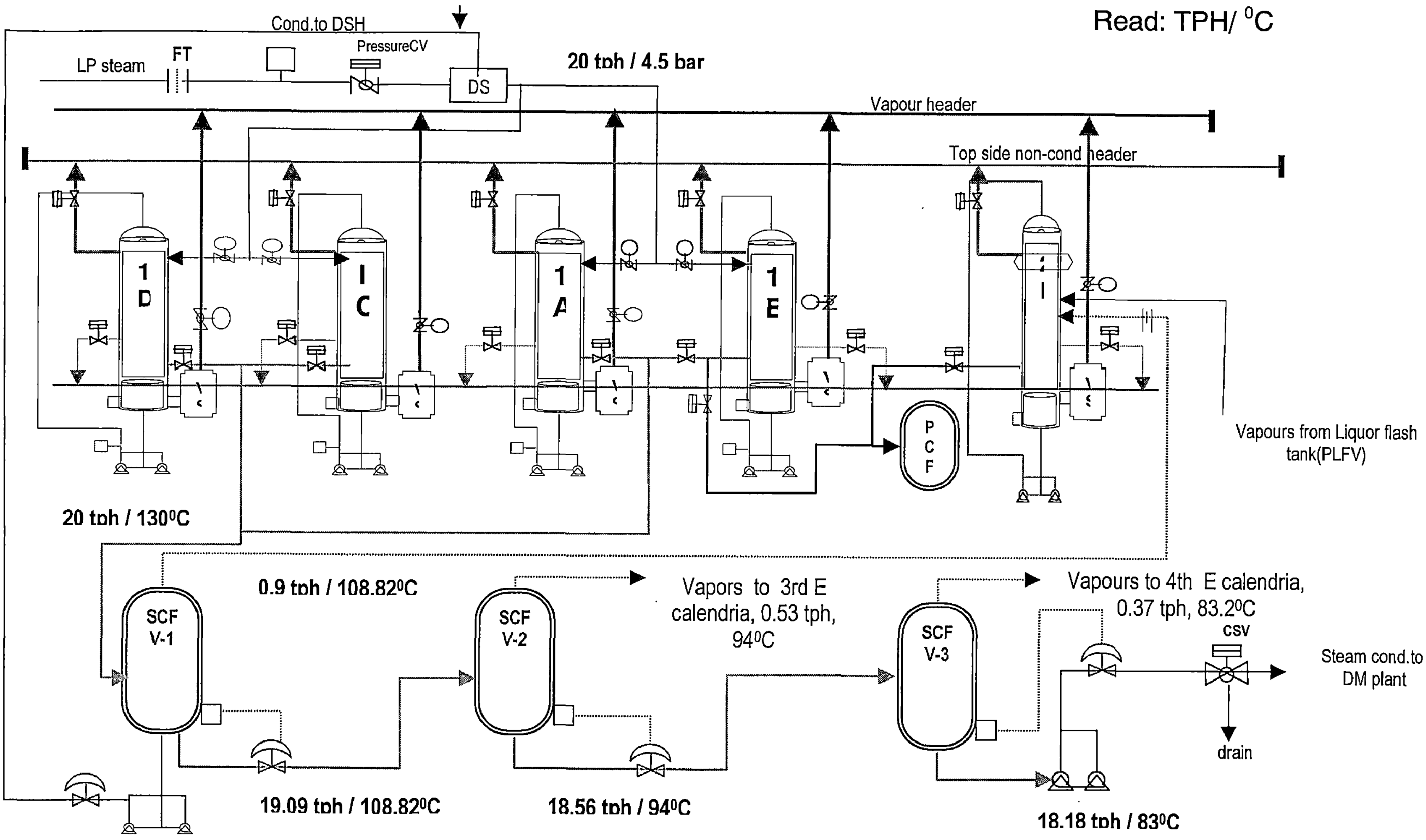
Appendix-I/4.1

Process (Foul) Condensate Flashing System in FFFF



Appendix- I/4.2

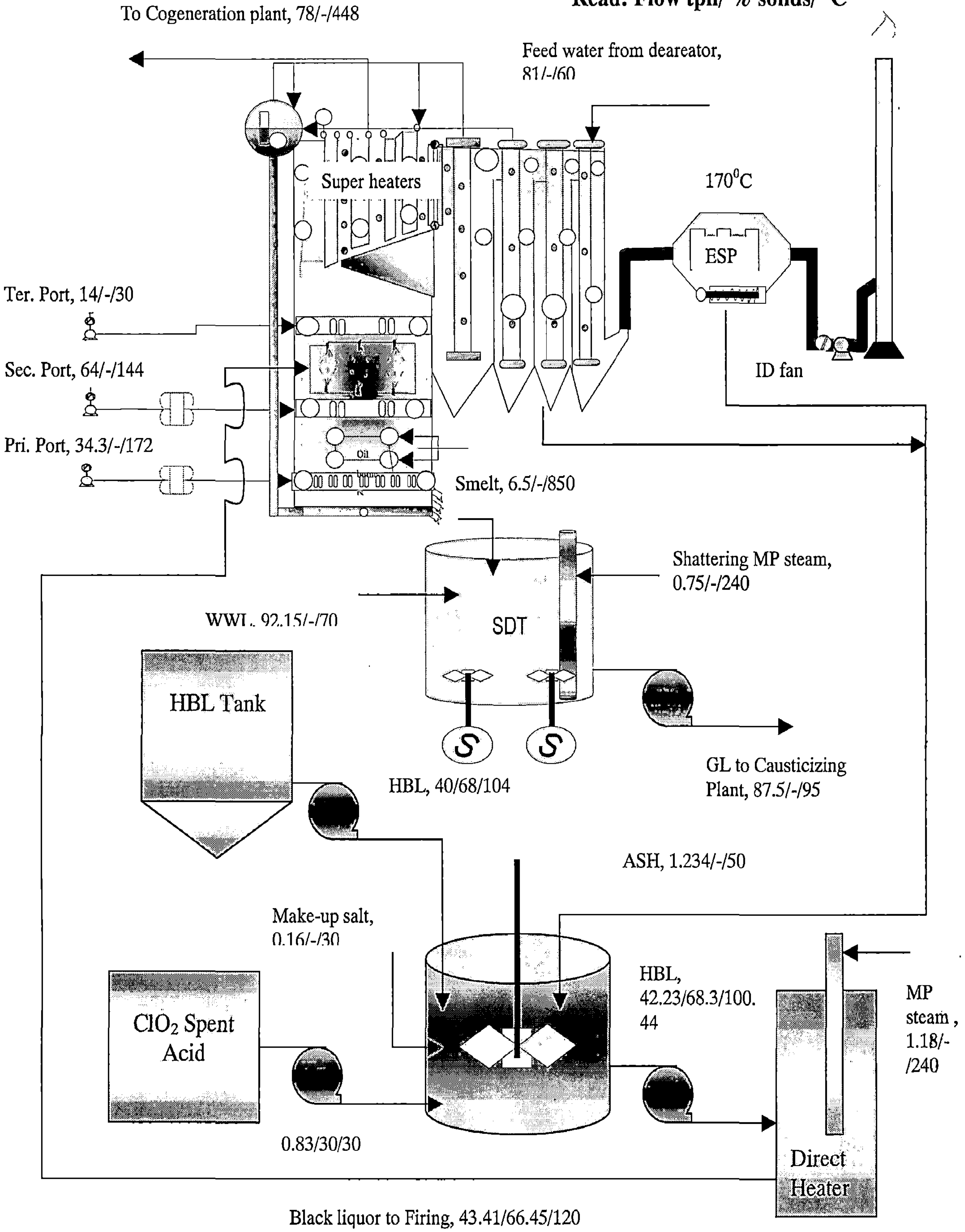
Steam Condensate Flashing System in FFFF



Appendix- I/5

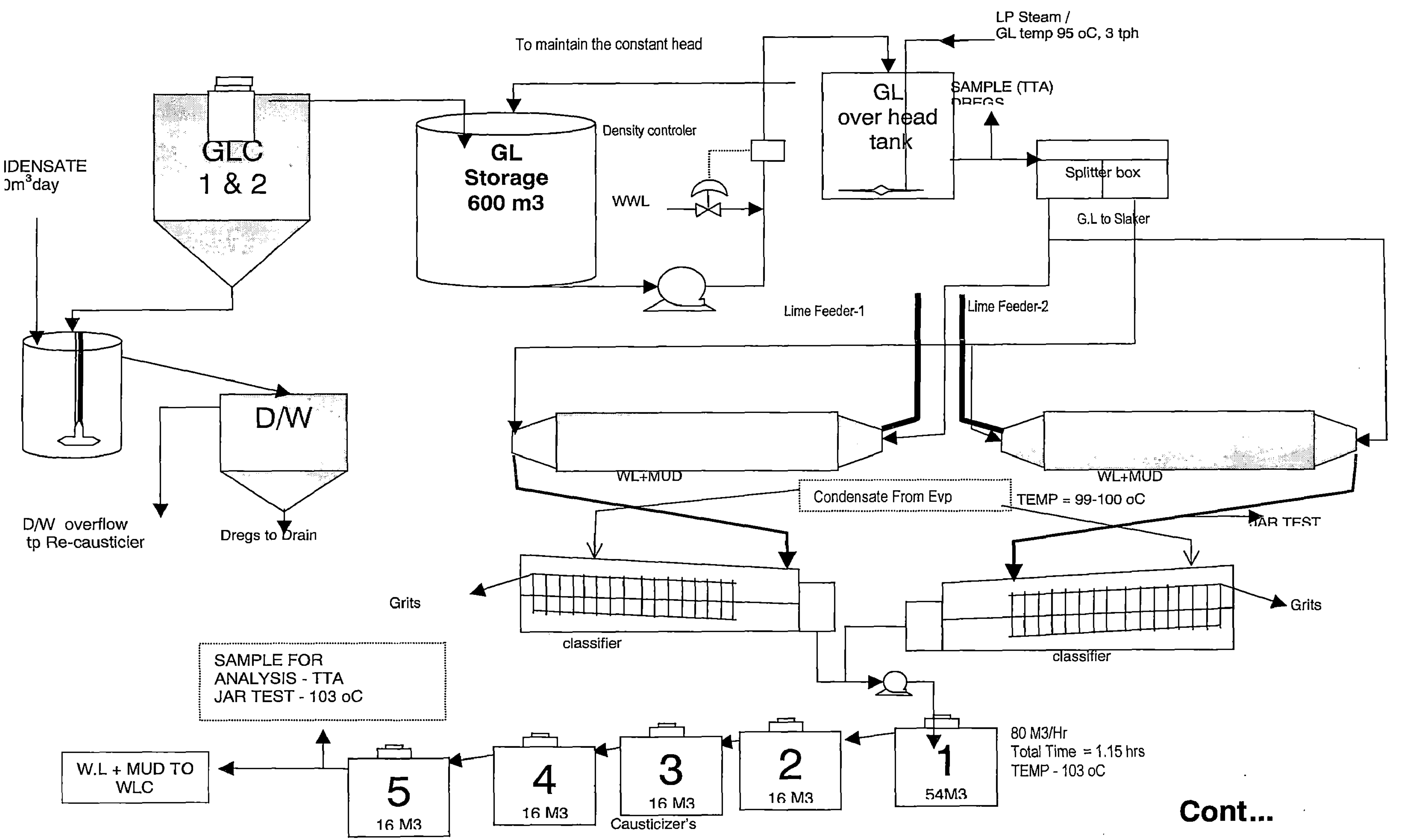
Process Flow Diagram of Recovery Furnace

Read: Flow tph/ % solids/ ^0C



Appendix- I/6

Process Flow Description of Recausticizing Section



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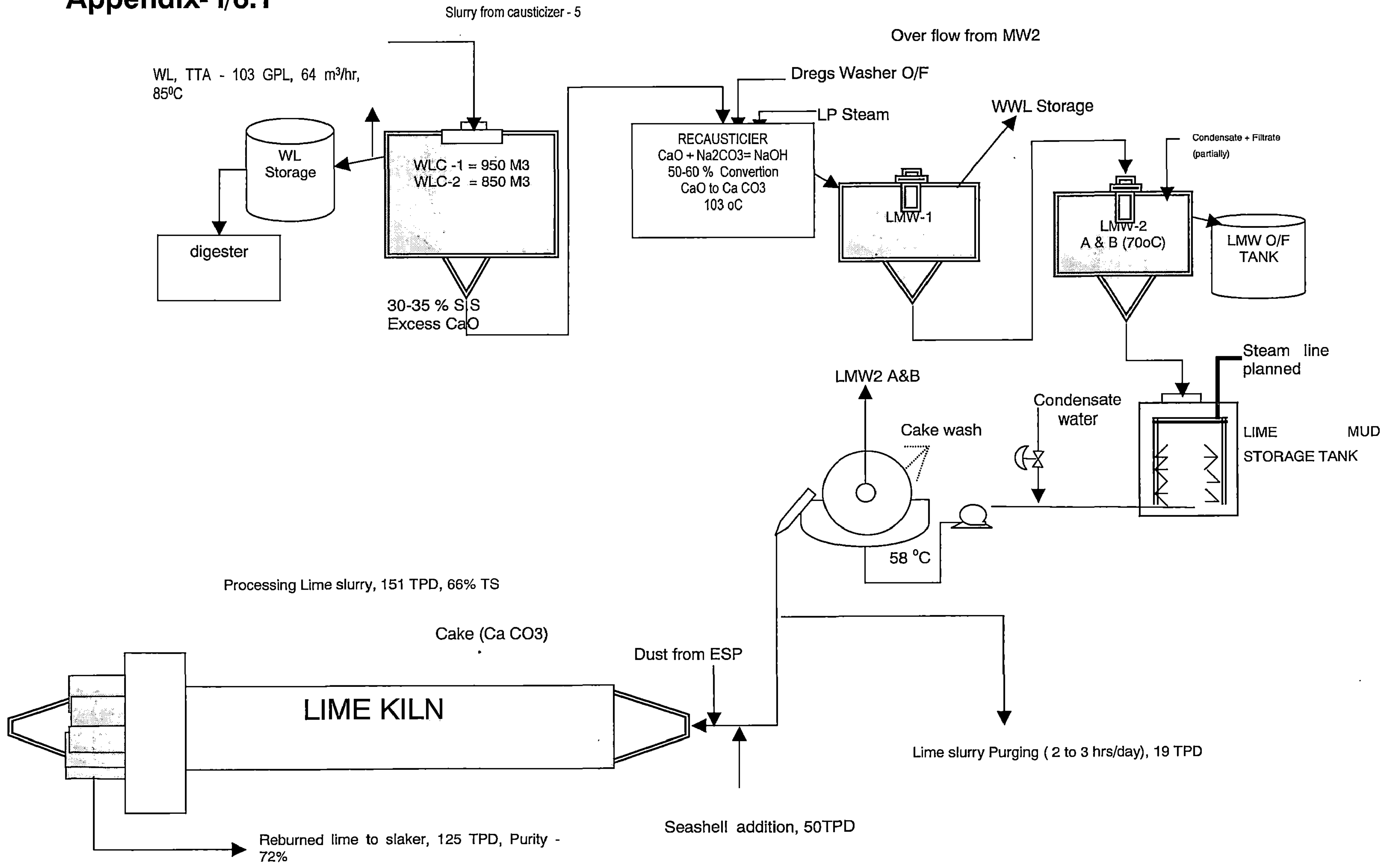
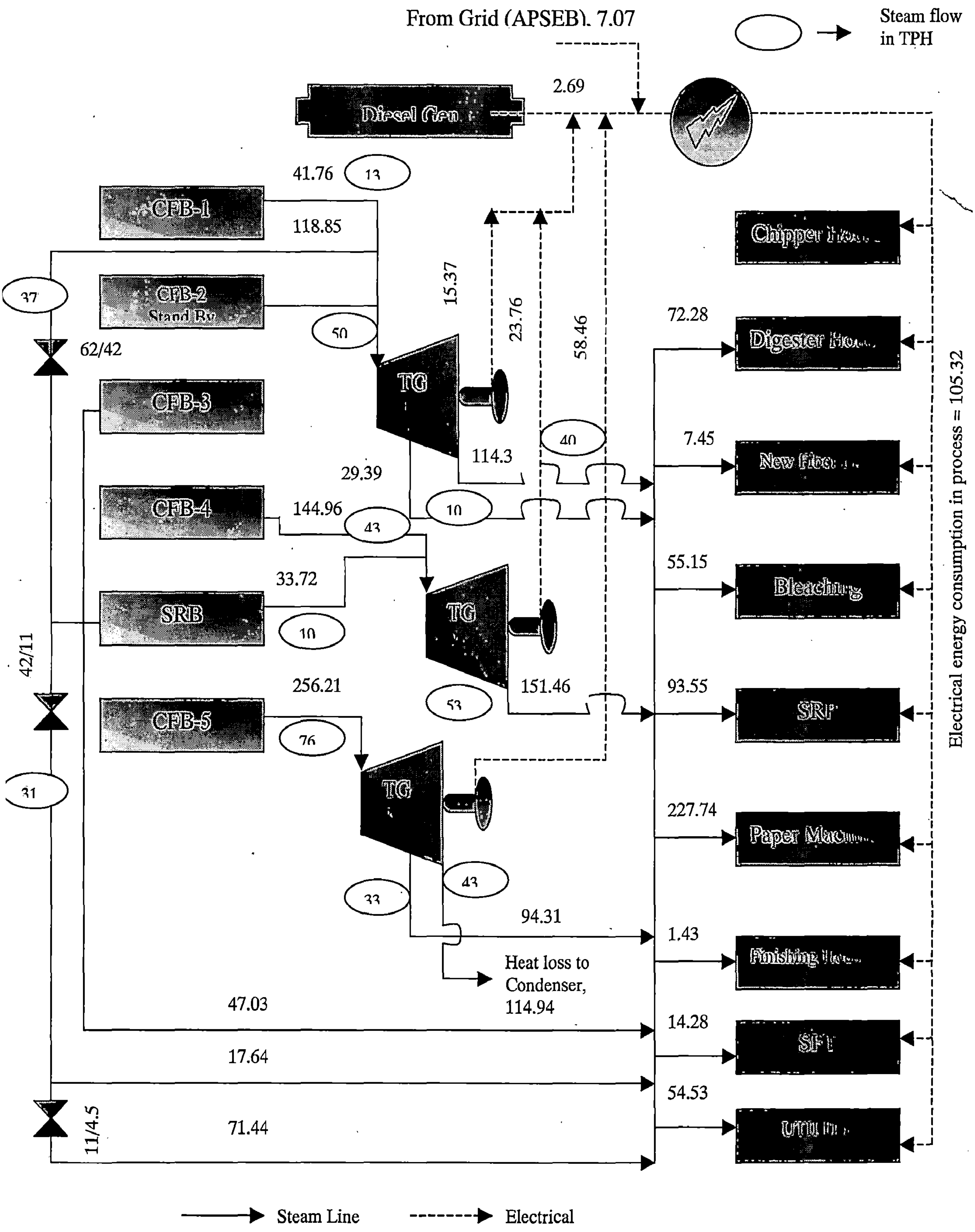
Appendix- I/6.1



Appendix- I / 7

System Energy Balance of the Mill Cogeneration Unit (Basis GJ/hr)



Appendix-II/1

Steam Requirement Calculation in Digester House

Process description:

S.no	Particulars	UOM	Values
1	Chip feed on BD basis		
	80 m3 digester	MT/digester	19
	100 m3 digester	MT/digester	24
2	Moisture in chips	% on BD chip	30
3	White liquor requirement	% on BD chip	17.5
4	Strength of white liquor	gpl as Na2O	88
5	White liquor feed temperature	oC	85
6	Yield of pulping	%	45
7	Cooking temperature	deg. C	162
8	Chip feed temperature	deg. C	30
9	Initial temperature of digester shell	deg. C	75
10	Black liquor feed temperature	deg. C	85
11	Black liquor feed concentration	% TS	18
12	Sp. Heat of white liquor	kJ/kg deg.C	3.8
13	Sp . Heat of black liquor	kJ/kg deg.C	5
14	Sp. Heat of chips	kJ/kg deg.C	1.38
14	Sp . Heat of digester shell	kJ/kg deg.C	0.627
15	Sp . Gravity of white liquor	-	1.14
16	Sp . Gravity of black liquor	-	1.08
17	Bath ratio	-	2.8
18	Number of digester		
	80 m3 digester	-	6
	100 m3 digester	-	2
19	Shell weight of digester		
	80 m3 digester	MT	35
	100 m3 digester	MT	44
20	Sp. Gravity of shell	-	7.86
21	Number of blows per day		
	80 m3 digester	-	30
	100 m3 digester	-	9

Process calculation:

Basis : Per cook.

For 80 m3 digester

1	White liquor charge	MT as Na2O	3.33
2	Volume of white liquor charge	m3	37.78
3	Mass of white liquor charge	MT as such	43.07
4	Total liquor in digester with 2.8 bath ratio	MT	53.2
5	Amount of black liquor charge	MT	4.43
6	Pulp produced (BD basis)	MT	8.55

For 100 m3 digester

1	White liquor charge	MT as Na2O	4.20
2	Volume of white liquor charge	m3	47.73
3	Mass of white liquor charge	MT as such	54.41
4	Total liquor in digester with 2.8 bath ratio	MT	67.2
5	Amount of black liquor charge	MT	5.59
6	Pulp produced (BD basis)	MT	10.8

Heat balance:

For 80 m3 digester per cook

s.no	Description	Mass, kg	Cp,kJ/kg · deg. C	Delta T, deg.C	Heat value, kJ
1	Heating of chips	19000	1.38	132	3461040
2	Heating of moisture in chips	5700	4.18	132	3145032
3	Heating of white liquor	43073.864	3.8	77	12603412.5
4	Heating of black liquor	4426.1364	5	77	1704062.5
5	Heating of shell	35000	0.627	87	1909215
6	Radiation losses @ 2.5%				570569.05

Enthalpy of steam @ 11 bar	kJ/kg	1984.3
MP steam consumption	MT	11.79

For 100 m3 digester per cook

s.no	Description	Mass, kg	Cp,kJ/kg deg. C	Delta T, deg.C	Heat value, kJ
1	Heating of chips	24000	1.38	132	4371840
2	Heating of moisture in chips	7200	4.18	132	3972672
3	Heating of white liquor	54409	3.8	77	15920100
4	Heating of black liquor	5591	5	77	2152500
5	Heating of shell	44000	0.627	87	2400156
6	Radiation losses @ 2.5%				720431.7

Enthalpy of steam @ 11 bar	kJ/kg	1984.3
MP steam consumption	MT	14.89

Total steam consumption for cooking in digester per day MT/day 488

Appendix-II/2

Two stage of steaming with LP and MP steam

Process description is similar to the existing one. Only change is indirect steaming partially carried with LP steam and with MP steam. Temperature rise from 80 deg. C to 112 deg.C is rised by using the LP steam and temperature from 112 to 162 deg.C rise is by MP steam.

For 80 m3 digester

Temp rise from 80 - 112 deg.C (32 deg.C) for 1hr

s.no	Description	Mass, kg	Cp, kJ/kg deg.C	Delta T, deg.C	Heat value, kJ
1	Heating of chips	19000	1.38	82	2150040
2	Heating of moisture in chips	5700	4.18	82	1953732
3	Heating of white liquor	43073.864	3.8	27	4419378.41
4	Heating of black liquor	4426.1364	5	27	597528.41
5	Heating of shell	35000	0.627	37	811965
6	Radiation losses @ 2.5%				248316.10

Enthalpy of steam @ 4 bar	kJ/kg	2107
LP steam consumption	MT	4.83

Temp rise from 112-122 deg.C (10 deg.C) for 1/2 hr

s.no	Description	Mass, kg	Cp, kJ/kg deg.C	Delta T, deg.C	Heat value, kJ
1	Heating of chips	19000	1.38	10	262200
2	Heating of moisture in chips	5700	4.18	10	238260
3	Heating of white liquor	43073.864	3.8	10	1636806.82
4	Heating of black liquor	4426.1364	5	10	221306.82
5	Heating of shell	35000	0.627	10	219450
6	Radiation losses @ 2.5%				64450.59

Enthalpy of steam @ 11 bar	kJ/kg	1984.3
MP steam consumption	MT	1.33

Temp rise from 122 to 162 deg.C (40 deg.C) for 45 min

s.no	Description	Mass, kg	Cp, kJ/kg deg.C	Delta T, deg.C	Heat value, kJ
1	Heating of chips	19000	1.38	40	1048800
2	Heating of moisture in chips	5700	4.18	40	953040
3	Heating of white liquor	43073.864	3.8	40	6547227.27
4	Heating of black liquor	4426.1364	5	40	885227.27
5	Heating of shell	35000	0.627	40	877800
6	Radiation losses @ 2.5%				257802.36

Enthalpy of steam @ 11 bar	kJ/kg	1984.3
MP steam consumption	MT	5.33
Total steam consumption per dig.	MT	11

For 100 m3 digester

Temp rise from 80 - 112 deg.C (32 deg.C) for 1hr

s.no	Description	Mass, kg	Cp, kJ/kg deg.C	Delta T, deg.C	Heat value, kJ
1	Heating of chips	24000	1.38	82	2715840
2	Heating of moisture in chips	7200	4.18	82	2467872
3	Heating of white liquor	54409.091	3.8	27	5582372.73
4	Heating of black liquor	5590.9091	1.08	27	163030.91
5	Heating of shell	44000	0.627	37	1020756
6	Radiation losses @ 2.5%				298746.79

Enthalpy of steam @ 4 bar	kJ/kg	2107
LP steam consumption	MT	6

Temp rise from 112-122 deg.C (10 deg.C) for 1/2 hr

s.no	Description	Mass, kg	Cp, kJ/kg deg.C	Delta T, deg.C	Heat value, kJ
1	Heating of chips	24000	1.38	10	331200
2	Heating of moisture in chips	7200	4.18	10	300960
3	Heating of white liquor	54409.091	3.8	10	2067545.45
4	Heating of black liquor	5590.9091	5	10	279545.45
5	Heating of shell	44000	0.627	10	275880
6	Radiation losses @ 2.5%				81378.27

Enthalpy of steam @ 11 bar	kJ/kg	1984.3
MP steam consumption	MT	2

Temp rise from 122 to 162 deg.C (40 deg.C) for 45 min

s.no	Description	Mass, kg	Cp, kJ/kg deg.C	Delta T, deg.C	Heat value, kJ
1	Heating of chips	24000	1.38	40	1324800
2	Heating of moisture in chips	7200	4.18	40	1203840
3	Heating of white liquor	54409.091	3.8	40	8270181.82
4	Heating of black liquor	5590.9091	5	40	1118181.82
5	Heating of shell	44000	0.627	40	1103520
6	Radiation losses @ 2.5%				325513.09

Enthalpy of steam @ 11 bar	kJ/kg	1984.3
MP steam consumption	MT	6.73
Total steam consumption per dig.	MT	14

Total Steam Consumption:

1 Total LP steam consumption per day	MT	197.28
2 Total MP steam consumption per day	MT	275.42
3 Total steam consumption per day	**MT**	**472.70**

Possible flash vapor recoverable from MP steam condensate:

Flashing to 4.5 bar pressure

1	Heat of MP steam condensate	kJ/kg	789.90
2	Heat of MP steam condensate from condensac tank	kJ/kg	652.80
3	Latent heat of LP steam to which condensate is flashed	kJ/kg	2098.10
4	% of flash vapor available @ 4.5 bar	%	6.53
5	LP steam recovered from MP condensate flash	MT/day	18.00

Flashing to atm. Pressure

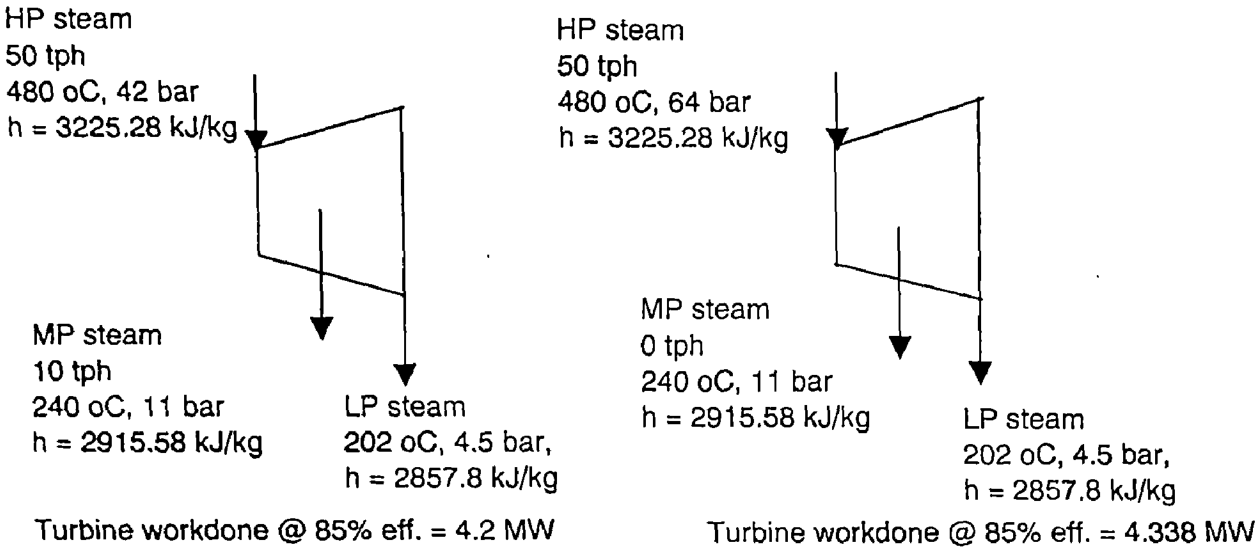
1	Heat of LP steam condensate from condensate tank	kJ/kg	504.70
2	Latent heat of steam at atm. Pressure	kJ/kg	2201.60
3	% of flash vapor available @ atm. Pressure	%	6.73
4	Steam recovered for heating warm water at atm. Pressure	MT/day	30.59

Benefits:

1	Total MP steam saving by two stage steaming	MT/day	212
2	Total Increase in LP steam consumption	MT/day	179

This saving of MP steam will results in additional power generation in TG plant. By reducing the extraction rate of MP steam from TG palnt.

Turbo-Generator-1



Before two stage steaming **After two stage steaming**

Savings by co-generation:

1	Power generation before two stage steaming @ 90% alternator and 95% transmission efficiency (kW)	3591
2	Power generation after two stage steaming @ 90% alternator and 95% transmission efficiency (kW)	3708.99
3	Saving in power generation, kW	117.99
4	Annual savings in power generation @ Rs.3.25 per kwh, Rs. Lakh	30
5	Total investment required, Rs. Lakh	50
6	Simple pay back period, months	20

Pay back period

1	Depreciation per year, Rs. Lakh per annum	5
2	Enterprise tax, Rs. Lakh per annum	3.81
3	Increased cash flow, Rs. Lakh per annum	27
4	Pay back period, months	23

Appendix-II/3

Blow heat recovery system of the Batch digester

Process description

s.no	Particulars	UOM	Values
1	Chip feed on BD basis		
	for 80 m3 digester	MT/digester	19
	for 100 m3 digester	MT/digester	24
2	Number of blows in a day		
	for 80 m3 digester	blow/day	30
	for 100 m3 digester	blow/day	9
3	Number of digester		
	for 80 m3 digester	-	6
	for 100 m3 digester	-	2
4	Blow time		
	for 80 m3 digester	min	15
	for 100 m3 digester	min	20
5	Flushing steam for 5min/blow	MT/hr	2
6	Cooking temperature	deg.C	165
7	Yield of pulping	%	45
8	White liquor consumption	% on BD chip	17.5
9	Bath ratio	Ratio	2.9
10	Consistency of blown pulp	%	11
11	Sp. Heat of pulp	kJ/kg deg.C	4
12	Hot water flow to new fiber line	m3/hr	230
13	Warm water temperature available	deg.C	35
14	Foul condensate temperature available to condenser	deg.C	66
15	Foul condensate temperature available to PHE	deg.C	85
16	Primary condensate pump details (running condition)		
	Discharge	m3/hr	380
	Head	mwc	15
	RPM	rpm	
	Motor power	kW	25
17	Secondary condensate pump details(running condition)		
	Discharge	m3/hr	190
	Head	mwc	15
	RPM	rpm	
	Motor power	kW	13
18	Foul condensate pump details (Name plate details)		
	Discharge	m3/hr	450
	Head	mwc	50
	RPM	rpm	1450
	Motor power	kW	90
19	Warm water pump details (Name plate details)		
	Discharge	m3/hr	400
	Head	mwc	25
	RPM	rpm	
	Motor power	kW	38.13
20	Storage capacity of tanks		
	Accumulator tank	m3	240
	Foul condensate tank	m3	100
	Warm water tank	m3	180
	Hot water tank	m3	180

21	Plate heat exchanger details		
	Make		Alfa-laval
	No of plates	-	107
	Heat transfer area	m2	83
	Each plate area	m2	0.79
	overall heat transfer co-efficient	W/m2 deg.C	5800
22	Actual foul water flow to PHE	m3/hr	370
23	Actual warm water flow to PHE	m3/hr	230

Energy calculation

Present system:

1	Quantity of pulp blown per day@11% Cy (for 80 m3 digester)	MT/day	2332
2	Quantity of pulp blown per day@11% Cy (for 100 m3 digester)	MT/day	884
3	Heat content in blown pulp (for 80 m3 digester)	kJ/hr.	25261364
4	Heat content in blown pulp (for 100 m3 digester)	kJ/hr	9572727
5	Flash steam generated by 80 m3 digester	MT/hr	11
6	Flash steam generated by 100 m3 digester	MT/hr	4
7	Total flash steam generated	MT/hr	15
8	Blow hours per day for 80 m3 digester	hrs/day	7.5
9	Blow hours per day for 100 m3 digester	hrs/day	3
10	Total Peak blow steam rate	MT/hr	35

BHR efficiency

1	Foul condensate inlet temperature to condenser	deg.C	66
2	Foul condensate outlet temperature from condenser	deg.C	85
3	Foul water required to condense this flash vapor	m3/hr	1031
4	Actual flow through primary and secondary cond. Pump	m3/hr	570
5	Mass of foul water / Mass of flash steam	MT / MT	29
6	Present amount of flash vapor condense by above condition	MT/hr	20
7	Efficiency of present BHR system	%	55

Across Plate heat exchanger:

1	Heat available with foul condensate	kJ/hr	29385400
2	Outlet temperature of hot water availability	deg.C	65.57
3	Present available heat transfer area	m2	83

Across the hot water tank

1	Hot water temperature required in new fiber line	deg.C	80
2	LP steam consumption @ 4.5 bar	MT/hr	5.74

Proposed system:

1	Making availability of foul condensate to condenser at temp of	deg.C	55
2	Taking the foul condensate temp from accumulator tank at	deg.C	80
3	Foul condensate required to condense flash vapor	m3/hr	791
4	Hot water available temperature across PHE with exisitng foul condensate flow across the PHE	deg.C	75
5	Additional LP steam required in HWT @ 80 oC hot water	MT/hr	2

Requirements

Across the PHE

1	Flow of Hot water	m3/hr	230
2	LMTD	deg.C	11
3	Heat transfer area required in PHE @ U= 5800 w/m2 oC	m2	174
4	Additional heat transfer area required compared to exisiting PHE	m2	91

Across the Condenser

1	Present flow rate to condenser	m3/hr	570
2	Additional FC flow required to condense the vapor	m3/hr	221
3	With present flow of FC across the condenser, amount of vapor condense	MT/hr	25

Condensation of remaining 35-25=10 MT/hr vapor taken care by additional flow of 221 m3/hr FC. For this pump (PC & SC) capacity have to increase.

Energy savings:

1	Present steam consumption across the hot water tank	MT/hr	5.74
2	Proposed steam consumption across the hot water tank	MT/hr	2
3	Saving in steam consumption	MT/hr	3.84
4	Expected annual steam savings @ Rs. 300 /MT	Rs. Lakh	91.24
5	Cost of implementation	Rs. Lakh	15
6	Simple pay back period	months	2

Pay back period:

1	Depreciation per annum	Rs. lakh	1.5
2	Enterprise tax per annum	Rs. lakh	13.46
3	Increased cash flow per annum	Rs. lakh	77.78
4	Pay back period	months	2

Appendix-III/1

Mass and Energy calculation for the 7 effect evaporator, feed to 5th calendria.

Effects	1	2	3	4	5	6	7
Delta T, deg.C	8.4	9.3	5.8	5.7	4.4	6.0	4.5
Liquor inlet temp, deg.C	99.5	88.2	77.5	53.5	88.0	70.5	61.5
Liquor outlet temp, deg.C	124.6	99.5	88.2	77.5	70.5	61.5	53.5
Body pressure, bar or mm of Hg	2.3	0.9	0.3	92.6	166.8	475.8	586.1
Body temp, deg.C	133.0	108.8	94.0	83.2	75.0	67.5	58.0
Vapor separation temp, deg.C	114.6	92.5	83.2	75.0	67.5	58.0	49.5
BPR, deg.C	10.0	7.0	5.0	2.5	3.0	3.5	4.0
Latent heat of vapor in, kJ/kg	2161.9	2230.0	2272.8	2306.9	2324.5	2332.0	2349.2
Latent heat of vapor out, kJ/kg	2230.0	2275.4	2306.9	2324.5	2332.0	2349.2	2374.2
Solids in, % TS	0.50	0.40	0.33	0.29	0.18	0.20	0.24
Solids out, % TS	0.69	0.50	0.40	0.33	0.20	0.24	0.29
Liquor flow in, kg/hr	61267.1	77642.6	92819.3	104341.0	163000.0	144810.5	123477.2
Liquor flow out,kg/hr	44500.0	61267.1	77642.6	92819.3	144810.5	123477.2	99472.9
Water evaporated, kg/hr	16767.1	16375.5	15176.7	11521.7	18189.5	21333.3	24004.2
Avg. sp. Heat of liquor, kJ/kg oC	3.8	4.3	4.6	4.1	5.1	5.0	4.8
Total hear transfer area, m2	1570.0	1570.0	1570.0	1570.0	1570.0	1570.0	1570.0

Solids out from PLFV, %TS	ratio	0.703231
Steam requirement @2.4 bar	kg/hr	20000
Water evaporated	tph	121278.3
Steam economy	**ratio**	**6.06**

Energy Balances:-

Effect-1

Heat of steam	kJ/hr	43237800
Heating of liquor	kJ/hr	5847149
Liquor flow in	kg/hr	61267.11
Water evaporated	kg/hr	16767.11

Effect-2

Heat of vapor	kJ/hr	37390651
Heat of SCFV-1	kJ/hr	2018600
Heat of PLFV	kJ/hr	1655623
Heating of liquor	kJ/hr	3804019
Liquor flow in	kg/hr	77642.63
Water evaporated	kg/hr	16375.52

Effect-3

Heat of vapor	kJ/hr	37260855
Heat of SCFV-2	kJ/hr	1208701
Heat of PCFV-1	kJ/hr	1117755
Heating of liquor	kJ/hr	4575675
Liquor flow in	kg/hr	92819.28
water evaporated	kg/hr	15176.66

Effect-4

Heat of vapor	kJ/hr	35011636
Heat of SCFV-3	kJ/hr	859466.3
Heat of PCFV-2	kJ/hr	1177533
Heating of liquor	kJ/hr	10266468
Liquor flow in	kg/hr	104341
Water evaporated	kg/hr	11521.74

Effect-5

Heat of vapor	kJ/hr	26782167
Heat of PCFV-3	kJ/hr	1048103
Cooling of liquor	kJ/hr	14588109
Liquor flow out	kg/hr	144810.5
Water evaporated	kg/hr	18189.54

Effect-6

Heat of vapor	kJ/hr	42418378
Heat of PCFV-4	kJ/hr	1180448
Cooling of liquor	kJ/hr	6516471
Liquor flow out	kg/hr	123477.2
Water evaporated	kg/hr	21333.28

Effect-7

Heat of vapor	kJ/hr	50115296
Heat of PCFV-5	kJ/hr	2135013
Cooling of liquor	kJ/hr	4741523
Liquor flow out	kg/hr	99472.93
Water evaporated	kg/hr	24004.24

PLFV

Flow of liquor to PLFV	kg/hr	44500
Temperature of liquor flow to HBL tank	deg.C	114
Liquor temperature to PLFV	deg.C	124.63
BPR	deg.C	11
Vapor temperature	deg.C	103
% conc. Of black liquor in	ratio	0.691592
% conc. of black liquor out	ratio	0.703231
Avg. Sp. Heat of black liquor	kJ/kg oC	3.5
Latent of vapor	kJ/kg	2248
Flash vapor flow	kg/hr	736.4869
Heat of vapor to 2nd calendria	kJ/hr	1655623

SCFV-1

Condensate flow in	kg/hr	20000
Temperature of condensate in	deg.C	133
Vapor temperature	deg.C	108.815
Sensible heat of condensate in	kJ/kg	558.03
Sensible heat of condensate out	kJ/kg	457.1
% of flash vapor	%	4.526009
Flash vapor to Cal-2	kg/hr	905.2018
Heat of vapor to cal-2	kJ/hr	2018600

SCFV-2

Condensate flow in	kg/hr	19094.8
Temperature of condensate in	deg.C	108.815
Vapor temperature	deg.C	94
Sensible heat of condensate in	kJ/kg	457.1
Sensible heat of condensate out	kJ/kg	393.8
% of flash vapor	%	2.785111
Flash vapor to Cal-3	kg/hr	531.8113
Heat of vapor to cal-3	kJ/hr	1208701

SCFV-3

Condensate flow in	kg/hr	18562.99
Temperature of condensate in	deg.C	94
Vapor temperature	deg.C	83.2
Sensible heat of condensate in	kJ/kg	393.8
Sensible heat of condensate out	kJ/kg	347.5
% of flash vapor	%	2.006988
Flash vapor to Cal-4	kg/hr	372.5568
Heat of vapor to cal-4	kJ/hr	859466.3

PCFV-1

Condensate flow in from cal-2	kg/hr	18408.8
Condensate flow in from cal-3	kg/hr	17399.13
Condensate temp in from cal-2	deg.C	108.815
condensate temp in from cal-3	deg.C	94
Vapor temp	deg.C	94
Sensible heat in from cal-2	kJ/kg	457.1
Sensible heat in from cal-3	kJ/kg	393.8
Sensible heat out	kJ/kg	393.8
% flash vapor	%	1.392555
Flash vapor flow	kg/hr	491.7966
Heat of vapor to Cal-3	kJ/hr	1117755

PCFV-2

Condensate flow in from cal-4	kg/hr	16059.64
Condensate flow in from PCFV-1	kg/hr	35316.13
Condensate temp in from cal-4.	deg.C	83.2
condensate temp in from PCFV-1	deg.C	94
Vapor temp	deg.C	83.2
Sensible heat in from cal-4	kJ/kg	347.5
Sensible heat in from PCFV-1	kJ/kg	393.8
Sensible heat out	kJ/kg	347.5
% flash vapor	%	1.003494
Flash vapor flow	kg/hr	510.4305
Heat of vapor to Cal-4	kJ/hr	1177533

PCFV-3

Condensate flow in from cal-5	kg/hr	11972.63
Condensate flow in from PCFV-2	kg/hr	50865.34
Condensate temp in from cal-5.	deg.C	74.96
condensate temp in from PCFV-2	deg.C	83.2
Vapor temp	deg.C	74.96
Sensible heat in from cal-5	kJ/kg	313.9
Sensible heat in from PCFV-2	kJ/kg	347.5
Sensible heat out	kJ/kg	313.9
% flash vapor	%	0.722739
Flash vapor flow	kg/hr	450.8959
Heat of vapor to Cal-5	kJ/hr	1048103

PCFV-4

Condensate flow in from cal-6	kg/hr	18695.73
Condensate flow in from PCFV-3	kg/hr	62387.08
Condensate temp in from cal-6	deg.C	67.52
condensate temp in from PCFV-3	deg.C	74.96
Vapor temp	deg.C	67.52
Sensible heat in from cal-6	kJ/kg	284.6
Sensible heat in from PCFV-3	kJ/kg	313.9
Sensible heat out	kJ/kg	284.6
% flash vapor	%	0.628211
Flash vapor flow	kg/hr	506.191
Heat of vapor to Cal-6	kJ/hr	1180448

PCFV-5

Condensate flow in from cal-7	kg/hr	22242.12
Condensate flow in from PCFV-4	kg/hr	80576.62
Condensate temp in from cal-7	deg.C	58.02
condensate temp in from PCFV-4	deg.C	67.52
Vapor temp	deg.C	58.02
Sensible heat in from cal-7	kJ/kg	242.7
Sensible heat in from PCFV-4	kJ/kg	284.6
Sensible heat out	kJ/kg	242.7
% flash vapor	%	0.891808
Flash vapor flow	kg/hr	908.8408
Heat of vapor to Cal-7	kJ/hr	2135013

Appendix-III/2

Performance of Calendria 1 & 2 for Controlling Water Soluble Scales

Effect-2

Time, hrs	2.30	3.00	3.30	4.00	4.30
Liquor temperature in, deg.C	87	87.8	87.9	86.7	86.4
Liquor temperature out, deg.C	102.72	102.5	101	100	99.5
Body pressure, bar	1.5	1.51	1.53	1.48	1.48
Next body pressure, mm Hg	30.53	18.06	12.48	8.7	8.9
Body temperature, deg.C	112	113	114	114.5	114.5
Vapor separation temp, deg.C	95.72	95.07	94.78	94.49	94.56
Delta T, deg.C	9.28	10.5	13	14.5	15
BPR, deg.C	7	7.43	6.22	5.51	4.94
Solids in, % TS	46.13	46.13	46.31	47.77	47.77
Solids out, % TS	52.48	52.48	51.83	53.65	53.65
Liquor flow in , kg/hr	78000	78000	78000	78000	77000
Liquor flow out, kg/hr	68562.12	68562.12	69692.84	69451.26	68560.86
Water evaporated, kg/hr	9437.88	9437.88	8307.16	8548.74	8439.14
Avg. sp heat of liquor, kJ/kg deg.C	4.3	4.3	4.3	4.3	4.3
Latent heat of vapor out, kJ/kg	2263.47	2265.97	2265.97	2268.91	2268.91
Balance heat of body, kJ/hr	26634849	26316335	23217511	23857146	23485065
Heat transfer area, m2	1570	1570	1570	1570	1570
Overall heat transfer co-efficient, w/m2 deg.C	507.81	443.44	315.99	291.10	277.01

Effect-IB(57%)

Time, hrs	2.30	3.00	3.30	4.00	4.30
Liquor temperature in, deg.C	102.72	102.5	101	100	99.5
Liquor temperature out, deg.C	132.4	132.32	132.02	132.25	127.56
Body pressure, bar	2.73	2.7	2.73	2.74	2.55
Next body pressure, mm Hg	1.5	1.51	1.53	1.48	1.48
Body temperature, deg.C	140.8	140.48	140.99	140.94	136.06
Vapor separation temp, deg.C	124	123	124	122	122
Delta T, deg.C	8.4	8.16	8.97	8.69	8.5
BPR, deg.C	8.4	9.32	8.02	10.25	5.56
Solids in, % TS	52.48	52.48	51.83	53.65	53.65
Solids out, % TS	58.48	58.48	57.83	58.85	58.85
Liquor flow in , kg/hr	66663.08	66663.08	65802.49	65647.07	65022.22
Liquor flow out, kg/hr	59823.50	59823.50	58975.33	59846.48	59276.84
Water evaporated, kg/hr	6839.58	6839.58	6827.16	5800.59	5745.38
Avg. sp heat of liquor, kJ/kg deg.C	3.8	3.8	3.8	3.8	3.8
Latent heat of vapor out, kJ/kg	2187.812	2190.74	2187.812	2193.664	2193.66
Balance heat of body, kJ/hr	22482238	22537729	22693087	20769595	19536597
Heat transfer area, m2	1570	1570	1570	1570	1570
Overall heat transfer co-efficient, w/m2 deg.C	473.54	488.67	447.61	422.87	406.66

Effect-IA(68%)

Time, hrs	2.30	3.00	3.30	4.00	4.30
Liquor temperature in, deg.C	132.4	132.32	132.02	132.25	127.56
Liquor temperature out, deg.C	137.14	136.67	136.28	136.04	134.6
Body pressure, bar	2.7	2.67	2.69	2.7	2.69
Next body pressure, mm Hg	1.5	1.51	1.53	1.48	1.48
Body temperature, deg.C	141.31	141.02	141.09	140.93	139.6
Vapor separation temp, deg.C	126	127	127	125	125
Delta T, deg.C	4.17	4.35	4.81	4.89	5
BPR, deg.C	11.14	9.67	9.28	11.04	9.6
Solids in, % TS	58.48	58.48	57.83	58.85	58.85
Solids out, % TS	65.98	65.98	65.33	66.1	65.85
Liquor flow in , kg/hr	66663.08	66663.08	65802.49	65647.07	65022.22
Liquor flow out, kg/hr	59085.43	59085.43	58248.25	58446.75	58110.21
Water evaporated, kg/hr	7577.65	7577.65	7554.24	7200.32	6912.01
Avg. sp heat of liquor, kJ/kg deg.C	3.5	3.5	3.5	3.5	3.5
Latent heat of vapor out, kJ/kg	2181.96	2179.034	2179.034	2184.886	2184.886
Balance heat of body, kJ/hr	17640062	17526894	17442067	16602690	16704091
Heat transfer area, m2	1570	1570	1570	1570	1570
Overall heat transfer co-efficient, w/m2 deg.C	748.45	712.88	641.58	600.71	591.09

Appendix-III/3

Present and Recommended Cycle of Evaporation

5 hr cycle	Present Cycle time of evaporator operation																							
Body	6	7	8	9	10	11	12	1	2	3	4	5	6	7	8	9	10	11	12	1	2	3	4	5
1D	57% body					wbl cir		PC cir		PC	68% body					57% body					wbl cir		PC cir	
1C	wbl cir		PC cir		PC	68% body					57% body					wbl cir		PC cir		PC	68% body			
1A	68% body					57% body					wbl cir		PC cir		PC	68% body					57% body			
1B	2 body					wbl cir		PC cir		PC	2 body					wbl cir		PC cir		PC	2 body			
2	wbl cir		PC cir		PC	2 body					wbl cir		PC cir		PC	2 body					wbl cir		PC cir	

Washing with WBL	hrs/cycle	2
Washing with PC	hrs/cycle	3
Process condensate addition in WBL tank	m3/day	250
Steam required to evaporate this water @ 6.06 steam economy and 2.4 bar pressure	MT/day	41.25
Equivalent cost of thermal energy @ Rs.300/ton	Rs.lakh/annum	41
Down time to attain product liquor after change of body	hrs/day	1.25
Equivalent cost of downtime in terms of steam generation at 80 tph	Rs.lakh/annum	99
Total cost per annum	Rs.lakh/annum	140

8 hrs cycle	Recommended cycle time of operation																							
Body	6	7	8	9	10	11	12	1	2	3	4	5	6	7	8	9	10	11	12	1	2	3	4	5
1D	57% body								WBL cir					PC cir		PC	68% body							
1C	wbl cir					PC cir		PC	68% body								57% body							
1A	68% body								57% body								WBL cir					PC cir		PC
1B	2 body								WBL cir					PC cir		PC	2 body							
2	wbl cir					PC cir		PC	2 body								WBL cir					PC cir		PC

Washing with WBL	hrs/cycle	5
Washing with PC	hrs/cycle	3
Process condensate addition in WBL tank	m3/day	150 (Calculated proportionally)
Steam required to evaporate this water @ 6.06 steam economy and 2.4 bar pressure	MT/day	24.75
Equivalent cost of thermal energy @ Rs.300/ton	Rs.lakh/annum	25
Down time to attain product liquor after change of body	hrs/day	0.75
Equivalent cost of downtime in terms of steam generation at 80 tph	Rs.lakh/annum	59.4
Total cost per annum	Rs.lakh/annum	84

6 hrs cycle	Recommended cycle time of operation																							
Body	6	7	8	9	10	11	12	1	2	3	4	5	6	7	8	9	10	11	12	1	2	3	4	5
1D	57% body						WBL cir			PC cir		PC	68% body						57% body					
1C	WBL cir			PC cir		PC	68% body						57% body						WBL cir			PC cir		PC
1A	68% body						57% body						WBL cir			PC cir		PC	68% body					
1B	2nd body						WBL cir			PC cir		PC	2nd body						WBL cir			PC cir		PC
2	WBL cir			PC cir		PC	2nd body						WBL cir			PC cir		PC	2nd body					

Washing with WBL	hrs/cycle	3
Washing with PC	hrs/cycle	3
Process condensate addition in WBL tank	m3/day	200
Steam required to evaporate this water @ 6.06 steam economy and 2.4 bar pressure	MT/day	33.00
Equivalent cost of thermal energy @ Rs.300/ton	Rs.lakh/annum	33
Down time to attain product liquor after change of body	hrs/day	1
Equivalent cost of downtime in terms of steam generation at 80 tph	Rs.lakh/annum	79.2
Total cost per annum	Rs.lakh/annum	112

WBL cir	Weak black liquor circulation
PC cir	Process condensate circulation
PC	Process condensate circulation by closing the inlet and outlet valves of body

Appendix-IV/1

Energy saving by Indirect black liquor heater

Process description

1	Flow of black liquor from AMT	MT/hr	42.23
2	Black liquor conc.from AMT	%TS	68.32
3	Temperature of black liquor from AMT	deg.C	100.44
4	Temperature of black liquor to firing	deg.C	120
5	Sp. Heat of black liquor	kJ/kg deg.C	2.96
6	Solids firing	MT/hr	28.85
7	Steam generation	MT/hr	83
8	Steam generation pressure	bar	64
9	Steam generation temperature	deg.C	480
10	MP steam consumption in direct heater	kg/hr	1183.5
11	Black liquor conc. from direct heater	% TS	66.46

System Calculation

1	Heat required to preheat the black liquor	kJ/kg	2445015.65
2	Equivalent MP steam required by indirect heating	kg/hr	1227.79
3	Steam generation increase in recovery boiler (avoiding BL dilution)	MT/hr	1.19
4	Additional MP steam required by indirect heating	MT/hr	0.044
5	Amount of condensate will be send to DM plant	MT/hr	1.228

Energy Savings

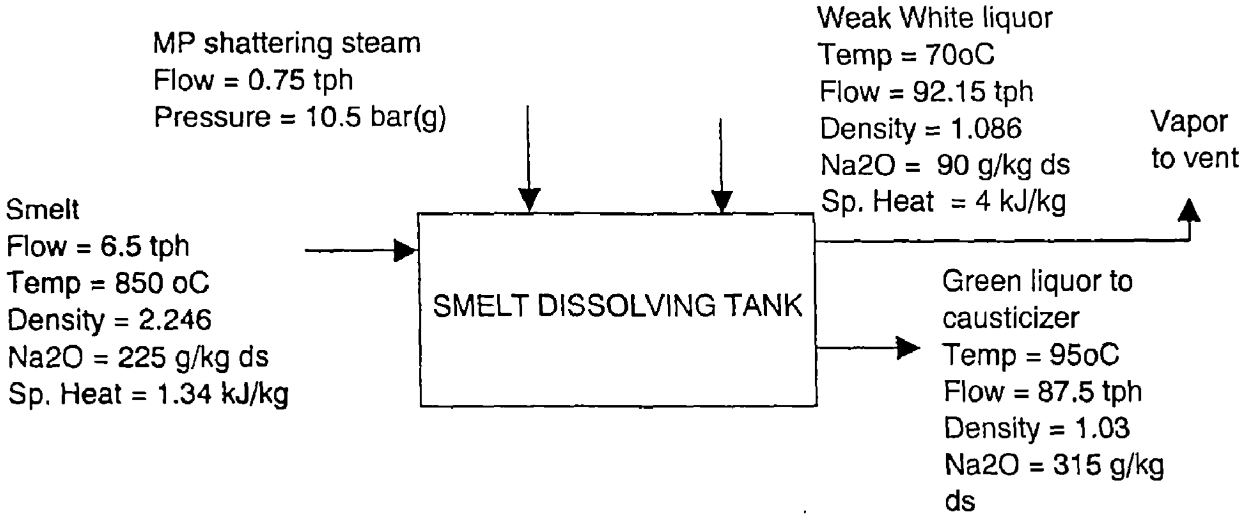
1	Annual cost of condensate saving @ Rs. 20 per cu.m	Rs.lakh	1.94
2	Annual cost of steam saving @ Rs. 300 per tonne of steam	Rs.lakh	27.22
3	Total annual savings	Rs.lakh	29.2
4	Total investment required	Rs.lakh	12
5	Simple pay back period	Months	4.9

Pay back period

1	Depreciation per annum	Rs. lakh	1.2
2	Enterprise tax per annum	Rs. lakh	4.20
3	Increased cash flow	Rs. lakh	24.97
4	Pay back period	months	6

Appendix-IV/2

SMELT DISSOLVING TANK VAPOR RECOVERY



Calculating amount vapor vent to atmosphere

Process Description

1	Smelt		
	Flow	MT/hr	6.5
	Temp	deg.C	850
	Density	t/m3	2.246
	Na2O	g/kg ds	225
	Sp. Heat	kJ/kg	1.34
2	Weak white liquor		
	Flow	MT/hr	92.15
	Temp	deg.C	70
	Density	t/m3	1.086
	Na2O	g/kg ds	90
	Sp. Heat	kJ/kg	4
3	Green liquor		
	Flow	MT/hr	87.5
	Temp	deg.C	95
	Density	t/m3	1.03
	Na2O	g/kg ds	315
	Sp. Heat	kJ/kg	3.65
4	MP shattering steam		
	Flow	MT/hr	0.75
	Heat available	kJ/kg	2781.3

Energy balance

1	Heat in smelt	kJ/hr	7403500
2	Heat in WWL	kJ/hr	25802000
3	Heat in GL	kJ/hr	30340625
4	Heat in shattering steam	kJ/hr	2085975
	Vapor release to atm	kg/hr	1850.093423

Conserve heat from Vent Vapor to Atmosphere

Option - 1

Condensing the vapor by spraying the water in jet type of condenser and exchanging the heat from foul condensate to celan water in hoat exchanger.

Process Parameter

1	Foul condensate temp		
	Inlet to vapor condenser	deg.C	50
	Outlet to vapor condenser	deg.C	80
2	Assuming vapor recovery	%	70
3	Hot water generation temp		
	Inlet Temp	deg.C	35
	Outlet temp	deg.C	80

Calculate foul water requirement and hot water generation in heat exchanger

1	Water required to condense smelt tank vapor	kg/hr	39480
2	Total foul condensate available	kg/hr	40776
3	Hot water generation in heat exchanger	kg/hr	26320
	Equivalent LP steam required	kg/hr	2362.16
	Annual saving @ Rs. 300 per tonne steam	Rs.lakh	56.12
	Total investment required	Rs.lakh	25
	Simple pay back period	months	5

Pay back period

1	Depreciation per annum	Rs.lakh	2.5
2	Enterprise tax per annum	Rs.lakh	8.0
3	Increased cash flow per annum	Rs.lakh	48.1
4	Pay back period	months	6

Option - 2

Condensing the Vapor by passing through surface condenser and using hot process condensate water from evaporator in process, here by handling of foul condensate is avoided.

Process Parameters

1	Hot water generation (Process condensate from evaporator)		
	Inlet Temp	deg.C	54
	Flow to process	m3/hr	75
2	Assuming vapor recovery	%	80

Energy Savings

Available hot water temp.	deg. C	65
Equivalent LP steam required	kg/hr	1593.77
Annual saving @ Rs. 300 per tonne of steam	Rs.lakh	37.87
Total investment required	Rs.lakh	20
Simple pay back period	months	6.3

Pay back period

1	Depreciation per annum	Rs.lakh	2.0
2	Enterprise tax per annum	Rs.lakh	5.4
3	Increased cash flow per annum	Rs.lakh	32.5
4	Pay back period	months	7

Appendix-IV/3

Blow down tank & Air perheater condensate tank flash vapor recovery in SRP

Process Description

1	Total blow down from boiler	% of steam generation	3
2	Steam generation	MT/hr	93
2	Boiler drum pressure	kg/cm2 abs	70
3	Temperature	deg.C	275
4	Boiler efficiency	%	70
5	Silica content in boiler feed water as SiO2	ppm	<0.02
6	Primary air requirement		
	Flow	MT/hr	34.3
	Temperature	deg.C	160
	Temperature increase by LP steam	deg.C	70
	Temperature by MP steam	deg.C	60
7	Secondary air requirement		
	Flow	MT/hr	64
	Temperature	deg.C	150
	Temperature increase by LP steam	deg.C	70
	Temperature increase by MP steam	deg.C	50
8	LP steam pressure	bar	4.5
9	MP steam pressure	bar	10.5
10	LP steam temperature	deg.C	155
11	MP steam temperature	deg.C	185

Steam requirement in primary and secondary Scaph

1	Latent heat of steam @ LP steam	kJ/kg	2098.1
2	Latent heat of steam @ MP steam	kJ/kg	1991.4
3	Mass of LP steam in Primary Scaph	kg/hr	1150.09
4	Mass of MP steam in Primary Scaph	kg/hr	1038.61
5	Mass of LP steam in Secondary Scaph	kg/hr	2145.94
6	Mass of MP steam in Secondary Scaph	kg/hr	1614.94
7	Total LP steam consumption	kg/hr	3296.03
8	Total MP steam consumption	kg/hr	2653.56

Blow down tank flash vapor calculation

1	Blow down condensate from boiler drum	tph	2.79
2	Heat of condensate from boiler drum	kJ/kg	1262.25
3	Heat of condensate from blow down tank	kJ/kg	504.7
4	Latent heat of flash vapor from blow down tank	kJ/kg	2201.6
5	Total Heat of flash vapor	kJ/kg	2706.3
5	Percentage of flash vapor	%	34.41
6	Amount of Vapor recoverable	MT/hr	0.96

Condensate tank flash vapor calculation

1	Heat of LP steam condensate from Scaph	kJ/kg	652.8
2	Heat of LP steam condensate from condensate tank	kJ/kg	504.7
3	Latent heat of flash vapor from condensate tank	kJ/kg	2201.6
4	Total Heat of flash vapor	kJ/kg	2706.3
5	Heat of MP steam condensate from Scaph	kJ/kg	789.9
6	Heat of MP steam condensate from tank	kJ/kg	504.7
7	Percentage of LP steam condensate flash vapor	%	6.73
8	Percentate of MP steam condensate flash vapor	%	12.95
9	Mass of LP steam condensate flash vapor recoverable	kg/hr	221.72
10	Mass of MP steam condensate flash vapor recoverable	kg/hr	343.75

Option-1

The flash vapor from blow down tank and condensate collection tank can be compressed by steam jet ejector using MP steam as motive steam. The compressed vapor (LP steam) can be used back to air pre-heater and deareator.

Thermo-compressor for blow down tank vapor(TC-1) and condensate tank flash vapor(TC-2) separately

1	Compression efficiency of thermocompressor(assumed)	%	90
2	Nozzle efficiency of thermocompressor	%	90
3	Extraction efficiency of thercompressor	%	90
4	Mass of motive steam required TC-1	MT/hr	4.028
5	Mass of motive steam required TC-2	MT/hr	3.44

System Calculation

1	Mass of LP steam from TC-1	MT/hr	4.99
2	Mass of LP steam from TC-2	MT/hr	4.01
3	Total mass of motive steam required @ 11.3 bar abs	MT/hr	7.47
4	Total LP steam available from the thermocompressor	MT/hr	8.99
5	Heat of MP steam	kJ/kg	2781.30
6	Heat of LP steam	kJ/kg	2750.90
7	Thermal efficiency of TC-1	%	**22.48**
8	Thermal efficiency of TC-2	%	**15.17**

Energy Saving Cost

1	LP steam saving in deareator	MT/hr	3.30
2	LP steam saving in air preheater	MT/hr	5.70
3	Motive steam required (MP steam)	MT/hr	7.47
4	Total saving in steam consumption	MT/hr	1.53
5	Annaul saving @ Rs. 300 per tonne of steam	Rs.lakh	36.25
6	Total investment required	Rs. lakh	10.00
7	Simple pay back period	Months	3

Pay back period

1	Depreciation per annum	Rs. lakh	1.00
2	Enterprise tax per annum	Rs. lakh	5.29
3	Increased cash flow per annum	Rs. lakh	30.96
4	Pay back period	Months	4

Option - 2

Flash vapor can be conserved by passing the vapor to heat exchanger where it gives its latent heat to heat feed water supplying to deareator.

System Calculation

1	Temperature of feed water to deareator	deg.C	60
2	Flow of feed water to deareator	MT/hr	93
3	By conserving the flash vapor in HE feed water avilable temp to deareator	deg.C	68.64
4	Thermal efficiency	%	81.35

Energy Saving Cost

1	LP steam saving in deareator	MT/hr	1.60
2	Annual saving @ Rs. 300 per tonne of steam	Rs. lakh	38.03
3	Total investment required	Rs. lakh	12.5
4	Simple pay back period	Months	4

Pay back period

1	Depreciation per annum	Rs. lakh	1.25
2	Enterprise tax per annum	Rs. lakh	5.52
3	Increased cash flow per annum	Rs. lakh	32.52
4	Pay back period	Months	5

Appendix-IV/4

Lime Kiln furnace oil heater condensate recovery

Process description

1	Furnace oil consumption	MT/hr	0.94
2	Density of oil	t/m3	0.98
3	Furnace oil inlet temperature	deg.C	81
4	Steam pressure	bar(g)	9.3
5	Furnace oil outlet temperature	deg.C	128
6	Latent heat of steam	kJ/kg	2010
7	Sensible heat of codensate leaving heater	kJ/kg	767.3
8	Sensible heat of codensate draining	kJ/kg	419.1
9	Latent heat of steam @ atm pressure	kJ/kg	2256.9
10	Heat value of steam @ atm pressure	kJ/kg	2676

Energy Savings

Steam consumption in heater	kg/hr	241.78
Flash steam available in condensate	%	15.43
Mass of flash steam available	kg/hr	37.30
Equivalent LP steam saving	kg/hr	40.13
Annual saving @ Rs. 300 per tonne of steam	Rs.lakh	0.95
Cost of condensate saving per annum	Rs.lakh	0.38
Total saving per annum	Rs.lakh	1.34
Investment required	Rs.lakh	Marginal
Payback period	Months	Nil

SRB furnace oil heater condensate recovery

Process description

1	Furnace oil in circulation	MT/hr	0.45
2	Density of oil	t/m3	0.98
3	furnace oil inlet temperature	deg.C	100
4	Steam pressure	bar(g)	10.5
5	Furnace oil outlet temperature	deg.C	125
6	Latent heat of steam @ 10.5 bar	kJ/kg	1991.4
7	Sensible heat of codensate leaving heater	kJ/kg	789.4
8	Sensible heat of codensate draining	kJ/kg	419.1
9	Latent heat of steam @ atm pressure	kJ/kg	2256.9
10	Heat value of steam @ atm pressure	kJ/kg	2676
11	Steam valve opening period	hrs/day	8.5
12	Avg. fuel consumption per month	kl/month	101.47
13	Avg. fuel consumption per hour	MT/hr	0.138

Energy Savings

Steam consumption in heater	kg/day	528.21
Steam required to pre-heat oil	kg/day	457.37
Total steam consumption in heater	kg/day	985.58
Flash steam available in condensate	%	16.41
Mass of flash steam available	kg/day	161.71
Equivalent LP steam saving	kg/day	173.95
Annual savings @ Rs. 300 per tonne steam	Rs.lakh	0.17
Cost of condensate saving per annum	Rs.lakh	0.07
Total saving per annum	Rs.lakh	0.24
Investment required	Rs.lakh	Marginal
Payback period		Nil

Energy saving by proper condensate recovery of sootblowing steam

Process description

1	Nmber of soot blowers	-	26
2	Steam consumption	MT/hr	4.5
3	Working pressure	bar	25
4	Working temperature	deg.C	350
5	Total time for each soot blower	minutes	8
6	Total time for sootblowing in a shift	hours	3.45
7	Orifice dia in condensate line	mm	8
8	Condensate flow velocity in line	m/s	2
9	Condensate drian temperature	deg.C	250
10	Diameter of the condensate drain pipe	mm	25.4
11	Sp. Volume of condensate	m3/kg	0.001251
12	Sensible heat of condensate	kJ/kg	1085.8
13	Latent heat of steam @ 1 bar	kJ/kg	2256.9

Energy Savings

1	Mass flow of condensate during sootblowing period	kg/day	30168
2	Flash vapor available from condensate	%	30
3	Mass of steam with condensate drain	kg/day	8912
4	Equivalent mass of LP steam	kg/day	8695
5	Annaul Cost of LP steam @ Rs. 300 per tonne	Rs.lakh	9
6	Investment required	Rs.lakh	Marginal
7	Payback period	Months	Nil

Appendix- V

Flux maxiox for furnace oil saying at Lime kiln

Trial Conclusion

Pre-installation status:	**UOM**	**Qty**
Total number of days considered for evaluation	-	47
Total product lime production	MT	5883
Total quantity of lime sludge handled at lime kiln	MT	7090
Average moisture in lime sludge during trial	%	44
Total water quantity in sludge handled	MT	3120
Total quantity of seashell handled at lime kiln	MT	2362
Constant moisture considered in seashell	%	3.2
Total water quantity in seashell handled	MT	76
Total water handled at limekiln	MT	3195
Furnace oil consumed during above period	kl	1087
Specific oil consumption for water evaporation	liter/MT	340
Specific oil consumption per ton of production	liter/MT	185
Post installation status:		
Total number of days considered for evaluation	-	46
Total product lime production	MT	5836
Total quantity of lime sludge handled at lime kiln	MT	6754
Average moisture in lime sludge during trial	%	45
Total water quantity in sludge handled	MT	3039
Total quantity of seashell handled at lime kiln	MT	2614
Constant moisture considered in seashell	%	3.2
Total water quantity in seashell handled	MT	84
Total water handled at limekiln	MT	3123
Furnace oil consumed during above period	kl	1052
Specific oil consumption for water evaporation	liter/MT	337
Specific oil consumption per ton of production	liter/MT	180
Energy Savings:		
Reduction in specific oil consumption for water evaporation	liter/MT	3
Savings:	%	1
Reduction in specific oil consumption per ton of production	liter/MT	5
Savings:	%	2
Avg. water evaporated from sludge and seashell	MT/day	68
Average lime kiln production	MT/day	127
Daily saving in Furnace Oil for water evaporation	liter/ day	204
Daily saving in Furnace Oil per ton of lime kiln production.	liter/ day	573
Water Evaporation- Savings		
Annual saving in Furnace Oil (330 days/annum)	kl/annum	67
Cost of FO	Rs/liter	12
Annual Savings	Rs.lakh	8

Specific Furnace Oil consumption - Savings		
Annual saving in Furnace Oil (330 days/annum)	kl/annum	189
Cost of Furnace Oil	Rs/liter	12
Annual Savings	Rs.lakh	23
Steam saving at Furnace Oil heater		
Furnace oil intlet temperature to heater	deg.C	81
Furnace oil outlet temperature to heater	deg.C	128
Density of furnace oil	MT/m3	0.98
Steam saving at Furnace Oil heater	MT/day	192
Annual savings @ Rs. 300 per ton	Rs. lakh	190
Total savings per annum	Rs. lakh	213
Total investment required	Rs. lakh	2
Simple pay back period	months	0.11
Pay back period		
Depreciation per annum	Rs. lakh	0.20
Enterprise tax per annum	Rs. lakh	31.85
Increased cash flow per annum	Rs. lakh	180.70
Pay back period	months	0.13

Appendix- VI/1
Rapid displacement pulping

Process description:

1	Chip feed on BD basis		
	80 m3 digester	MT/digester	19
	100 m3 digester	MT/digester	24
2	Moisture in chips	% on od	30
3	White liquor requirement	% on od	17.5
4	Strength of white liquor	gpl as Na2O	88
5	White liquor feed temperature	deg.C	85
6	Yield of pulping	%	45
7	Cooking temperature	deg.C	162
8	Chip feed temperature	deg.C	30
9	Initial temperature of digester shell	deg.C	75
10	Black liquor feed temperature	deg.C	85
11	Black liquor feed concentration	% TS	18
12	Sp. Heat of white liquor	kJ/kg oc	3.8
13	Sp . Heat of black liquor	kJ/kg oc	4
14	Sp. Heat of chips	kJ/kg oc	1.38
14	Sp . Heat of digester shell	kJ/kg oc	0.627
15	Sp . Gravity of white liquor	-	1.14
16	Sp . Gravity of black liquor	-	1.08
17	Bath ratio	-	4.2
18	Number of digester		
	80 m3 digester		6
	100 m3 digester		2
19	Shell weight of digester		
	80 m3 digester	MT	35
	100 m3 digester	MT	44
20	Sp. Gravity of shell		7.86
21	Number of blows per day		
	80 m3 digester	-	30
	100 m3 digester	-	9

General Assumption:

1	Hot black liquor solids	%TS	21.5
2	Warm black liquor solids	%TS	18.36
3	Cool black liquor solids	%TS	17.8
4	White liquor charge during cool pad	% AA od	3
5	White liquor charge during warm fill	% AA od	3
6	White liquor charge during hot fill C1	% AA od	1
7	White liquor charge during hot fill C2	% AA od	10.5
8	C2 Hot black liquor accumulator	deg. C	155
9	C1 Hot black liquor accumulator	deg. C	140
10	HW Hot white liquor tank	deg. C	160
11	B Warm black liquor accumulator	deg. C	120
12	A Cool black liquor tank	deg. C	90

Energy calculation: Basis on 80 m3 digester

Step-1 Chip filling and steam packing

Asuuming after presteaming bring the chips temperature to 75 oC

sl. No	Description	Mass, kg/cook	Sp. Heat, kJ/kg oC	Temp difference, oC	Heat requirement, kJ/cook
1	Heating of chips	19000	1.38	45	1179900
2	Heating of moisture in chips	5700	4.18	45	1072170
3	Heat given by cool black liquor	16684.5	4	15	1001070
4	White liquor charge @ 3% on od chips	7375.8	3.8	10	280280.4

Enthalpy of LP steam @ 4 bar	kJ/kg	2107.4
LP steam requirement	kg/cook	460.6243

Step-2 Warm liquor filling

Assuming the warm liquor available at 120 oC(B) and final temperature of the system is T1 oC and taking warm liquor feed equivalent to sum of white and black liquor feed.

sl. No	Description	Mass, kg/cook	Sp. Heat, kJ/kg oC	Temp difference, oC	Heat requirement, kJ/cook
1	Heating of chips	19000	1.38	T1-75	26220(T1-75)
2	Heating of moisture in chips	5700	4.18	T1-75	23826(T1-75)
3	Heating of shell	35000	0.627	T1-75	21945(T1-75)
4	Heat lost by warm liquor	108612	4	120-T1	434448(120-T1)
5	White liquor charge @ 3% on od chips	7375.8	3.8	85-T1	28028.04(85-T1)
6	Liquor displacement	75934.8	4	T1-90	303739.2(T1-90)

Final temperature of the bath in digester after warm liquor feed(T1) = 104.09 deg. C

Step-3 Hot white and black liquor filling (Hot fill C1)

Assuming the white liquor available temp at 160 oC and black liquor available temp at 140 oC(C1)

Let final temperature of the system after addition of hot W/L and BL is T2 oC

sl. No	Description	Mass, kg/cook	Sp. Heat, kJ/kg oC	Temp difference, oC	Heat requirement, kJ/cook
1	Heating of chips	19000	1.38	T2-104.09	26220(T2-104.09)
2	Heating of moisture in chips	5700	4.18	T2-104.09	23826(T2-104.09)
3	Heating of shell	35000	0.627	T2-104.09	21945(T2-104.09)
4	Heat given by hot black liquor	16087	4	140 - T2	64346.9(140 - T2)
5	White liquor charge @ 1% od chips	2461.36	3.8	160 - T2	9353.18(160-T2)
6	Liquor displacement	18975.6	4	T2-120	75902.4(T2-120)

Final temperature of the bath in digester after hot liquor addition(T2) = 122.33 deg. C

Step-4 Hot white and black liquor filling (Hot Fill C2)

Assuming the white liquor available temp at 160 oC and black liquor available temp at 155 oC(C2)

Let final temperature of the system after addition of hot W/L and BL is T3 oC

sl. No	Description	Mass, kg/cook	Sp. Heat, kJ/kg oC	Temp difference, oC	Heat requirement, kJ/cook
1	Heating of chips	19000	1.38	T3-122.33	26220(T3-122.33)
2	Heating of moisture in chips	5700	4.18	T3-122.33	23826(T3-122.33)
3	Heating of shell	35000	0.627	T3-122.33	21945(T3-122.33)
4	Heat given by hot black liquor	52477	4	155 - T3	209908.8(155 - T3)
5	White liquor charge @ 10.5% od chips	25844.32	3.8	160 - T3	98208.41(160-T3)
6	Liquor displacement to B	28080	4	T3-120	112320(T3-120)
7	Liquor displacement to C1	48373.2	4	T3-140	193492.8(T3-140)

Final temperature of the bath in digester after hot liquor addition(T2) = 142.32 deg. C

Step-5 Time to temperature

Taking the cooking temperature 162 oC

sl. No	Description	Mass, kg/cook	Sp. Heat, kJ/kg oC	Temp difference, oC	Heat requirement, kJ/cook
1	Heating of chips	19000	1.38	19.68	516010
2	Heating of moisture in chip	5700	4.18	19.68	468896
3	Heating of shell	35000	0.627	19.68	431878
4	Heating of black liquor	74239	4	19.68	5844110

Enthalpy of MP steam @ 11 bar	kJ/kg	1984.3
MP steam requirement (direct heating)	kg/cook	3659

Exchanging of heat in heat exchanger

Assuming that the hot black liquor in hot black liquor (C1) is available at 140 oC.

a. Hot black liquor exchanges heat in heat excahnger 1 to white liquor

Intial temperature of white liquor available	deg.C	85
Final temperature of white liquor	deg.C	120
Intial temperature of black liquor	deg.C	140
Final temperature of black liquor to B fill.	deg.C	120
White liquor flow across HE	kg/cook	28306.2
Black liquor flow across HE	kg/cook	47059.06

b. White liquor preheating to 160 oC by MP steam in heat exchanger 2

Intial temperature of white liquor available	deg.C	120
Final temperature of white liquor	deg.C	160
White liquor flow across HE	kg/cook	28306.2
Enthalpy of steam @ 11 bar	kJ/kg	1984.3
MP Steam required	kg/cook	2168

c. Flash vapor recovery from MP steam condensate

sensible heat of condensate	kJ/kg	798.4
sensible heat of condensate @ atm pr	kJ/kg	419.1
Latent heat of vapor recoverable @ atm pr	kJ/kg	2256.9
% of flash vapor recoverable	%	16.80624
vapor recoverable @ atm pr	kg/cook	510.4633

d. Black liquor from C1 fill prehated to C2 fill Temperature 155 oC in heat exchanger 3

Intial temperature of black liquor	deg.C	140
Final temperature of black liquor to C2 fill.	deg.C	155
Black liquor feed to heat exchanger	kg/cook	28741
Latent heat of MP steam @11 bar	kJ/kg	1984.3
MP steam required	kg/cook	869.0521

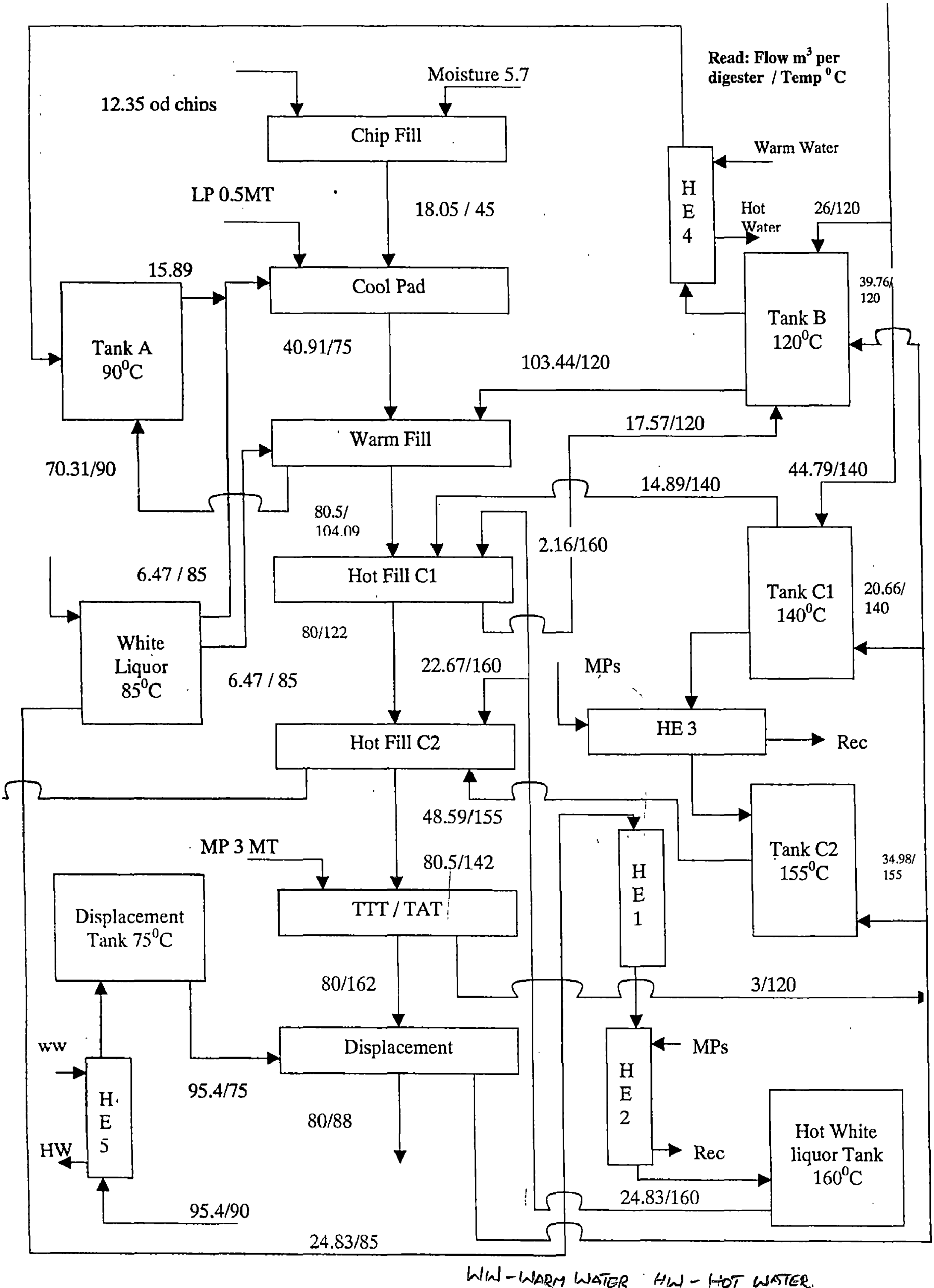
e. Hot water generation

Warm water temperature	deg.C	50
Hot water temperature	deg.C	80
In Heat exchanger -4		
Black liquor flow from B fill	kg/cook	32400
Intial temperature of black liquor	oC	120
Final temperature of black liquor to C2 fill.	oC	90
Amount of hot water generation	kg/cook	31004.78

In Heat exchanger -5		
Filtrate flow to displacement tank	kg/cook	95400
Inlet temperature of filtrate	deg.C	90
Final temperature of filtrate to displacement tank	deg.C	75
Amount of hot water generation	kg/cook	47700
In heat exchanger-6		
Available flash vapor	kg/cook	510.4633
Amount of hot water generation	kg/cook	9187.119
Total hot water generation	kg/cook	87891.9

Energy Savings (Comparative made to 80 m3 digester production with conventional process)

MP steam consumption in present process	kg/cook	6687
MP steam savings compared to conventional process	kg/cook	5102
Equivalent power generation in TG @ 30 cooks / day	kW	102
LP steam consumption in this process	kg/cook	460.62
Heat recovered by hot water generation its equivalent LP steam savings	kg/cook	5230
LP required for additional hot water generation of 54 m3 per cook compared to conventional process	kg/cook	4840
Additional LP steam required by this process	kg/cook	5301
Savings with respect to grid power		
Annual savings compared to conventional process @ Rs. 3.25 per kwh grid	Rs. Lakh	26



Appendix-VI/1.1 : Process Flow Description of RDH Process (Calculated on basis of present 80m³ digester

Appendix- VI/2

Heat treatment of Black liquor

Operational Details:

1. Feeding details of the black liquor to the evaporator plant:

   Concentration level = 18%

   Black liquor temperature = 80^{0}C

   Feed rate = 169000 kg / hr

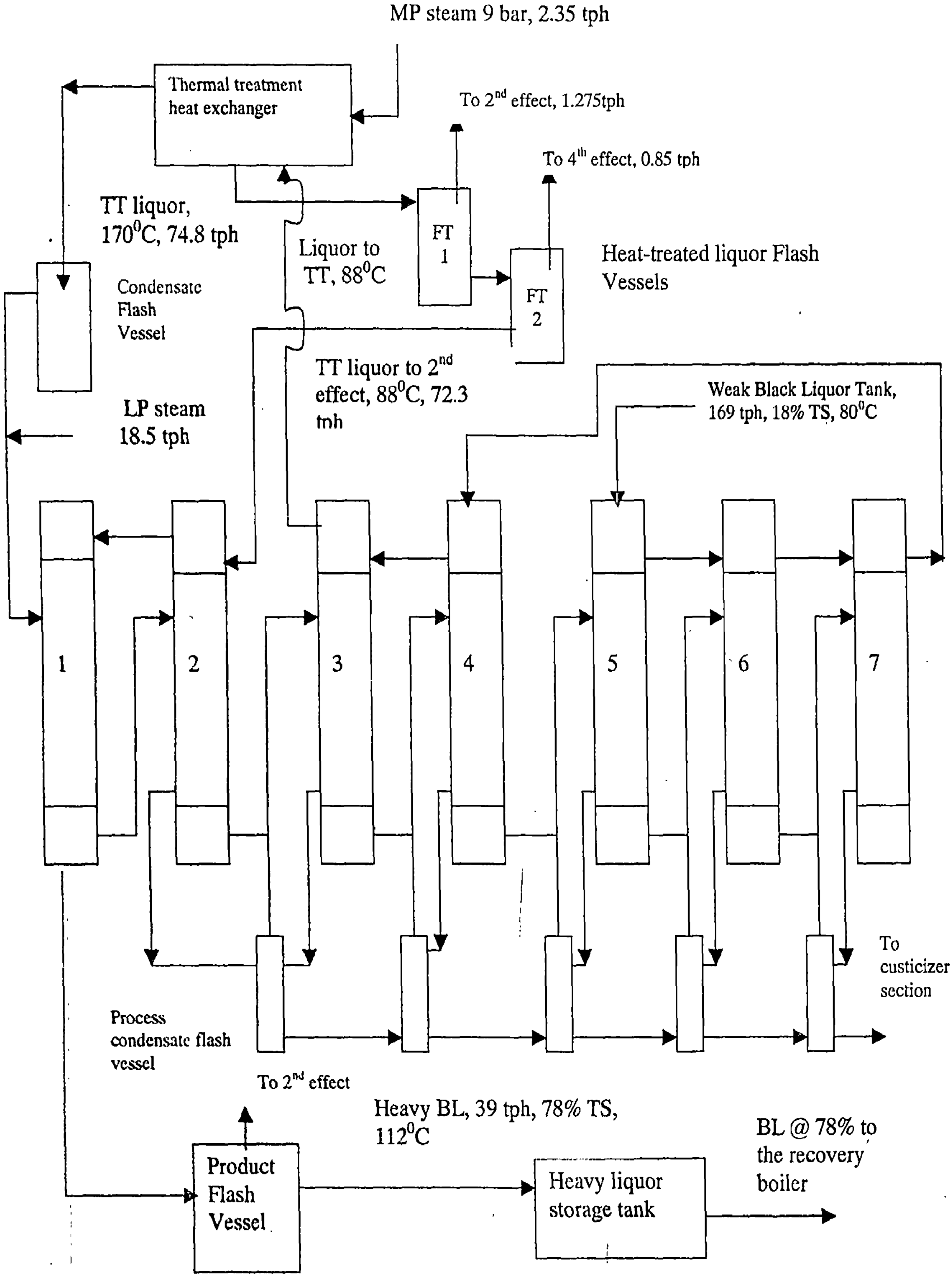
   Density = 1.08 tonne / m3

2. Fixing the final concentration level to the recovery furnace = 78% TS
3. Assuming the black liquor temperature to the recovery furnace as per design value = 112^{0}C.
4. Amount of water to evaporate from BL @ 78% = 130 tonne per hour
5. Amount of BL feed @78% to recovery boiler = 39 tonne per hour.
6. Thermal treatment temperature = 170^{0}C

Mass Balance:

Effect	1A/1B	2	3	4	5	6	7
Liquor temp, ^{0}C	125	102	88	79	71	64	54
BPR, ^{0}C	16	13	8	6	2.5	3	4
Vapor generation temp, ^{0}C	109	89	80	73	68.50	61	50
Calendria temp, ^{0}C	132	109	89	80	73	68.50	61
Driving force, ^{0}C	7.0	7.0	1.0	1.0	2	4.5	7.0
Water evaporated, tonne/hr	16.46	17.72	18.85	16.94	19.74	18.30	21.08
Feed Black Liquor, tonne/hr	55	72.32	93.70	109.37	169	149.26	130.96
Discharge Black liquor, tonne/hr	38.13	54.60	74.86	93.00	149.26	130.96	109.88

Liquor heat treatment plant: -



Energy balances: -

Energy balances in GJ per hr

effect	Heat in steam	Heat from PCFV	Heat from SCFV	Heat from TTFV	Heat from PLFV	Liquor cooling	**Heat input**	Heat in vapor out	Heating of liquor	**Heat output**
1	40.82	-	-	-	-	-	40.82	36.76	4.02	40.78
2	36.76	-	2.31	2.85	1.96	-	43.88	40.51	3.24	43.75
3	40.51	1.78	1.82	-	-	-	44.11	43.51	2.66	46.17
4	43.51	1.59	0.75	1.96	-	-	47.81	39.42	8.00	47.42
5	39.42	1.83	-	-	-	4.86	46.11	46.16	-	46.16
6	46.16	1.56	-	-	-	3.34	51.06	43.13	-	43.13
7	43.13	3.19	-	-	-	4.04	50.36	50.24	-	50.24

PCFV- Process Condensate Flash Vessel
SCFV – Steam Condensate Flash Vessel
TTFV – Thermal Treatment Flash Vessel
PLFV- Product Liquor Flash Vessel.

Energy requirement:

Live steam Consumption = LP steam for Evaporation + MP steam for heat treatment

$$= (18.83 - 0.274) + 2.35$$

$$= 20.9 \text{ tonnes per hour}$$

Steam Economy = 130 / 20.9 = 6.22

Estimation of Improvement in Steam Generation: -

- ✓ Present plant of 70% TS (design) steam generation = 90 tph.
- ✓ For 78% TS to the recovery boiler, increase in Steam generation = 93.82 tph

Energy Savings

LP steam required for evaporation = 18.5 tph

MP steam required for evaporation = 2.35 tph

Increase in steam generation in recovery boiler = 3.82 tph

Annual savings compared to present system @ Rs. 300 per tonne =3.82* 330*24*300.

≑ Rs. 91 lakhs.