OCCASION

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

DISCLAIMER

FAIR USE POLICY

CONTACT

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

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

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

1C

RESTRICTED

April, 1987

English

ENERGY MANAGEMENT IN THE INDIAN FERTILIZER INDUSTRY

DP/IND/85/006/11-01

TECHNICAL REPORT : EVALUATION OF ENERGY CONSUMPTION IN INDIAN FERTILISER PLANTS AND THE IDENTIFICATION OF AREAS OF INEFFICIENT ENERGY USE AND REMEDIAL MEASURES

Prepared for the Government of India
by the United Nations Development Organisation
acting as executing agency for the United Nations
Development Programme

Based on the work of Warwick J. Lywood
Expert in Energy Management

United Nations Industrial Development Organisation
Vienna

This report describes discussions held at Fertilizers and Chemicals of Travancore Limited (FACT) at Cochin and the Fertilizer Corporation of India Limited (FCI) at Ramagundam and Gorakhpur in February/March, 1987 regarding Energy Management and Conservation.

CONTENTS

SECTION A.

SUMMARY OF CONCLUSIONS AND RECOMMENDATIONS FOR EACH SITE

1. SUMMARY - FACT COCHIN DIVISION

Many of the efficiency losses on the site have been identified, but further work is necessary to characterise the ammonia plant in order to find other reasons for lost efficiency and to quantify the known reasons for losses.

The following actions and modifications are proposed as offering a significant potential for efficiency improvement on the site, and it is recommended that they should be investigated more fully.

Ammonia Plant

		Savings $x10^6$ rupees/yr.	Capital Cost $x10^6$ rupees
1.	Reline primary reformer)		
2.	Repair vetrocoke reboiler) leak. }	in ha.d	
3.	Install lean or semilean soln flash to reduce steam carbon ratio.	4.5	10
4.	Raise CO_2 regeneration pressure	N/E	
5.	Recommission vetrocoke let-down turbine	2.5	Low
6.	Stop make gas leaks and pipe seal losses to fuel	13.2	Low
7.	Improve efficiency of syn gas compressor	up to 30	N/E
8.	Modify steam control system	14.5	Low
9.	Pipe ammonia gas to NPK plant	16	N/E
10.	Install hydrogen recovery unit	3.5	8-10

Urea Plant	Savings $\times 10^6$ rupees/yr.	Capital Cost $\times 10^6$ rupees
Supply hydrolyser reboiler with low pressure steam	7	Low
Sulphuric Acid Plant		
Install motor on air blower and export 37 ata steam and install new turboalternator	23.5	N/E
Steam Power Plant		
Replace burners to reduce excess air	16	Low

TOTAL	131	

N/E - Not Estimated

A number of these modifications involve new interplant piping for which costs are not readily avilable.
There are also a number of minor efficiency improvements which are detailed under the respective plants.

1. SUMMARY - FACT UDYOGAMANDAL DIVISION

Most of the efficiency losses on the composite ammonia plant have been identified and proposals made to eliminate the losses and to reduce naphtha consumption further below flowsheet.

The following measures can be taken to improve the efficiency of the plant:-

Monitor excess air on primary reformer and trim furnace.

Check design of fluegas waste heat boiler and check amount of bypassing.

Renew seals on combustion air heater and restrict inlet temperature.

Reduce auxiliary firing on auxiliary burners and import LP steam for MEA.

Clean or replace naphtha cooler E-002.

Reduce steam ratio.

Monitor excess air on fired heaters and trim burners.

Increase size of process air heater coil in naphtha vaporiser.

Check mass balance on process stream and check for losses.

Monitor air compressor and synthesis gas compressor efficiencies.

Install water separator in top of LP scrubber.

Consider isolating top row of burners in Primary Reformer.

Retrofit MEA system to MEA - Amine guard and install lean solution flash.

Major schemes for revamping the 140 te/day partial oxidation ammonia plant have been considered. These are likely to be expensive, and it is recommended that an outline proposal with rough efficiency improvements and capital cost is provided by the retrofit vendor, so that an approximate evaluation can be done, before paying for a full feasibility study.

The proposal made by Technical Services to install a low pressure boiler in the synthesis loop of the partial oxidation ammonia plant and to rationalise the site steam system by installation of a turboalternator looks to be sound and should be evaluated further.

1 SUMMARY - FCI RAMAGUNDAM UNIT

The efficiency of the ammonia plant is very much worse than design on an annual basis. A large part of the inefficiency is due to operating at part rate with only one or two of the gasifiers working. The greatest single efficiency improvement could be obtained by improving the availability of gasification by installing a fourth gasifier.

Areas of plant and equipment where there are efficiency losses were discussed and a number of areas where action can be taken to improve efficiency are recommended. The major areas of possible improvement below, could give an efficiency improvement worth 8.1 crores rupees per annum.

	Saving 106 rupees/yr
Clean first compressor stage and intercoolers of air compressor	4
Raw gas compressor	9
Synthesis gas compressor	6
Return nitrogen wash tail gas to RGC suction	14
Return ammonia loop flash gas to SGC suction	3.5
Burn gas flared during start-up	N/E
Avoid running two nitrogen compressors at two gasifier rate	16
Eliminate continuous oil firing on SGP	15
Modify SGP import steam superheater	0.5
Control of steam system	12
Improve cooling system	N/E
Connect urea low pressure steam to ammonia plant system	1.2
TOTAL:	81
Install fourth gasifier	about 139

N/E = Not estimated

1. SUMMARY - FCI GORAKHPUR UNIT

The capacity utilisation and efficiency of the plants on the site is poor. Only part of the difference between the achieved and flowsheet efficiencies is accounted for on the ammonia plant. While some of the efficiency losses on the plants are due to operating at below design steam rates and due to breakdowns and start-ups, these reasons only account for about half of the total loss on the ammonia plants. There are major improvements to be made by better operation and maintenance and rectifying design errors. Better monitoring and complete loss accounting is needed to identify and quantify the reasons for these efficiency losses.

Areas of plant and equipment where there are known to be problems were discussed and a number of areas where action can be taken to improve efficiency are recommended. These are given in detail in the main text, but the major areas of improvements are summarised below :

Ammonia Plant

Increase plant operating rate to design rate
- install new nitrogen compressor
- uprate oxygen compressors

Clean or redesign refrigeration condensers

Stop throttling suction of syn gas compressors

Install liquid nitrogen storage for rapid start-up of Air Separation Units

Improve operation of cooling water system

Urea Plant	Saving 10^6 rupees/yr
Improve performance of ammonia recovery column)	
Improve design of ammonia absorber cooler)	58
Refit scrubbing system on urea prilling towers)	
Convert to modern urea process	up to 47
Steam Generation Plant	
Improve boiler thermal efficiency	up to 8.3
Operate two out of three boilers	1.9
Install/run steam driven BFW pump	2.7
Send hot urea condensate to SGP	N/E
Install site turbo alternator	52

SECTION B.

GENERAL CONCLUSIONS AND RECOMMENDATIONS

There are a number of actions that can be taken or techniques that can be used to identify and remedy efficiency losses which are applicable to a number of the sites visited. These are :-

1. Plant characterisation
2. Equipment performance monitoring
3. Control system design and dynamic simulation
4. Radioactive injection for flow measurement
5. Infrared photography
6. Methods of cleaning heat exchangers
7. Correct doping of cooling water systems
8. Piping blow-off gas to fuel

1. Plant Characterisation:

The procedure and results of a plant characterisation are shown in the appendix. While a full plant characterisation is most worthwhile on a steady running plant, the technique can be usefully applied to sections of any plant in order to determine reasons for efficiency losses in that section.

2. Equipment Performance Monitoring:

There was little equipment performance monitoring for efficiency in the sites visited. Where it was being done, it was not always clear which measurements should be recorded and how they should be interpreted. Particular areas where more monitoring should be done are: machine efficiencies, heat exchanger performance, cooling tower performance and catalyst performance.

3. Control System Design and Dynamic Simulation:

There were a number of areas where better control system design using a dynamic simulation could give significant efficiency improvements. Two sites had steam control systems where there were significant amounts of steam blown off due to poor control. The control system can be designed to avoid this. A dynamic simulation of the control system can be written to ensure that the control loops are tuned so that equipment does not trip or relief valves lift spuriously when there is an upset to the system.

4. Radioactive Injection for Flow Measurement:

A radioactive species is injected into a stream and monitored downstream. The method can be used for simple flow measurement where no facility currently exists, or can be used for detecting bypassing or leakage in heat exchangers, machines or catalytic convertors.

5. Infrared Photography (thermal imaging):

This is used to indicate hot spots on furnace walls or refractory lined header systems. The reason for the hot spot can be found and remedial action taken during a following shutdown.

6. Methods of Cleaning Heat Exchangers:

There were a number of instances where plant personnel were not sure of the best way of cleaning particular heat exchangers and expertise in this area would be useful.

7. Correct Doping of Cooling Water Systems:

The FACT plants had recently had a visit from a UNIDO sponsored water chemist to advise on water treatment problems. The FCI sites visited could benefit from similar advice particularly on treatment of cooling water systems.

8. Piping Blow-off Gas to Fuel:

In most of the ammonia plants there is scope for returning blown off process gas to fuel or low pressure feed. This gas is either from plant start-up blow-offs or small amounts of gas lost during continuous operation.

APPENDIX - PLANT CHARACTERISATION

A plant characterisation is the analysis of the data describing the performance of the plant. The following steps are needed :-

1. Collect necessary data to describe each section of the plant to to the required level.
2. Check data for consistency.
3. Evaluate the performance of each section of the plant.
4. Compare performance with design performance.
5. Collect extra data or check data or do tests to resolve data inconsistencies and to distinguish between possible causes of poor performance.
6. Put together model of the plant incorporating performance data. If the plant is complicated a model using a computer flowsheeting package is needed in order to enable prediction of the plant performance at different conditions and with modified equipment.

The following areas of the plant and their performance characteristics are normally evaluated relative to design:-

Area	Characteristics
1. Overall plant	- heat and mass balance - heat losses, material losses. - plant limits
2. Catalysts	- activity, pressure drop, bypassing.
3. Furnaces	- radiant efficiency, excess air, heat losses.
4. Process heat exchangers	- fouling, pressure drop, bypassing.
5. Machines	- efficiency, fouling, bypassing, intercooler performance, catchpot carry-over, condensing pressure.
6. Cooling system	- approach to wet bulb temperature, circulation rate.
7. CO_2 removal systems	- specific heat and power requirements.

SECTION C.

EVALUATION OF ENERGY CONSUMPTION AND IDENTIFICATION OF AREAS OF INEFFICIENT ENERGY USE AND REMEDIAL MEASURES

FERTILIZERS AND CHEMICALS OF TRAVANCORE LTD
COCHIN DIVISION

CONTENTS

1. SUMMARY - FACT COCHIN DIVISION

Many of the efficiency losses on the site have been identified, but further work is necessary to characterise the ammonia plant in order to find other reasons for lost efficiency and to quantify the known reasons for losses.

The following actions and modifications are proposed as offering a significant potential for efficiency improvement on the site, and it is recommended that they should be investigated more fully.

Ammonia Plant

		Savings $x10^6$ rupees/yr.	Capital Cost $x10^6$ rupees
1.	Reline primary reformer)		
2.	Repair vetrocoke reboiler) leak.)	in hand	
3.	Install lean or semilean soln flash to reduce steam carbon ratio.	4.5	10
4.	Raise CO_2 regeneration pressure	N/E	
5.	Recommission vetrocoke let-down turbine	2.5	Low
6.	Stop make gas leaks and pipe seal losses to fuel	13.2	Low
7.	Improve efficiency of syn gas compressor	up to 30	N/E
8.	Modify steam control system	14.5	Low
9.	Pipe ammonia gas to NPK plant	16	N/E
10.	Install hydrogen recovery unit	3.5	8-10

	Savings $x10^6$ rupees/yr.	Capital Cost $x10^6$ rupees
Urea Plant		
Supply hydrolyser reboiler with low pressure steam	7	Low
Sulphuric Acid Plant		
Install motor on air blower and export 37 ata steam and install new turboalternator	23.5	N/E
Steam Power Plant		
Replace burners to reduce excess air	16	Low
TOTAL	131	

N/E - Not Estimated

A number of these modifications involve new interplant piping for which costs are not readily avilable.

There are also a number of minor efficiency improvements which are detailed under the respective plants.

2. INTRODUCTION

The Cochin Division of FACT has the following Fertilizer Plants:

	Capacity	Commissioned
Ammonia	600 tpd	1973
Urea	1000 tpd	1973
Sulphuric Acid	1000 tpd	1976
Phosphoric Acid	360 tpd	1976
NPK Plant	1600 tpd	1976

The following work was done to improve the efficiencies of these plants and of the site in general.

1. Identify for each plant the efficiency losses compared to flowsheet.
2. Identify changes that could be made to plant flowsheets in order to improve the efficiency.
3. Identify where the integration of the site steam and ammonia systems could lead to a saving in site energy usage.

In order to assess the benefits of efficiency improvements the costs used were :

Naphtha	2084 rupees/te	= 0.183 rupee/cal HHV
LSHS	2000 rupees/te	
Power import	660 rupee/MWh (average)	
54 ata steam from boilers	189 rupees/te	
13 ata superheated steam	141 rupees/te	

3. AMMONIA PLANT

The 600 te/day Ammonia Plant is based on naphtha reforming and was commissioned in 1973. The front end was designed by FEDO in collaboration with Davy Power Gas Corporation, while the ammonia synthesis section was designed by FCI based on the Fauser Montecatini Process in collaboration with Technimont. The plant has operated up to 90% rate on an air compressor limit but is currently operating at 70% rate due to the poor condition of the reformer. The plant efficiency is significantly worse than design.

	Flowsheet	Ideal Design	Current Operating
Naphtha to process G cal/ te (HHV)	6.45	6.12	6.32
Naphtha to Fuel G cal/ te (HHV)	2.29	2.17	3.44
Total Naphtha G cal/te	8.74	8.28	9.76
Power import kWH/te	429	407	431
3o ata steam import te/te	0	0	0.70
Variable cost rupee/te	1882	1784	2203

The original flowsheets by Davy and Technimont each allowed for a 2.5% loss of make up gas around the compressor giving a total loss of 5%. This is excessive. The 'ideal design' figures assume no loss of make up gas.

The possible energy saving improvements can be divided into two types:-

3.1 Improvements to get back to flowsheet efficiency.

3.2 Improvements to the flowsheet.

3.1 Improvements to get back to Flowsheet Efficiency.

The necessary data is not available to enable a complete breakdown of the efficiency losses compared to flowsheet and a plant characterisation is needed to establish these data. The data needed from a plant characterisation and the procedure needed to establish it are given in the appendix of Section B. It is recommended that a plant characterisation of the ammonia plant be done after the next overhaul when the reformer has been relined and the plant is closer to design rates.

The following efficiency losses have been identified compared to the ideal design figure.

	Loss
Low plant rate	N/E
High primary reformer heat loss	N/E
Poor reformer firing distribution	N/E
High Steam/Carbon ratio	0.13 G.cal/te
Inefficient process air compressor	N/E
Vetrocoke reboiler leak	0.044 G.cal/te
Other make gas stream leaks	0.40 G.cal/te
Vetrocoke hydraulic turbine not running	19 KWH/te
Low front end pressure	N/E
Inefficient synthesis gas compressor turbine	up to 0.93 G.cal/te
Steam system control	0.4 te/te

There are other major heat losses to account for the very poor fuel efficiency compared to design. More data is needed to identify these and to quantify some of the known losses.

3.1.1 Low Plant Rate

The electricity usage will hardly change as the rate is reduced, so the electrical efficiency will be worse at low rates. The efficiency of turbines and compressors should not change significantly until the plant rate falls below 70-80%. The efficiency of heaters and the reformer will improve as the plant rate is reduced.

3.1.2 Reformer Heat Loss

The refractory in the primary reformer is in a very poor condition causing high convective and radiant heat losses from the furnace walls. The reformer is being relined in the next overhaul. Areas of poor internal insulation can be identified by using infrared photography of the outside wall of the furnace. If this is done before subsequent overhauls, the refractory can be repaired in the shutdown before there is buckling of the wall panels and hence further deterioration of refractory.

3.1.3 Reformer Firing Distribution

From looking into the reformer box, it was evident that there is poor distribution of firing and a large variation in flame lengths. This is giving a poor radiant efficiency in the furnace and causing some tubes to run hot. Some means should be provided for adjusting the firing distribution. A range of exit pigtail temperatures should be measured in order to identify the hot and cold areas of the furnace.

3.1.4 High Steam:Carbon Ratio

The plant is running at a steam :carbon ratio of 4:0 instead of the design of 3:4. The main reason for this is in order to provide enough heat for the vetrocoke reboiler - see later. However, there is also concern about the accuracy of measurement of the steam:carbon ratio, so the plant is run at a higher ratio to give a safety margin. If this is a problem then the steam and naphtha flows can be automatically corrected for temperature and pressure and duplicate measurements can be taken to ensure accurate measurement.

The cost of running at a steam ratio of 4:0 instead of 3:4 is about 0.13 G.cal/te or 4.5×10^6 rupees/year.

3.1.5 Process Air Compressor/Turbine

The output of the process air compressor and the turbine steam consumption are worse than design. The delivery pressure from the first stage of the machine is low, but the efficiency of the second stage is slightly better than design. There is no leakage on the antisurge bypass and the air flow measurement is taken before the blow-off. The pressure exit the 1st stage is measured after the intercooler.

The following actions should be taken to find the cause of the poor performance or to rectify it.

1. Measure temperatures and pressures inlet and exit of the steam turbine and calculate the compressor power and steam turbine efficiency compared to design.
2. Measure the pressure drop across the interstage cooler to check that it is not high.

3. Measure the water knocked out from the intercooler catchpot and check that it corresponds to the correct amount for the intercooler exit temperature and pressure and the humidity of the air. Intercooler separators often do not perform satisfactorily.
4. Calculate cost and benefit of a larger intercooler in order to give an approach between inlet cooling water and exit air of 3^{0} C or 5^{0} C compared to the present 11^{0} C.

If the rate from the process air compressor cannot be improved sufficiently to meet design, then a reciprocating compressor in parallel with the existing machine or a suction booster for the compressor should be evaluated.

There is a proposal to replace the current air compressor with a new one because the existing compressor casing was damaged during a breakdown and spares are not available. If a compressor replacement is approved or if a parallel compressor is installed then a margin should be added so that plant output can be increased if a hydrogen recovery unit is installed. (See Section 3.2.7)

3.1 6 Vetrocoke Hydraulic let-down Turbine:

A hydraulic turbine was installed but was never run. There should be no problem with operating a let-down turbine when the vetrocoke pumps are motor driven. Either the original turbine should be recommissioned or a new turbine installed. The cost saving from operating the let-down turbine will be 19 KWH/te or 2.5×10^{6} rupees/year.

3.1.7 Vetrocoke Steam Reboiler

The vetrocoke steam reboiler is in the bottom of the regenerator tower. This is unusual and careful design is needed for an internal reboiler to perform satisfactorily. Above a certain steam rate, vibration occurs in the regenerator and it is reported that the performance of the gas heated reboiler gets worse. For this reason, the steam rate to the reboiler is kept below design. The steam ratio is increased to provide enough heat for reboil and excess LP steam is vented.

With an internal reboiler there is a two phase mixture in the base of the regenerator, so the current level measurements in the base of the tower - sight glass and DP cell will indicate a lower level than that is actually attained. The high level could be causing flooding of the internal baffle or packing causing vibration problems. The true level in the base of the tower should be measured by a level switch on the vessel or by a Dp cell with a bottom tapping which is just below the correct operating level. The two phase mixture could also be causing vapour entrainment in the feed line to the gas heated reboiler giving poor performance of the gas heated reboiler. Vapour entrainment can be checked for by using a radioactive technique to measure the density in the line to the gas heated reboiler.

The problems caused by the internal steam reboiler will be eliminated if a lean solution flash modification is made to the vetrocoke system. (See Section 3.2.3)

3.1.8 High Hydrogen Level in CO_2

This is probably being caused by a leak in the gas heated reboiler. A leak test should be performed and leaking tube or tubes plugged in the next overhaul. The other possible cause of the high hydrogen level is due to gas entrainment in the bottom of the absorber. This can be remedied by raising the level in the absorber base.

3.1.9 Other make Gas Stream Leaks:

A mass balance between feed naphtha and ammonia product from current plant data shows that there is a 5% loss of make gas. The vetrocoke reboiler leak accounts for only 1.0% of this loss. The rest of the losses are either through make gas blow-off control valves, relief valves or compressor seal losses.

The following actions should be taken :

1. Check the mass balance from process feed to ammonia product.
2. Measure the amount of make gas which is going up vents. This can be done by injecting a known amount of nitrogen into the base of a vent and taking a gas analysis further up the vent.
3. If the loss is due to blow off control valve leaks, change to soft seat control valves.
4. If the loss is from compressor seal leaks, pipe the blow off gas to fuel through a cooler and glass wool oil separator.

3.1.10 Low front end Pressure:

The front end pressure is normally run at lower than design pressure because the plant is on an air compressor limit.

At the moment, the reformer instead of the air compressor is limiting so the front end pressure should be run as high as possible in order to reduce the steam consumption on the synthesis gas compressor turbine.

3.1.11 Synthesis Gas Compressor/Turbine:

The synthesis gas compressor turbine consumes 25% more 130 ata steam than it should do at 80% plant output. The cost of the inefficiency could be up to 0.93 G.cal/te or 30 x 10^6 rupees/year. A characterisation is needed to establish the cause of the high steam rate. Possible reasons are poor compressor stage efficiency, poor turbine efficiency, higher proportion of steam extraction at 30 ata than design, poor vacuum on condenser, lower than design compressor suction pressure or high intercooler pressure drop.

If the cause of lost efficiency is due to a poor steam turbine efficiency, then replacement of the turbine rotor with a more efficient design should be evaluated.

3.1.12 Steam System Control:

At 90% ammonia plant rate about 10-15 te/hr of 54 ata steam is imported from the boiler plant and let down to 30 ata, 13 te/hr of steam is let down from 30 ata to 3.4 ata and 10 te/hr of 3.4 ata steam is vented. This gives a plant efficiency loss of 0.42 G.cal/te and is worth 14.5 x 10^6 rupees/year.

The steam control system should be modified so that either no steam is let down from 30 ata to 3.4 ata or so that no 3.4 ata steam is vented.

3.2 Improvements to the Flowsheet:

The following improvements to the flowsheet have been evaluated:-

Install selectoxo unit.

Replace vetrocoke with alternative solution

Install lean and/or semi-lean solution flash in vetrocoke.

Raise CO_2 regenerator pressure and bypass CO_2 gas holder.

Install parallel ammonia converter.

Install molecular sieve driers.

Replace ammonia converter cartridge with low pressure drop design.

Install hydrogen recovery.

Export ammonia as gas to NPK and shutdown refrigeration compressor during normal operation.

3.2.1 Install Selectoxo Unit:

This is only worthwhile when the plant is not on a primary reformer limit and not on a synthesis gas compressor limit. The capital cost is high and proven experience is not good. It should not be considered further for this plant.

3.2.2 Replace Vetrocoke with Alternative Solution:

The only replacement solvent for vetrocoke that has been successful is Benfield. An increased mass transfer area is required with Benfield than Vetrocoke, but this can often be achieved by better tower packings and distributors. Pipework and heat exchanger bundles need to be changed to stainless steel. Parts of the towers need to be stainless steel lined and the other metal surfaces need to be sandblasted or passivated. Union Carbide Corporation have done a number of conversions from Vetrocoke to Benfield and are the experts in this field. Replacement of vetrocoke solution should not be pursued further unless there is strong pressure due to safety considerations.

3.2.3 Install lean and/or semi-lean solution flash in Vetrocoke

The design reboiler load on this vetrocoke system is 95 KJ/Kmole. Values of 76 KJ/Kmole can be obtained with lean solution flash using an ejector to recompress flash steam. Further reductions can be achieved using mechanical recompression of flash in addition to steam ejectors. A solution flash system can be designed in order to provide all the reboil heat needed using only the gas heated reboiler at the design steam:carbon ratio.

The amount of steam used on the ejector can be adjusted in order to maintain a plant LP steam balance.

Lean solution flash retrofits have proved successful in operation. The design is best done by Union Carbide Corporation.

3.2.4 Raise CO_2 Regeneration Pressure

If the operating pressure of the regenerator is raised then the higher CO_2 pressure will enable a reduction in power on the CO_2 compressor for urea. The CO_2 gas holder would need to be bypassed. The lean solution flash system can be designed to give an LP steam balance at the higher regeneration pressure. However, a higher regeneration pressure will give higher temperatures of the vetrocoke solution. This may give increased corrosion rates, and the operating conditions should be checked with UCC.

3.2.5 Install Moiecular Sieve Driers

Molecular sieve driers are used to dry the make up gas to the synthesis loop, so that the make up gas can be added to recycle gas after the loop catchpots instead of before the catchpots. The benefits from synthesis gas driers are small in a high pressure synthesis loop and they will give added complication. The size will be large unless the make up gas is chilled before entering the driers. Synthesis gas driers are unlikely to be an economic retrofit on this plant and should not be pursued.

3.2.6 Install Parallel Ammonia Converter or Replace Converter Cartridge

These modifications are used to reduce the pressure drop in the synthesis loop to enable a higher loop hydrogen efficiency or a lower loop operating pressure. These are unlikely to be economic in the Montecatini synthesis loop because the loop interchanger and boiler are inside to converter shell so it would be difficult and expensive to change the existing integrated arrangement.

3.2.7 Install Hydrogen Recovery

A hydrogen recovery Unit will normally enable the synthesis loop hydrogen efficiency to be increased to 98 or 99% compared to the design loop hydrogen efficiency of 93.7% for this plant. Because some purge gas is used for hydrofining, the loop efficiency on this plant can not be increased above 98%. At the purge rate on this plant a membrane type hydrogen recovery unit will be cheaper than a cryogenic or PSA system. The most widespread membrane units for hydrogen recovery are the Monsanto Prism, but other companies have membrane systems and should be approached to ensure competitive quotes. As long as the process air compressor is not limiting, the recovered hydrogen from the loop can be used to increase the plant output as well as improving efficiency. The hydrogen recovery unit can be sized to take a higher purge rate than that currently achieved in order to lower the synthesis loop pressure and save steam on the synthesis gas compressor turbine. The optimum size of the hydrogen recovery unit can be determined by running a computer simulation of the plant for different cases. The saving from a hydrogen recovery unit will be about 0.1 G.cal/te or 3.5×10^6 rupees/year, while the cost will be about $8\text{-}10 \times 10^6$ rupees. An increase in plant output should also be possible.

3.2.8 Export Ammonia as Gas to NPK Plant

The ammonia refrigeration compressor unit liquifies 7.25 te/hr ammonia at 2.7 ata and 8.7 te/hr ammonia at 4.5 ata. The service refrigeration compressor unit liquifies 1.35 te/hr ammonia at 4.5 ata. The NPK plant imports 20 te/hr liquid ammonia and vaporises it. It should be possible to install a 2.7 ata ammonia gas main to take ammonia vapour from the ammonia plant and Hortensphere direct to the NPK plant. Some modification may be needed on NPK plant to use low pressure ammonia and a variable speed motor may be needed on the ammonia refrigeration compressor, so that it can be run more efficiently at low rates when the NPK plant is not at full rate. The saving when the NPK plant is above 86% rate will be 123 KWH/te NH3 or 16 x 10^6 rupees/year at average electricity cost.

3.3 Conclusion:

The following potential efficiency improvements have been identified which will give a significant improvement to the efficiency of the ammonia plant. (P.T.O)

	Savings efficiency	Savings 10^6 rupees/ year	Capital Cos 10^6rupees
1. Reline primary reformer		in hand	
2. Repair vetrocoke reboiler leak		in hand	
3. Install lean and or semi-lean solution flash to reduce steam : carbon ratio	0.13 G.cal/te	4.5	8-10
4. Raise CO_2 regeneration pressure	N/E		
5. Recommission vetrocoke let down turbine	18 KWH/te	2.5	Low
6. Stop make gas leaks and pipe seal losses to fuel	0.37 G.cal/te	13.2	Low
7. Improve efficiency of synthesis gas compressor	up to 0.93 G.cal/te	up to 30	N/E
8. Modify steam control system	0.42 G.cal/te	14.5	Low
9. Pipe ammonia gas to NPK plant	123 KWH/te	16	N/E
10. Install hydrogen recovery unit	0.1 G.cal/te	3.5	8-10
Total	up to 1.95 G.cal/te + 142 KWH/te	up to 84	

A plant characterisation is needed to identify the reasons for other losses compared to flowsheet.

4. UREA PLANT

The Urea Plant was commissioned in 1973 and is based on the Montecatini total liquid recycle process in collaboration with M/s. Technimont of Italy. Most of the plant is in two steams each with a capacity of 500 500 tonnes per day. The plant will run at design rate but the measured efficiencies are worse than design.

	Design	Operating	Extra cost per te Urea rupee
Ammonia te/te urea	0.59	0.61	44
13 bar steam te/te	1.8	1.91	13
54 bar steam te/te	0	0.29	55
Power KWh/te	180	189	6

The following losses compared to design have been identified:

4.1 Ammonia Efficiency

4.1.1 Ammonia loss in inerts purge.

The ammonia lost in the inerts purge is 3.5 te/day instead of the design of 1.2 te/day for each steam. This amounts to a loss compared to design of 0.005 te/te. Reasons for the extra loss are :-

1. The design inerts level of CO_2 to the Urea Plant on the Urea Plant flowsheet is 0.7% whereas the design inerts level in CO_2 from the Ammonia Plant flowsheet is 1.37%.
2. There is a leak on the vetrocoke reboiler which gives a higher than design inerts level in the CO_2 of about 2.8%.

3. The recycle water scrubbing steam on the inerts washing column is not operated, so the proportion of ammonia in the inerts purge is higher than design. The recycle is not operated, in order that the gas mixture in the column is further away from the explosive limit. However, there is a large margin between the gas composition and the explosive limit even when operating the water recycle. So the operation of the recycle should be reconsidered.

4.1.2 Dump from Ammonia Absorber

Solution in the ammonia absorber is frequently dumped because of a transient high level in the absorber. The loss of ammonia is about 3.2 te per day per steam which is equivalent to 0.006 te/te of urea. The reason for the dumping of the solution is due to a control problem on the ammonia absorber or on the ammonia rectification column. The root cause of the problem should be determined and the control system modified to correct it.

4.1.3 Measurement Accuracy/Urea Losses

Since the hydrolyser has been installed to remove urea, ammonia and CO_2 from the condensate and recycle it to the process, there should not be any other reason for significant ammonia losses. The remaining discrepancy between measured operating and design ammonia efficiency is only 1.7% and could easily be due to errors in measurement of the ammonia feed and urea product, or to unaccounted urea losses.

4.2 Steam Efficiency

4.2.1 Hydrolyser

The main reason for extra steam usage 0.29 te/te is the 12 te/hr of 54 ata steam which is used on the hydrolyser, which was installed after the original plant. It may be possible to reduce this steam rate, but this was not examined. However, it should certainly be possible to use low pressure steam for the hydrolyser reboilers (10 tes/hr). This should be considered as part of the rationalisation of the site steam system - see separate section.

4.2.2 13 ata Steam Efficiency

The urea reactor is run at a pressure of 195 atas instead of the design operating pressure of 221 atas. This causes the concentration of urea exit the reactor to be 29.8% instead of the design of 30.7%. The temperature of the urea solution to the distillers is below design. These two differences from design running will cause higher than design recycle rates leading to a loss of efficiency. The steam rate to each distiller cannot be measured and the flowmeter for reading the solution recycle rate is not working. The following actions should be taken:-

Raise the urea reactor pressure to the design value.

Measure the steam flow to the different steam users using a radioactive injection technique.

Determine the heat exchange performance of the distiller compared to design.

Recommission the solution recycle flowmeter.

4.3 Flowsheet Modifications

It has been proposed that the urea plant be retrofitted to use the Montedison isobaric double recycle (IDR) process in order to reduce the ammonia and steam consumptions to 0.57 te/te and 0.6 te/te respectively. The 0.57 te/te of ammonia is an ideal figure and in practice the improvement to the ammonia efficiency that could be obtained by using a new process would be minimal. All modern urea processes promise a reduction in steam usage to 0.6 to 1.05 te/te. However, in most of these processes including IDR the steam pressure required is about 25 ata. This would not fit into the existing site steam balance, so the benefits would be small or non-existant. An exception is the Mitsui-Toatsu (MTC) Recycle D Process with a steam usage of 0.78 te/te at a pressure of 14 bar, which is similar to the current steam pressure. The cost of a retrofit to the MTC Recycle D Process is unknown, and the existing plant may be incompatible with MTC processes. Alternatively there may be retrofits that can be done by Montedison to improve the existing plant without having to change the steam pressure. While not reducing the efficiency as far as the IDR process, they may be better value for money.

5. SULPHURIC ACID PLANT

The Sulphuric Acid Plant was commissioned in 1976 and has a capacity of 1000 tonnes per day. The design was by FEDO in collaboration with Davy Power Gas based on double conversion double absorption process under licence from Bayer.
The plant runs at the design rate and generates the design steam rate. However, the steam raised is not used efficiently.
47 te/hr of steam is raised at 37 ata. 25 te/hr of this steam

is let down through turbines to 3.5 ata, 8.5 te/hr of steam is let down through valves to 3.5 ata and the remaining 13.5 te/hr is let down through a desuperheater to give 15.5 te/hr of 8 ata steam. Of the 8 ata steam 2.5 te/hr is used in the melting pits, 5 te/hr is exported at 8 ata to the NPK and Phosphoric Acid Plant, while the remaing 7 te/hr is let down to 5.5 ata and exported to the NPK plant. Of the 3.5 ata steam, 6 te/hr is used for deareation and the remaining 27.5 te/hr. is vented. It was intended that the spare 3.5 ata steam should be exported to the Phosphoric Acid Plant concentrators. However, the concentrators on the Phosphoric Acid Plant do not need to be run. There are three approaches to improving the efficiency of the steam system:-

1. Install a turboalternator to take in excess 37 ata steam and 3.5 ata steam, pass out steam at 8 ata and condense the remainder. A turboalternator for this duty would be complicated and expensive and efficiency would be lost when the steam rates were not at design valves.

2. Install a motor on the air blower and export the excess 37 ata steam to the rest of the site. The export steam could be used to supply any 30 ata steam import to the Ammonia Plant and to supply some of the 13 bar steam to urea. This will enable more power to be generated on the existing site turboalternator with the current firing on the site boiler.

3. Install a motor on the air blower, export to site all the 37 ata steam except that required on the BFW pump turbine and install a turboalternator to let the steam down to 6.5-8 ata. This is evaluated in the section on the site steam system. Work is needed to cost the export steam main, turboalternator and motor.

6. STEAM/POWER PLANT

The Power Plant consists of 3 x 60 te/hr capacity boilers and a 14 MW capacity turboalternator. The boilers were designed for furnace oil feedstock but now run on low sulphur heavy stock (LSHS). It was intended that only 2/0/0/3 boilers should be run at any one time, but higher than design steam demand from plants on the site has meant that all three boilers must be run for most of the time. The boilers raise about 122 te/hr of steam at 54 atas and 460°C. 27 te/hr is exported while the rest is passed to the turboalternator. 80 tes/hr is passed out at 14 atas and 300°C to the Urea Plant and the remainder is condensed in a condensing turbine.

6.1 Boilers

The boilers run with 90% excess air inlet the combustion air heater and a fluegas temperature of 200°C compared to a design of 175°C. The LHV efficiency is 86% compared to a design of about 94%. There is no measurement of the excess air exit the radiant chamber or in the stack, so it is not possible to be sure whether there is air leakage or whether the excess air is at the burners. It is unlikely that air leakage could account for more than 10% excess air, so most of the excess air will be at the burners. It is reported that if burner air is reduced then the stack becomes smoky. This is probably because the burners are not correctly designed for burning LSHS.

It is recommended that the following actions should be taken:

1. Organise a test on a burner vendors test furnace using LSHS with a view to replacing the burners on the boilers. The following burner vendors are known to have suitable test facilities :

   John Zinc

   Airoil Flaregas

   Babcock and Wilcox

   Laidlaw Drew

2. Measure the oxygen concentration at points exit the radiant chambers using a water cooled sample probe, in order to check that the combustion air is adequately distributed to the burners and to find out how much excess air is due to leakage in the duct.

3. Measure the oxygen concentration to the stack in order to establish whether there is a leak on the combustion air heater.

   The higher than design fluegas stack temperature can probably be accounted for entirely by the high excess air, but other possible contributing factors could be fouling of the tubes or bypassing between tubes and the boiler walls.

   The saving to be gained from reducing ezxcess air in in order to return to design efficiency is 16×10^6 rupees per year on a steam raising rate of 122 te/hr.

6.2 Site Steam System

The site steam system is inefficient, with a lot of steam being let down across control valves and steam being blown off. The steam system should be rationalised in order to reduce the losses. There are a number of ways of doing it and the best option will depend on whether the consumption of steam on the Ammonia Plant steam turbines remains the same or is reduced.

One method of rationalising the steam system is proposed assuming that the Ammonia Plant machine steam usage remains as it is now. The following changes would be made to the steam system:

1. Install a site steam main at 6.5 - 8 ata.
2. Convert air blower on the Sulphuric Acid Plant to a motor drive.
3. Install a turboalternator letting down Sulphuric Acid Plant export steam to 6.5 - 8 ata.
4. Supply all low pressure steam users including the hydrolyser reboiler from the new main.
5. Modify the steam control system on the ammonia plant so that it does not import steam and blow it off.

   The pressure of the new main will need to be optimised to get as much power from the new turboalternator while still enabling heating duties to be done on phos.acid and sulphuric acid plants. e.g. the pressure can be below 8 ata if the heating coils in the sulphur melting pits are made larger.

   The changes will mean that it is only necessary to run two of the three boilers on the boiler plant, so a fourth boiler will not have to be installed.

   The steam balance for the 'as now' and with the new system are shown overleaf.

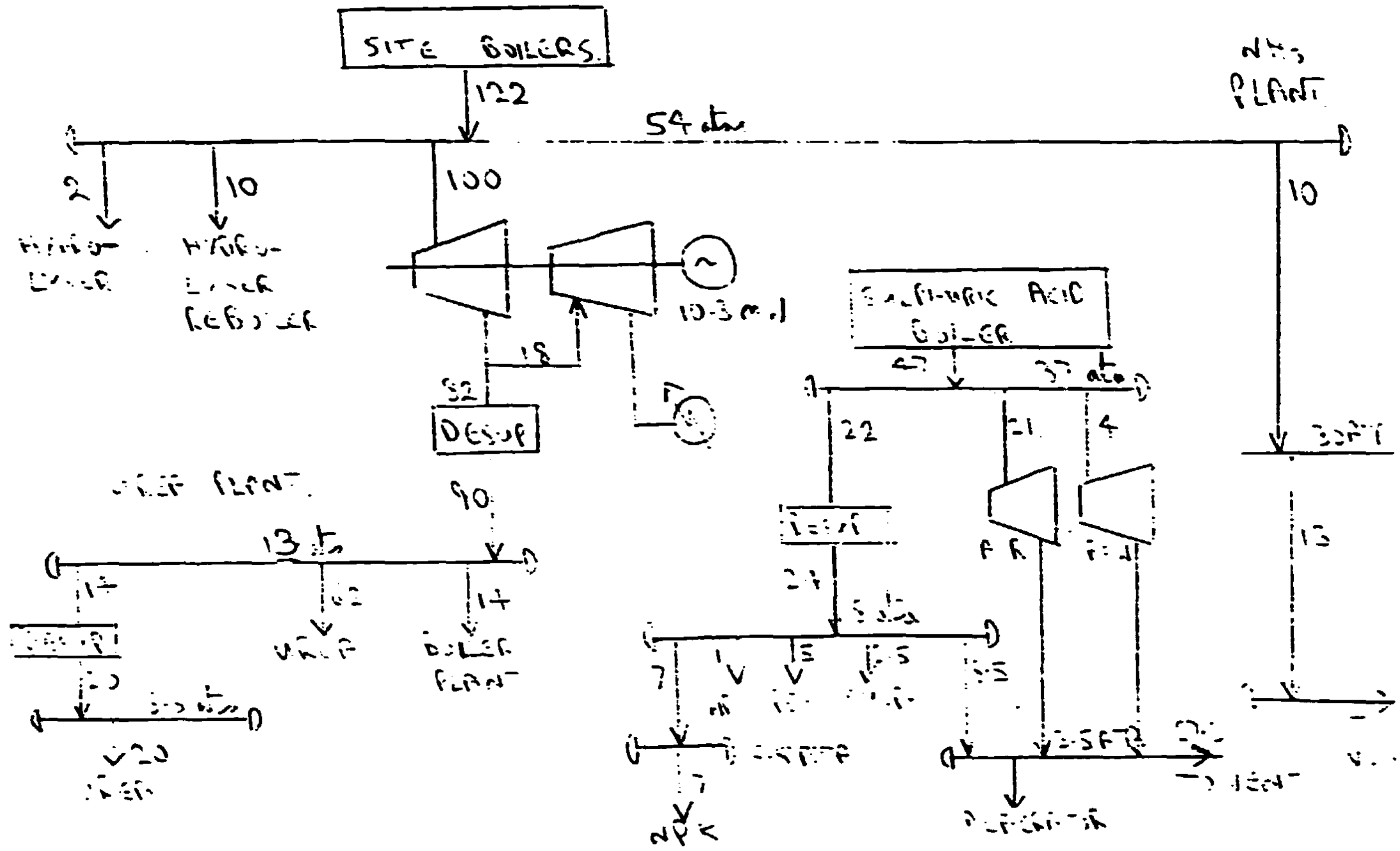
The economic benefit of the modified system is :-

Power from new turboalternator	3.1 MW
Power for urea air blower	2.0 MW
Extra power from existing turboalternator	0
Net gain in power generation	1.1 MW
Reduction in LSHS firing on boilers	1.83 te/hr
Total variable cost saving	5640 rupee/hr
	$= 45 \times 10^6$ rupee/yr

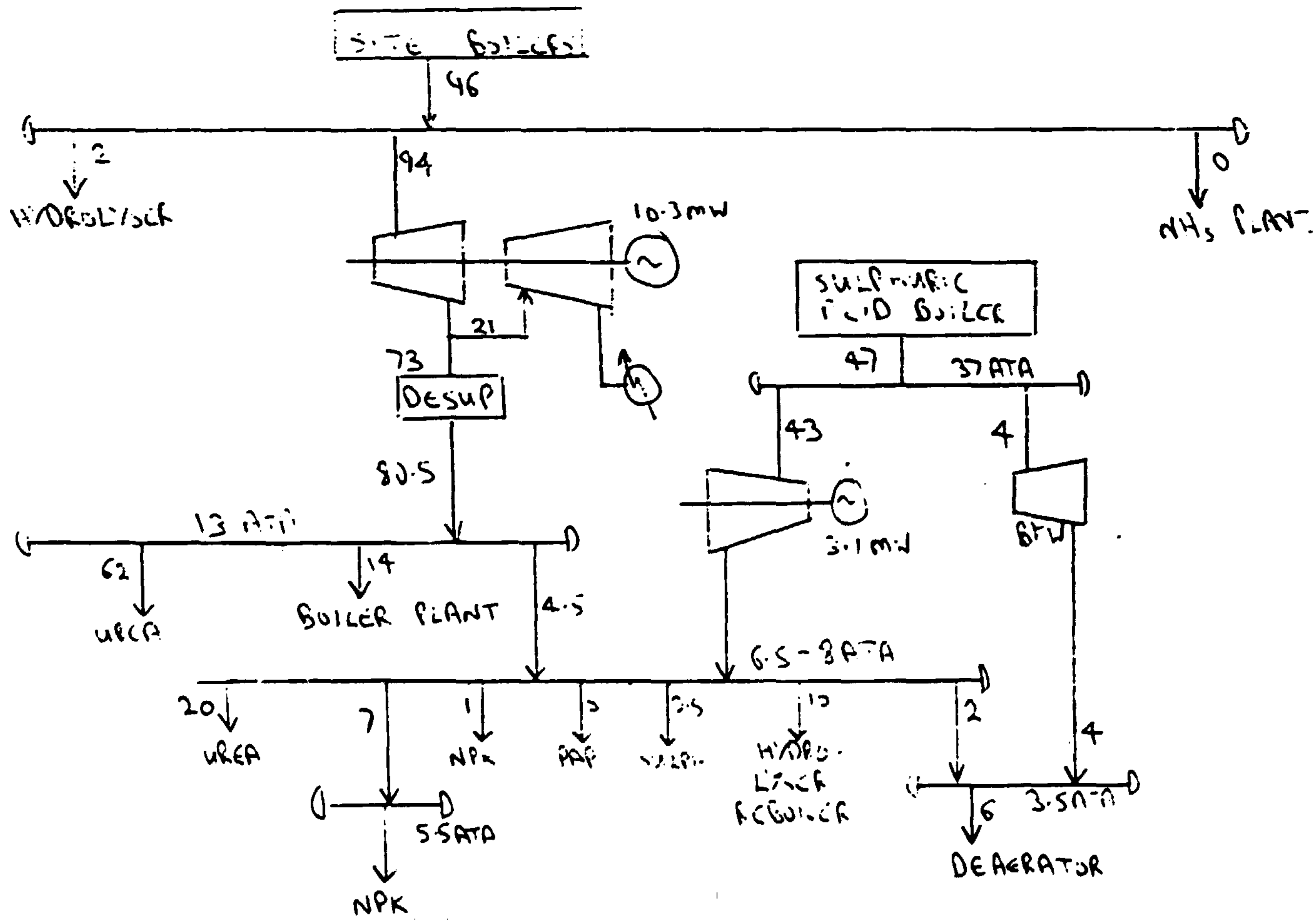
COCHIN - SITE STEAM SYSTEM

SIMPLIFIED SITE STEAM BALANCES

1) CURRENT SYSTEM



2) POSSIBLE NEW SYSTEM



SECTION D

EVALUATION OF ENERGY CONSUMPTION AND IDENTIFICATION OF AREAS OF INEFFICIENT ENERGY USE AND REMEDIAL MEASURES

FERTILIZERS AND CHEMICALS OF TRAVANCORE LTD

UDYOGAMANDAL DIVISION

CONTENTS

1. SUMMARY - FACT UDYOGAMANDAL DIVISION

Most of the efficiency losses on the composite ammonia plant have been identified and proposals made to eliminate the losses and to reduce naphtha consumption further below flowsheet.
The following measures can be taken to improve the efficiency of the plant:-
Monitor excess air on primary reformer and trim furnace.
Check design of fluegas waste heat boiler and check amount of bypassing.
Renew seals on combustion air heater and restrict inlet temperature.
Reduce auxiliary firing on auxiliary burners and import LP steam for MEA.
Clean or replace naphtha cooler E-002.
Reduce steam ratio.
Monitor excess air on fired heaters and trim burners.
Increase size of process air heater coil in naphtha vaporiser.
Check mass balance on process stream and check for losses.
Monitor air compressor and synthesis gas compressor efficiencies.
Install water separator in top of LP scrubber.
Consider isolating top row of burners in Primary Reformer.
Retrofit MEA system to MEA - Amine guard and install lean solution flash.

Major schemes for revamping the 140 te/day partial oxidation ammonia plant have been considered. These are likely to be expensive, and it is recommended that an outline proposal with rough efficiency improvements and capital cost is provided by the retrofit vendor, so that an approximate evaluation can be done, before paying for a full feasibility study.
The proposal made by Technical Services to install a low

pressure boiler in the synthesis loop of the partial oxidation ammonia plant and to rationalise the site steam system by installation of a turboalternator looks to be sound and should be evaluated further.

2. INTRODUCTION

The Udyogamandal Division of FACT has the following Fertilizer Plants:-

	Capacity	Commissioned
Ammonia - Naphtha partial oxidation	80 tpd	1962
Ammonia - Naphtha partial oxidation	140 tpd	1965
Ammonia - Naphtha reforming	120 tpd	1971
Sulphuric acid	160 tpd	
Sulphuric acid	600 tpd	
Phosphoric acid	100 tpd	
Ammonium sulphate	2 x 300 tpd	
Ammonium chloride	75 tpd	

The small ammonia plant will shortly be shutdown and the ammonium sulphate plants are going to be replaced.

In order to assess the benefits of efficiency improvements the costs used were:-

Naphtha	2084 rupees/te = 0.183 rupee/cal (HHV)
Power Import	660 rupees/MWh (average)
Steam	150 rupees/te

3. NAPHTHA REFORMING AMMONIA PLANT

This ammonia plant operates at design rate but the efficiency is worse than design.

	Flowsheet	Current Operating	Annual Average
Naphtha fuel te/te	0.547	0.588	
Naphtha fuel te/te	0.267	0.262	
Total naphtha te/te	0.814	0.85o	0.931
G.cal/te LHV	8.65	8.93	
Power KWH/te	810	848	1100
Var. Cost rupee/te	2231	2333	2668

The areas of efficiency improvements are dealt with under the following main headings :-

3.1 Naphtha Efficiency

3.2 Electric Power Efficiency

3.3 Average Annual Efficiency

3.4 Improvements to Flowsheet

3.1 Naphtha Efficiency

The plant does not have sufficient flowmeters to do a full characterisation to establish where the efficiency losses are taking place. However, the areas of known losses that have been identified by the operating and technical staff are expected to account for the deviation from flowsheet efficiciency.

The following reasons for losses have been identified:-

1. Heat losses on primary reformer.
2. High excess air on primary reformer.
3. Poor performance of fluegas waste heat boiler.
4. Leakage on combustion air heater.
5. High auxillary firing.

6. High steam ratio.
7. Poor fired heater efficiency.
8. Losses in make gas steam.
9. Venting of scrubbed flash gas.

3.1.1 Heat losses on Primary Reformer

The refractory in the primary reformer is in bad repair and areas of the outside reformer panels are very hot. There is a proposal that the ceramic fibre insulation be added on the inside of the existing brick insulation. The only case where this has been done was because there was no time during the shutdown to remove the fire brick lining. It is more normal to remove the brick insulation and replace it with ceramic fibre, and it is recommended that this be done if time during the shutdown will allow.

The low fluegas temperature exit the primary reformer of 900^{0}C against a design temperature of 1000^{o}C will be caused by excess air on the primary reformer or by inaccurate measurement of the temperature rather than by reformer heat losses.

3.1.2 Excess air on primary reformer

The primary reformer runs with 15 - 20% excess air and there is a proposal that the burners be replaced to reduce the excess air rate. 15% excess air is quite reasonable for a primary reformer, especially with a side fired reformer with its large number of burners. The benefits to be gained from changing the burners will be small and it is better to continue to monitor the excess air and trim the reformer to keep it low, rather than replace the burners.

3.1.3 Performance of Fluegas Waste Heat Boiler

The steam raising rate measured from the boiler is not consistent and does not match the rate calculated from

the fluegas temperature drop across the coil. The flowmeter should be overhauled and historic readings from the flowmeter treated with utmost caution.
The performance of the waste gas boiler (ie, heat load divided by log mean temperature difference) which is calculated from the temperatures and estimated flue gas rate was calculated for the following cases :-

	Relative Fluegas Rate	Fluegas Temp'ture		Performance Relative to Design
		Inlet	Exit	
Design	1	828	288	1
24/8/71 (apprx)	1.1	750	320	0.81
2/9/86 "	1.2	745	457	0.41
20/2/87 "	1.1	760	320	0.82

The performance soon after commissioning was 81% of design. There was a gradual increase in auxiliary burner firing to try to maintain the indicated steam rate, which led to a very poor performance of 41% design in 1986. Since then the auxiliary firing has been decreased and the performance of the boiler has improved to 81% of design. There would appear to be two problems:-

1. Poor intrinsic performance of the boiler.
2. Fouling at high auxiliary firing rates.

The performance of the boiler now is no worse than soon after commissioning. This suggests that the reason for only getting 81 - 82% of design performance is due to a design fault rather than fouling. Either there is insufficient heat transfer surface or there is some bypassing. The diagram of the boiler coil shows that there are tubes missing through the middle of the boiler coil and that there are gaps between the boiler coil and the walls of the duct. This would cause bypassing of gas and reduced heat transfer effectiveness. The heat transfer of the boiler and the amount of bypassing should be checked by calculation and if bypassing is shown to be a problem, baffles should be installed to prevent it.

The fouling of the boiler at high firing rates could be because of lack of combustion air causing soot formation and deposition of the external boiler surface.

3.1.4 Performance of Combustion Air Heater

The combustion air heater is of the Lungstrum rotary type. From plant data it can be seen that the thermal performance (heat load divided by LMTD) has remained clsoe to design in all cases. The high stack temperature is due to a high inlet temperature to the air heater and due to high fluegas rates. The temperature of the combustion has generally been higher than design.

The air leakage on the combustion air heater is 35-45% compared to a design leakage of 28%. The high air leakage at the present time is not unduly surprising, because the inlet fluegas temperature has been operating up to 460°C against a design of 288°C. The high temperature causes increased expansion and wearing of seals. Air leakage on the combustion air heater is not a large problem unless it causes a limit on the ID or FD fans. It is recommended that the seals are renewed at the next overhaul and that the inlet temperature to the heater is restricted. The case for changing to a new tubular combustion air heater is poor.

3.1.5 High Auxiliary Firing

Naphtha is fired on the auxiliary burners during start-up in order to raise steam. During normal operation, hydrogen rich gas from the hydrofining section is burnt on the boilers. There is some involuntary gas from the refining section, but the rate has until recently been increased by raising the temperature of the naphtha stripper, in order to increase auxiliary firing, to maintain a plant steam balance. Since the steam pressure required for the MEA reboiler is only about 4 ata, and 8 ata steam is blown off elsewhere on site, it is better to import steam for MEA reboil and to reduce auxiliary firing as far as possible. The naphtha stripper temperature should be reduced from the current operating of 80°C to the design temperature of 54°C or until the point where the sulphur level at the bottom of the stripper reaches the design value.

The performance of the naphtha cooler E002 is significantly worse than design. This is causing increased hydrogen rich flash gas from the separator V001. The reason for the poor performance of E002 should be investigated and the exchanger should be cleaned or replaced if necessary.

The pressure downstream of the naphtha heater E001 is lower than design due to a high pressure drop in the heater. This is causing an increase in hydrogen rich gas from the separator V001 and from the naphtha stripper. The heater should be cleaned so that the design pressure downstream of the heater can be maintained.

There is concern that if import site steam is used for MEA reboil, a sulphuric acid plant trip will cause loss of steam and high CO_2 slip. Operators must be made aware of the situation, or an automatic control system installed, in order to increase gas firing or initiate naphtha firing on the auxiliary burners in the case of a sulphuric acid plant trip.

3.1.6 High Steam Ratio

The steam carbon ratio on the primary reformer is run at 4.0 instead of the design of 3.8. The reason for this is partly because of poor LT Shift performance, but there is also a general concern at getting carbon laydown on the primary reformer catalyst. The design steam carbon ratio of 3.8 is conservative. The Cochin Division ammonia plant is designed for a steam: carbon ratio of 3.4 and the side fired Selas reformer will enable lower steam ratios than with a top fired reformer due to a better heat flux profile. There should be no concern about running at the design steam ratio of 3.8. If there is concern about accuracy of measurement of flows, then the steam and naphtha flows can be automatically corrected for temperature and pressure and duplicate measurement can be taken to ensure accurate measurement.

The cost of running at a steam ratio of 4.0 instead of 3.8 is 0.03 G.cal/te or 260,000 rupees/year.

3.1.7 Poor Fired Heater Efficiency

There are two fired heaters - naphtha heater and naphtha vaporiser. The naphtha heater does not have a convection coil to recovery low grade heat, but the naphtha vaporiser has a convection coil to heat process air. The fluegas temperature to stack from the naphtha vaporiser is 410°C compared to a design of 325°C. The fluegas temperature and excess air rates should be measured on these heaters and the burners adjusted to keep at low excess air rates (high excess air rates will cause high stack temperatures). If the high stack temperature from the naphtha vaporiser is due to poor performance of the air heater coil (as opposed to high excess air) then there may be an economic case for increasing the size of the coil to improve efficiency. The duty of the naphtha heater is small, so there is unlikely to be an economic case for installing a convection coil.

3.1.8 Losses in Make Gas Stream

The current plant data indicates that the poor naphtha efficiency is due to high process naphtha rather than a high fuel rate. This may be due to errors in measurement of the naphtha flows, but also could be due to blow off control valve leaks or excessive compressor seal losses. The following actions should be taken :-

1. Check the gas balance from process feed to ammonia product.
2. Measure the amount of make gas that is going up vents. This can be done by injecting a known amount of nitrogen into the base of vent and taking a gas analysis further up the vent.
3. If the loss is due to blow off control valve leaks, change to soft seat control valves.
4. If there is a large loss from compressor seal leaks, pipe the blow off gas to feed through a cooler and glass wool oil separator.

Venting of scrubbed flash gas:- scrubbed flash gas from the LP water scrubber T502 is vented instead of being used for fuel because it contains water. A demister or proprietary separator should be installed in the top of the scrubber to separate water droplets.

3.2 Electric Power Efficiency

The main usage of electric power is on the process air compressor and syn gas compressor. The following work should be done to establish why extra power is being used on these machines:

1. Measure polytropic efficiency of each stage of the compressor and compare it with design.

2. Check performance of compressor intercoolers compared with design.
3. Check pressure drop of compressor intercoolers compared to design.

3.3 Average Annual Efficiency

The average annual naphtha efficiency of the ammonia plant is 10% worse than the continuous running efficiency. The main reason for this is the large number of start-ups from cold each year. Some of these shutdowns are caused by power cuts, so the planned installation of a UPS system to enable the reformer to be kept running for a short period should partly alleviate the problem. However, a significant further reduction in naphtha loss can be achieved by piping the gas that is flared during start-up to reformer or auxiliary firing fuel. This has been done successfully on most ICI ammonia plants in the UK, where it was justified on the basis of about 3 cold start-ups per year. There are about 15 cold start-ups per year on this plant of which say 5 can be prevented by installation of an UPS. About 1/3 of the 60 tes of naphtha used during a start-up can be saved giving a total saving of 200 tes/

year of naphtha. This is with 0.06 Kcal/te or 420,000 rupees/year.

A reduction of naphtha usage and power usage can be achieved by reducing the plant start-up time. A lot of progress has been made in reducing the start up times of ammonia plants and ICI plants have attained a reduction in start-up time from 40 hours to 24 hours. The experts in the area of reducing ammonia plant start-up times are Hays Mayo Enterprises Inc., Kansas, USA.

3.4 Improvements to Flowsheet

The naphtha efficiency can be improved by reducing the amount of fuel fired on the primary reformer at the expense of reduced steam raising. The lost steam can be imported from site for use on MEA. Ways of reducing fuel are :-

1. Reduce Steam:Carbon Ratio below Design
   This will require a computer simulation of the plant or a carefully monitored plant test to establish the optimum steam:carbon ratio. A ratio as low as 3:4 should be safe from a carbon laydown point of view, but an accurate assessment can be made by ICI by doing a reformer computer simulation.

2. Isolate top row of burners on Primary Reformer
   By stopping firing on the top row of burners and increasing the firing pressure on the other burners, the reformer process exit temperature is kept the same but the fluegas temperature is reduced giving a reduction in total firing. Again this can be simulated using a computer model. On the ICI Severnside IA plant which has a Selas reformer, the top two rows of burners have been isolated.

3. Increase size of air preheat coil in top of naphtha vaporiser. This is discussed previously.

4. Modify CO_2 Removal System.
If the proposed turboalternator is installed on the site to condense excess steam, then LP steam import to the ammonia plant will reduce the power from the turboalternator. The LP steam import can be significantly reduced by modifying the MEA system. Union Carbide have promoted inhibition for MEA that permits the use of higher strength solution which reduces heat consumption compared to original systems. Further reduction of reboil load is achievable using lean solution flash.

Other possible retrofits such as Selectoxo, molecular sieve diers, hydrogen recovery unit and low pressure synthesis loop will be uneconomic for this size plant.

3.5 Conclusion

The following measures can be taken to improve the efficiency of the plant:-
Monitor excess air on primary reformer and trip furnace.
Check design of fluegas waste heat boiler and check amount of bypassing.
Renew seals on combustion air heater and restrict inlet temperature.
Reduce auxiliary firing on auxiliary burners and import LP steam for MEA. Clean or replace naphtha cooler E-002.
Reduce steam ratio.
Monitor excess air on fired heaters in trim burners.
Increase size of process air heater coil in naphtha vaporiser.

Check mass balance on process stream and check for losses.
Monitor air compressor and synthesis gas compressor efficiencies.
Install water separator in top of LP scrubber.
Consider isolating top row of burners in Primary Reformer.
Retrofit MEA system to MEA - Amine guard and install lean solution flash.

4. NAPHTHA PARTIAL OXIDATION AMMONIA PLANTS

A 80 te/day plant was commissioned in 1962 and a 140 te/day plant was commissioned in 1965. The smaller plant will shortly be shutdown.

The energy usage of the partial oxidation plants is very high compared to modern plants.

	Current Operating	Annual Average
Naphtha te/te	0.84	0.926
Oxygen te/te	1.0	
Power (including ASU) MWH/te	1.75	2.5
Steam te/te	0.4	1.4
Variable Cost Rs./te.	2965	3785

The only energy improvements which were discussed were changes to improve the flowsheet efficiency. These were :-

Conversion to air partial oxidation.
Installation of a PSA unit.
Installation of a boiler or BFW heater after the ammonia converter.

4.1 Conversion to Air Partial Oxidation

Using air instead of oxygen partial oxidation will give a large reduction in the power requirement of the plant. However, there would be excess nitrogen in make gas that would have to be removed and nitrogen wash would have to be replaced for final carbon monoxide removal. The high nitrogen level in the make gas stream will give a high pressure drop.

Foster Wheeler have proposed using a cryogenic unit to remove excess nitrogen. In order to remove the CO as well as excess nitrogen the cold box would have to be of the Braun type which would incur an additional pressure drop of about 3 bar.

KTI have proposed using a PSA Unit in the make gas stream. However, KTI only have experience of using a PSA unit to remove everything except hydrogen from the process steam and add nitrogen downstream.

It would be important for anybody offering air partial oxidation technology to demonstrate the reactor using high aromatic naphtha on a pilot scale unit before using it on a new plant or as a retrofit.

It is likely that the cost of any such retrofit would be high with a significant plant downtime. It is recommended that an outline proposal with rough efficiency improvements and capital costs is provided by the retrofit vendor so that an approximate evaluation can be carried out before paying for a full feasibility study to be done.

4.2 Installation of PSA Unit

Water wash, MEA, caustic scrubbing and nitrogen wash can be replaced by a PSA unit with nitrogen addition downstream of the PSA unit. This would save about 1.6 MW in power, but on this scale plant only 80-85% of the hydrogen would be recovered in the PSA unit. This modification will not therefore be economic.

4.3 Installation of Boiler or BFW Heater in Synthesis Loop

The gas from the converter passes direct to a cooler, so the heat in the gas stream is lost. The temperature of boiler feed water into the steam drum is fairly close to saturation, so a BFW heater in the synthesis loop would cause steaming in the final BFW heater. This would mean redesign of steam drum internals and also probably the final

BFW heater. This option is unlikely to be economic. Installation of a low pressure boiler to export steam to the site should be feasible and warrants further evaluation. It was reported that about 12 te/hr can be raised at 4 ata. However, this appears optimistic and a rough calculation shows that 4 te/hr is more realistic.

5. SITE STEAM SYSTEM

The sulphuric acid plant raises 28 te/hr of steam at 16 ata. This is let down to the 8 ata site main and passed to LP steam consumers. The largest LP steam users are the Ammonium Sulphate Plants, but they are soon to be replaced. The site will then have a large steam surplus.

A proposal has been made by Technical Services to rationalise the steam system and install a turboalternator to condense 16 ata steam generating 4.0 MW power. The proposal looks good and should be evaluated in more detail. The 4.0 MW power is worth about 22 x 10^6 rupees/year, while the capital cost of the turboalternator will be about 40 x 10^6 rupees

SECTION E.

EVALUATION OF ENERGY CONSUMPTION AND IDENTIFICATION OF AREAS OF INEFFICIENT ENERGY USE AND REMEDIAL MEASURES

FERTILIZER CORPORATION OF INDIA LTD

RAMAGUNDAM UNIT

RESTRICTED

APRIL 1987

ENGLISH

ENERGY MANAGEMENT IN THE INDIAN FERTILISER INDUSTRY - RAMAGUNDAM

DP/IND/85/006/11-01

TECHNICAL REPORT: EVALUATION OF ENERGY CONSUMPTION IN INDIAN FERTILISER PLANTS AND THE IDENTIFICATION OF AREAS OF INEFFICIENT ENERGY USE AND REMEDIAL MEASURES

Prepared for the Government of India by the
United Nations Development Organisation
acting as executing agency for the
United Nations Development Programme

Based on the work of Warwick J Lywood Expert
in Energy Management

United National Industrial Development Organisation
Vienna

CONTENTS

1 SUMMARY - FCI RAMAGUNDAM UNIT

The efficiency of the ammonia plant is very much worse than design on an annual basis. A large part of the efficiency loss is due to operating at part rate with only one or two of the gasifiers working. The greatest single efficiency improvement could be obtained by improving the availability of gasification by installing a fourth gasifier.

Areas of plant and equipment where there are efficiency losses were discussed and a number of areas where action can be taken to improve efficiency are recommended. The major areas of possible improvement below, could give an efficiency improvement worth 8.1 crores rupees per annum.

	Saving 10^6 rupees/yr
Clean first compressor stage and intercoolers of air compressor	4
Raw gas compressor	9
Synthesis gas compressor	6
Return nitrogen wash tail gas to RGC suction	14
Return ammonia loop flash gas to SGC suction	3.5
Burn gas flared during start-up	N/E
Avoid running two nitrogen compressors at two gasifier rate	16
Eliminate continuous oil firing on SGP	15
Modify SGP import steam superheater	0.5
Control of steam system	12
Improve cooling system	N/E
Connect urea low pressure steam to ammonia plant system	1.2
TOTAL:	81
Install fourth gasifier	about 139

N/E = Not estimated

2 INTRODUCTION

The Ramagundam Factory was designed to produce 900 te/day of ammonia and in turn 1500 te/day of urea. A facility was added to produce 120 NM^3/hr of argon. The project was started in 1971 and commissioned in 1980.

The Feedstocks for the factory are:

Coal	3000 te/day
Power	55 MW
Water	15 million gallons/day

The project was executed and completed by FCI Ltd with their Planning and Development Department (now Planning and Development of India Limited) supplying design and engineering data.

Coal gasification technology was supplied by Krupp-Koppers, rectisol purification by Lurgi and ammonia synthesis by Technimont. Technimont also licenced the urea process.

In order to assess the cost of efficiency losses, the costs used were:

Coal	384 rupees/te
Fuel oil	1798 rupees/m^3
Power	658 rupees/MWH
Superheated HP steam	102 rupees/te
Saturated HP steam	92 rupees/te
MP steam	82 rupees/te
LP steam	46 rupees/te
Ammonia	3500 rupees/te

3 AMMONIA PLANT

The ammonia plant was designed to produce 900 te/day of ammonia. The plant was started in 1971 and commissioned in 1980. The main sections of the plant are:

	No of streams	Licensor
Coal preparation	1	
Air separation	2	
Coal gasification	3	Krupp Koppers
Raw Gas compression	1	
Rectisol gas purification	2	Lurgi
Nitrogen wash	2	
Syn gas compression	1	
Ammonia synthesis	1	Technimont

The performance of the plant is given on the next page:

	Design	Anticipated Performance	Achieved: Good Day	Achieved: Annual Average Aril 86 - January 87
Capacity MTD	900	900	757	(2) 454
(1) Coal te/te	1.828	1.91[illegible]	2.07	2.43
Super heated 90 ata steam te/te	9.01	10.1	11.3	19.5
Sat HP Steam export te/te	-3.45	-2.9		-1.84
35 ata steam to urea te/te	-1.37	-1.5	2.1	-1.95
LP steam export te/te	-0.1	-0	1.7	-2.65
Net steam te/te	4.29	5.7	7.5	13.1
Power kWH/te	699	789	1070	1486
Variable cost rupees/te	1649	1894	2422	3447

(1) Corrected to 5100 kcal/kG

(2) On basis of 3[illegible]0 stream day/year

The main reason for the very poor steam consumption figures compared to design is that the plant is run most of the time on only one or two gasifiers. With only one or two gasifiers working, the steam generation from the plant is low, but the demand for steam from the Raw gas and syn gas compressors is still very high. This means that steam generated in the SGP must be higher than design, so the steam efficiency is much worse.

The electricity usage is high because both ASUs, nitrogen compressors and oxygen compressors have to be run when only 2/0/0/3 gasifiers are in commission.

The best way of improving the efficiency of the ammonia plant will be to increase the availability of gasification by installing an extra gasifier. For example the variable cost could be reduced to about 2800 rupee/te and the annual output increased to about 650 te/day. This is worth 139 x 10^6 rupee/yr in variable costs on an ammonia value of 3447 rupee/te.

The areas for plant efficiency improvement are dealt with section by section.

3.1 COAL PREPARATION

Heater for Drier

Impure nitrogen from the gasholder and recycle gas are heated in an oil fired drier heater to 250°. The fluegases are used for drying coal in a ball mill and lifting the powdered coal to a hopper at 90°C. Most of the hot gases are recycled to the fired heater but some gas is purged and vented in order to remove the moisture from the system. There was a proposal that the purge gas be used to heat up the nitrogen addition or the combustion air for the fired heater. A typical economic approach temperature on a combustion air heater is 50°C. The amount of heating that can be done with gas at 90°C will be very small and the scheme will not be economical.

3.2 COAL GASIFICATION

3.2.1 Surplus Low Pressure Steam

The steam generated in the gasifier jackets is 12 te/hr instead of a design of 10 te/hr each. The deaerator steam requirement is about 13/te instead of 19 te/hr because of the lower steam raising rate. The steam requirement for gasification is 13 te/hr compared to 17 te/hr design. However, the steam import of excess steam from urea which was designed to be 3 te/hr has not been commissioned. The LP steam surplus on gasification compared to design is 13 te/hr. To avoid blowing off this steam, a dump condenser has been installed to condense the excess steam on a heater to heat the demin water to the steam generation plant deaerator. This will enable the condensate to be recovered and will reduce the demand for 5 ata steam on the SGP deaerator. This in turn will save steam let down from 35 ata and 90 ata giving a reduction in HP steam generatior. The surplus 3 ata steam from urea plant will be exported to gasification so that a similar benefit can be obtained. This should be done quickly.

3.2.2 Quenching of Raw Gas

The gasifiers are operated at 1600°C. The gas is cooled by directly quenching to about 850-900°. the heat in the gas down to 300-250°C is used to raise steam in a waste heat boiler. The design quench temperature was 1100°C with a temperature exit the boiler of 350°C. As a result of the increased quenching, the steam raising from the boiler has been reduced from 25 te/hr design to 15 te/hr.

While it is probably not possible to recover any more heat above 900° it is possible to recover more heat below 350°C. At present, the minimum temperature approach on the waste heat boiler is 250°C (350-100). By enlarging the boiler and economiser coils, the exit temperature from the waste heat boiler can be reduced to 200 or may be even 150°C. This would give an increase in steam raising of 4-5 te/hr per gasifier worth 2.3 x 10^6 rupees/yr. However, the enlarging of the coils would need a complete redesign of the waste heat boiler and would be expensive. While it could not be justified on its own, if the boiler were redesigned in order to improve the reliability, the fluegas temperature should be reduced.

3.2.3 Operating Waste Heat Boilers at Lower Pressure

A proposal has been made that the boiler be operated at a lower pressure in order to reduce the frequency of tube failures. The main reason for tube failures is erosion caused by fly ash hitting the tubes. Tubes will fail when the erosion reduces the tube thickness to a certain minimum value. A reduction in pressure from 105 ata to 35 ata will reduce the minimum allowable thickness by a factor of 3. However, the original thickness of the tubes is about 3 or 4 times the minimum thickness for failure so assuming erosion at a constant rate the life of the tube will only be increased by about 20%.

Problems in operating at a lower pressure could be:

Reduced circulation
Insufficient water/steam separation
Possibility of flow instabilities
Steaming in economiser coil

Reduced circulation may well not cause dryout of tubes and flow instabilities will probably not be a problem, but the conditions will be outside the range of experience of boiler vendors, so no guarantee of well being can be given. Insufficient water/steam separation will definitely be a problem as operating pressure is reduced and the steam drum primary separators and demisters will need to be replaced. Steaming in the economiser will need primary separation installed for the feed water to the drum and could cause flow instabilities in the economiser coil.

3.3 AIR SEPARATION UNIT

At full ammonia plant rate, the steam rate on the ASU air compressors is about 54 te/hr each instead of an expected figure in the site steam balance of 42 te/hr. There are three reasons for this:

1) Poor performance of the process air compressor. This causes an extra steam rate of 4 te/hr.

2) The air rate needed on the ASUs is higher than design for a given ammonia rate. This gives extra steam of 6 te/hr.

3) Scaling the guaranteed steam rate on the air compressor at full output of 52 te/hr to account for the design margin in the ASUs gives a steam rate of 44 te/hr at full ammonia plant output and flowsheet oxygen usage. The 42 te/hr figure that was used on the steam balance is optimistic.

Calculations were done using operating data to find the performance of the machines.

The efficiency of the first compression stage is below design giving an increased steam consumption of 2 te/hr.

Inspections during overhauls have shown the first compression stage, intercoolers and aftercoolers to be fouled. This is due to poor performance of the inlet air scrubber filter and also due to cooling water fouling on the exchangers.

It is recommended that:

1) The design of the air scrubber filter is improved so that dust carryover does not occur.

2) The intercoolers and aftercoolers are cleaned and if possible, are redesigned to increase the cooling water velocity.

3) The air flow from the air compressor, the net oxygen flow from the ASU, the oxygen flow to gasification and the raw gas flow are measured while the raw gas flow analysis is done. This will enable the determination of the cause of the high air rate per unit raw gas rate. ie whether the cause is due to performance of the ASU or to extra oxygen for gasification.

4) A test is done on the ASUs to find the relationship between oxygen product rate and oxygen purity. This will enable a calculation to be done to find the optimum oxygen purity for gasification.

3.4 RAW GAS COMPRESSOR

The raw gas compressor is using much more steam than design. Calculations from a set of plant data showed:

The first two stages of the compressor had lower temperature rises than design.

The last two stages of the compressor had higher temperature rises than design, indicating poor efficiency. This accounts for an extra 3 te/hr of HP steam to condensing.

All the intercoolers are performing worse than design. This accounts for an extra steam usage of 3 te/hr.

No 15 bar steam is imported to the machine. This accounts for extra HP steam usage of 8 te/hr.

The vacuum on the steam turbine was 0.265 bar instead of a design of 0.125 bar. This accounts for an extra 5 te/hr of steam.

Taking these factors into account and assuming design efficiencies for the first two stages the steam rate is close to what would be expected. The low temperture rises on the first two stages of the compressor are probably due to water carry-over from the upstream catchpots. Evaporation of the water in the compressor will then cool the gas. The efficiency of the first two stages of the compressor cannot therefore be calculated and they could also be fouled.

It is recommended that:

1) Carry-over from the suction catchpot and first intercooler

catchpot be checked by measuring the water condensed in the first and second intercooler catchpot to see whether it matches the rate calculated from the temperature and pressure of the compressor.

2) The compressor rotor is cleaned and the reason for fouling is investigated.

3) The compressor intercoolers are cleaned and if possible redesigned to increase the cooling water velocity.

4) The vacuum on the steam turbine is checked and if correct, the reason for the poor vacuum determined either air leakage or poor ejector performance.

3.5 SYNTHESIS GAS COMPRESSOR

At 92% of design gas rate and a significantly lower delivery pressure than design, the HP steam rate was 220 te/hr and the IP steam rate was 73 te/hr compared to expected rates on the site steam balance of 223 and 32.8 te/hr respectively.

The figures given at the top of the log sheets indicating design operating temperatures exit each compressor stage are too high and probably correspond to mechanical design temperatures rather than operating design temperatures. The correct design operating temperatures exit each stage of the machine were not available, so the design polytropic efficiencies could not be calculated. Using estimated efficiencies for this sort of machine, the power calculated for the design rate coresponded to the design power of 14.9 MW. However, the steam rates used on the site steam balance are shown by the turbine characteristic to only give a power of 13.5 MW. To get the flowsheet power, the steam rates are 215 te/hr HP and 53 te/hr IP from the guarantee point on the turbine characteristic. These rates should be taken as the basis for determining the performance of the machine.

The performance data for the machine showed that:

The suction cooling and all the intercoolers were performing worse than design. This accounts for an extra 8 te/hr of steam

from HP to LP. The temperature rise on the first stage of the machine was low, but the polytropic efficiency of all the other stages of the compressor were close to what would be expected.

The adiabatic efficiency of at least one stage of the turbine is worse than design. The data from 1984 indicated the HP stage efficiency to be poor, while data from 1986 indicated the IP stage efficiency to be poor. This would account for an extra steam usage from HP to LP of 7 te/hr.

The known reasons for higher steam usage above do not account for all the increase in steam on the tubine. The rest of the increase in steam rate may be due to leaking of the anti-surge control valve.

It is recommended that:

1 The guarantee figures for steam usage on the turbine characteristic are used as a base case for determining steam usage.

2 The intercoolers are cleaned and if possible redesigned to increase the cooling water velocity.

3 The temperature measurements inlet and exit the first stage of the compressor are checked.

4 Leakage on the antisurge control valve is checked by injecting a gas in the machine and analysing for the gas upstream of the injection point.

5 The efficiencies of the turbine stages are monitored and the turbines examined for reasons for lost efficiency when they are next opened up.

3.6 METHANOL CONSUMPTION ON RECTISOL

The loss of methanol from the Rectisol system is 2 or 3 times greater than the expected consumption of 3 tes/day. The calculated losses during continuous running due to the vapour pressure of methanol in exit gases at the design operating temperatures are reported to be only about 1.5 te/day. The additional losses could be due to:

Higher losses in exit gases due to the higher than design continuous operating temperatures when only one gasifier is running. Higher losses in exit gases due to higher than design operating temperatures during start-up of the rectisol system. Droplets of methanol being entrained in exit gases due to poor performance or poor design of gas/liquid separators. Liquid losses from the rectisol system.

Recommendations to find the reasons for higher than design losses are:

1 Calculate losses in exit gases due to higher vapour pressure of methanol at high temperature continuous running conditions and at start up.

2 Check the design of gas/liquid separator. If the viscosity of the Rectisol solution is around that of water then the area of the separators should be such that under all operating conditions:
density x velocity^2 < 11 kg/Msec^2 for wire mesh pads
density x velocity^2 < 45 kg/Msec^2 for Peerless separator
If viscosity of Rectisol solution is high then larger areas are needed.

3 Monitor the level of methanol in the HT shift water circuit.

4 Monitor the level of methanol in the drains from the Rectisol plant area.

3.7 TAIL GAS FROM NITROGEN WASH

About half of the tail gas from the Nitrogen wash is burnt in the coal preparation plant and the remainder is flared. The gas that is currently flared could be burnt in a gas burner in the new boiler to be installed.

However, the hydrogen and CO in the tail gas represents 4% of the potential hydrogen made in the gasifers and is worth more as feed than as fuel. If the tail gas is returned to the raw gas compressor suction, the CO will be converted to hydrogen in shift conversion and when on a gasification limit the plant output can be increased by 3.4%. When the plant is operating on one or two gasifiers the main machines are on anti-surge, so there is practically no cost in making ammonia from tail gas that would otherwise be flared. If the alternative value of tail gas is as supplement for coal on a boiler, the cost of the extra ammonia would be 450 rupees/te. If the tail gas used for the drier in the coal preparation section were returned to the raw gas compressor and replaced by fuel oil, the effective cost of the extra ammonia would be 1140 rupees/te. Some fraction of the tail gas - say 15% must always be burnt in order to stop the level of argon from building up too high in the section of plant between raw gas compression and nitrogen wash.

When three gasifiers are operating, it will always pay to put tail gas to the suction of the raw gas compressor instead of flaring it, or burning it on a coal boiler, but the case for replacing tail gas with fuel oil on the drier will be marginal.

When the plant is able to achieve running at full rate for a significant proportion of the time, there will be a good case for installing a low temperature shift after the high temperature shift in order to convert more CO into hydrogen. This will avoid the cost and any limitations in raw gas compression or H_2S removal that are incurred when tail gas is returned to the raw gas compressor suction.

It is recommended that:

1 When one or two gasifiers are running, about 90% of the tail gas is returned to the raw gas compressor suction and fuel oil is used to replace tail gas on the coal preparation drier.

2 When three gasifiers are running, tail gas is used on the coal preparation drier and the remainder is passed to the suction of the raw gas compressor.

3 When three gasifiers are running for a significant proportion of the time then installation of an LT shift should be examined.

The benefit from returning tail gas to the suction of the raw gas compressor suction will be about 14 x 10^6 rupees/yr, valuing ammonia at 3500 rupees/te.

3.8 FLARING OF GAS DURING START-UP

There are about 25 start-ups a year during which gas is flared from the make gas stream flares for about 48 hours at a time. Most ICI ammonia plants in the UK have been retrofitted to use flared gas during start-up as fuel on the primary reformer. The control equipment and extra pipework was justified on the basis of about 3 start-ups per year.

It is recommended that a gas burner is installed in the new coal fired boiler, so that flared gas can be burnt during start-up.

3.9 NITROGEN COMPRESSORS

For about 75% of the time the plant is running at about 65-70% design rate, but the nitrogen compressors are each designed for 55% of design rate. It has been proposed that instead of running two nitrogen compressors, only one nitrogen compressor should be run, while the remaining nitrogen required is added at the suction of the raw gas compressor. This looks to be a good idea. A computer simulation of the nitrogen wash system would need to be done in order to evaluate this idea on paper and even this may not be conclusive. The potential savings are about 16 x 10^6 rupees/year. In view of the large savings and since a break-in to the raw gas compressor suction will be needed anyway for tail gas recycle, it is recommended that the pipework for passing nitrogen to the raw gas compressor suction is installed anyway and the feasibility of this idea evaluated on the plant.

3.10 AMMONIA CONVERTER BOILER

The boiler is only anticipated to give 45 te/hr of steam at full rate instead of the 55 te/hr of steam predicted in the design. This was assumed by Krupp Koppers to be due to low convertor temperatures. However, this cannot be the cause of the problem.

A rough check of the flowsheet heat balance around the converter gave a steam rate of 51 te/hr instead of the 55 te/hr design. The rest of the steam shortfall is due to the temperature approach on the cold end of the interchanger being greater than design. This could be due to two possible reasons:

1 The interchanger is fouled or is underdesigned.

2 The interchanger is being partly bypassed in order to reduce the converter operating temperatures. High converter operating temperatures would be caused by fouling or underdesign of the boiler coils.

It is recommended that:

1 The flowsheet heat balance around the converter is checked accurately.

2 The bypass rate on the interchanger be determined.

3 The heat exchanger and boiler are inspected for fouling by catalyst dust when the converter is next opened up.

3.11 AMMONIA LOOP FLASH GAS

Flash gas from the ammonia let-down vessel is designed to be used as fuel, but is currently being flared. This gas represents 0.7% of the potential ammonia rate. It is recommended that this gas is returned to the suction of the synthesis gas compressor. Since the syn gas compressor and ammonia loop are never on a limit. This will result in an additional ammonia make of 0.7% worth 3.6 x 10^6 rupees/yr.

4 ANCILLARY EQUIPMENT

4.1 STEAM GENERATION PLANT

The SGP consists of three coal fired boilers designed to raise 90 ata steam. The performance is given below:

	Design	Achieved	
		Good day	Annual average April 86-Jan 87
Production rate		358	279
Coal te/te	0.153		0.153
Oil te/te	0		0.007
Power kWh/te	23		23 assumed
Steam te/te	0.15		0.15
Cost rupees/te	88		102

The cost of steam at other pressures and conditions can be calculated from the cost of superheated 90 ata steam by considering the additional HP steam raising needed if the consumption rate of steam at other levels is changed.

	Cost rupees/te
Superheated 90 ata steam	102
Saturated 105 ata steam	92
35 ata steam	82
5 ata steam	46

4.1.1 FURNACE OIL FOR STABILITY OF FLAME

Oil burners are needed to maintain stability of the furnace coal flames when the boiler is operating below 60% load. Since the coal for firing each boiler is provided by two mills, the furnace load will fall to 50% when one mill stops. An oil burner is always kept on line, so that the oil firing can be increased to prevent the boiler from tripping when one coal mill fails. This is resulting in the use of more furnace oil which is 4-5 times costlier than coal on an equivalent energy basis. If 70% of the furnace oil firing could be eliminated, the saving in cost would be 15×10^6 rupees/year.

There are two problems:

1 Keeping Oil Hot

The oil circulating system is kept hot by a steam heater. There is a leak on the steam control valve to the heater which is reported to cause overheating of the oil circulating system unless some oil is used on the boiler. The cause of the steam valve leaking is probably that it is operating for a large proportion of the time nearly shut because only one oil burner is being run at minimum firing. This would cause wearing of the valve trim leading to a leaking valve. The problem would probably disappear if the valve were replaced and the system were operated as designed. An alternative way of solving the problem is to replace the existing valve, but also install a small steam valve in parallel which would be used to control the steam rate if there were only one oil burner firing at low rate.

2 Lighting of Burner

There is insufficient time to light the oil burner and turn up the oil firing between the coal flame failing due to a mill stopping and the boiler tripping. There is sufficient coal dust in the bowl of the mill to keep the coal burner alight for about 1/2-1 minute, while it takes about 15 seconds to light the oil burner when the operator is in position. When a mill stops, the boiler trips due to a high furnace pressure, but the reason for the high pressure occurring could not be explained.

It is recommended that:

1 A dynamic simulation of the boiler and its control system be written in order to find the exact cause of the boiler tripping.

2 A delay in tripping the boiler of say 20 seconds be built into the control system in order to give time for the control system to respond to the high pressure in the boiler. This is standard in many natural gas based primary reforming furnaces on ammonia plants.

3 The feasibility be examined of using a control signal from the mill stopping, initiating automatic start-up of the oil burners. The oil burners would then be lighted before the coal burner went out.

4.1.2 IMPORT STEAM SUPERHEATER

There is a coil in the convection section of the boilers before the economiser and combustion air heater to heat up import saturated steam from the ammonia plant. The import steam rate is

much lower than design so the import steam superheater is kept isolated in one or two boilers. This results in a stack temperature of 200°C instead of a design temperature of 180°C. A solution to this problem is to transfer part of the import superheater coil to increasing the BFW temperature. The design operating temperature of the BFW exit the existing economiser is well short of saturation temperature, so there should be no problem in raising this temperature. The size of tubes and number of parallel paths in the superheater coil and economiser coil are different but, the lower velocity of water in superheater coil should not be a problem.

It is recommended that the boiler vendor be asked to examine the feasibility of using part of the import superheater coil for BFW heating. The cost of the high stack temperature is about 0.5×10^6 rupees/yr.

4.2 STEAM SYSTEM

There are a number of problems with control of the steam system.

The boiler firing control on the coal boilers to control the HP steam main pressure is not functioning due to a lack of spare parts for the control system. This means that blow-offs are kept open on the MP and LP steam headers in order to provide control for the system. It is recommended that spare parts are obtained and the control system is recommissioned.

The original plant was designed with the low pressure steam to the SGP deareator being supplied directly from HP steam main. The plant has been set up so that the LP steam to the SGP deaerator is supplied from the 5 ata ammonia plant header. At the design machine performance the extra steam needed could all be supplied without let-down from HP to MP or MP to LP by

adjusting the extraction rates from the raw gas compressor and syn gas compressor turbines. However, the performance of the raw gas compressor is below design, so the turbine is operating close to the limit of HP steam rate and steam extraction rate. This means that it is not possible to avoid let-down from HP to MP pressure under high rate operating conditions.

The control loops, that were designed to control the MP and LP pressures by adjusting extraction from the main compressor turbines are not commissioned. This is because vibration was caused by the control system not functioning properly. As a result of these loops not being in commission the extraction rates are not adjusted to the optimum values when there are changes in operating conditions of the plant.

The losses due to the poor control of the steam system are in the order of 15 te/hr of HP steam worth 12 x 10^6 rupees/hr.

It is recommended that the steam control system is analysed by an experienced control engineer and a dynamic simulation model of the system is put together and run in order to establish the optimum settings of control loop gains etc. It is standard practice in ICI to do a dynamic simulation of steam systems for all of its new ammonia and methanol plants.

4.3 COOLING SYSTEM

The cooling water system has a number of problems:

1 The cooling water velocities in many heat exchangers are low causing fouling problems.

2 The flowrates of cooling water to many exchangers are not known. There is no knowledge of the system being checked at the late design stage to see that pressure drops in the system were such as to give the design water flow to each exchanger.

3 Fouling in heat exchangers is worsened by biological matter binding together mineral particles.

4 There is poor air distribution in the cooling tower cells.

The approach between wet bulb temperature and recooled water temperature is not monitored so the performance of the cooling towers is not known.

In some cooling tower cells, the original aluminium fan blades have been replaced by Fibre Reinforced Plastic (FRP) in order to save fan power. While fan power has certainly been saved, the air rates through the tower were in many cases decreased. There is normally more benefit in increasing the tower cell air throughput rather than reducing the fan power, so it is not clear that the FRP blades are giving an overall benefit.

The poor air distribution in the towers is probably a result of poor water distribution.

Recommendations:

1 The cooling water flowrate to each heat exchanger should be determined. This can be done by injecting a radioactive tracer pulse at the pump suction and determining how long it takes the pulse to reach different points in each leg of the system. The flow rate can be determined from the volume of the system between measuring points.

2 It should be assessed what is the optimum distribution of cooling water and modifications made to improve the distribution. For example compressor intercoolers will perform quite well with a cooling water temperature rise of $15^{o}C$-$20^{o}C$ as long as the velocities are adequate, whereas in condensers, the cooling water temperature rise must be kept low. Modifications would include installing orifice plates, installing larger lines, putting some exchangers in series instead of parallel and redesigning heat exchangers.

3 Control biological growth by shock doping with chlorine. Chlorine should be added once per shift over the period that the cooling water takes to go round the system and at a level to ensure 1 ppm free chlorine in the return cooling water.

4 Install on line cleaning eg Taproge in turbine condensers and air raising nozzles between baffles in exchangers where cooling water is on the shell side.

5 Monitor cooling tower performance and check water distribution in the towers.

A five degree centigrade reduction in cooling water temperature is worth about 9×10^6 rupees/yr.

5 UREA PLANT

The urea plant uses the Technimont process with detailed design by PDIL. The plant has a capacity of 1500 te/day in 2 x 50% streams. The design and achieved performance data are:

	Design	Achieved	
		Good Day	Annual Average April.86-Jan.87
Production rate MTD	1500	1320	746
Ammonia te/te	0.58	0.574	0.61
35 ata steam te/te	0.82	1.20	1.19
Power KWh/te	190	237	316
Cost rupees/te	2157	2233	2408

The urea production rate is limited by ammonia availability.

5.1 STEAM CONSUMPTION

In the plant design it was intended to use a steam compressor to compress 4 te/hr per stream of 3 ata steam from the partial condensers up to 6 ata to be used in the process. These steam compressors did not work satisfactorily and due to various losses, the demand for 6 ata steam is 6.6 te/hr per stream instead of 4 te/hr. As a result of this, 6.6 te/hr of 35 ata steam is let down to 5 ata while excess 3 ata steam is blown off.

It is proposed that a steam ejector should be installed to use 4 te/hr of 35 ata steam to compress 2.8 te/hr of 3 ata steam in order to give the required rate of 6 ata steam. This would reduce the demand of 35 ata steam for both streams by 5 te/hr and be worth 1.6×10^6 rupees/yr.

If the urea plant is considered on its own then installation of an ejector would be a good project. However, it is planned to commission the connection between the urea 3 ata steam system and the ammonia plant 2 ata system so that the excess steam on urea can be used for heating the demin water to the SGP deaerator. When this has been done the installation of the ejector will simply mean that the let-down on the ammonia plant from 35 ata to 5 ata to supply steam for the SGP deaerator will have to be increased by about 5 te/hr with no overall gain in site steam efficiency.

However, if the ammonia plant steam control system could be recommissioned and the power on the raw gas compressor could be reduced, then the let-down from 35 ata to 5 ata could be eliminated and the ejector on the urea plant would become a worthwhile investment.

5.2 CONDENSATE

Part of the urea condensate is used for preheating the ammonia. Due to minor leaks in this heat exchanger, the condensate is contaminated with ammonia so it is drained instead of being re-used. This results in overloading of the demin water plant.

The cause of leaks in the heat exchanger is thought to be in the tubesheet. There is sometimes a problem of the tubesheet being porous due to poor manufacture causing leakage. This can sometimes be overcome by covering the tubesheet with weld metal to give a non-porous layer.

The level of ammonia in the condensate at which it would be safe to re-use the condensate was not known and the advice of a water chemist should be sought.

The polishing unit should be regenerated, so that at low levels of contamination, the condensate can be passed through the unit and be re-used.

The use of ammonia instead of morpholine for boiler feed water pH control is non-standard and direction from a water chemist should be sought.

SECTION F.

EVALUATION OF ENERGY CONSUMPTION AND IDENTIFICATION OF AREAS OF INEFFICIENT ENERGY USE AND REMEDIAL MEASURES

FERTILIZER CORPORATION OF INDIA LTD

GORAKHPUR UNIT

CONTENTS

1. SUMMARY - FCI GORAKHPUR UNIT

The capacity utilisation and efficiency of the plants on the site is poor. Only part of the difference between the achieved and flowsheet efficiencies is accounted for on the ammonia plant. While some of the efficiency losses on the plants are due to operating at below design steam rates and due to breakdowns and start-ups, these reasons only account for about half of the total loss on the ammonia plants. There are major improvements to be made by better operation and maintenance and rectifying design errors. Better monitoring and complete loss accounting is needed to identify and quantify the reasons for these efficiency losses.

Areas of plant and equipment where there are known to be problems were discussed and a number of areas where action can be taken to improve efficiency are recommended. These are given in detail in the main text, but the major areas of improvements are summarised below :

Ammonia Plant

Increase plant operating rate to design rate
- install new nitrogen compressor
- uprate oxygen compressors

Clean or redesign refrigeration condensers

Stop throttling suction of syn gas compressors

Install liquid nitrogen storage for rapid start-up of Air Separation Units

Improve operation of cooling water system

Urea Plant	Saving 10^6 rupees/yr
Improve performance of ammonia recovery column)	
Improve design of ammonia absorber cooler)	58
Refit scrubbing system on urea prilling towers)	
Convert to modern urea process	up to 47
Steam Generation Plant	
Improve boiler thermal efficiency	up to 8.3
Operate two out of three boilers	1.9
Install/run steam driven BFW pump	2.7
Send hot urea condensate to SGP	N/E
Install site turbo alternator	52

2. INTRODUCTION

The FCI factory at Gorakhpur has the following fertilizer plants.

	Capacity	Commissioned
Ammonia A&B streams	2x190 tpd	1968
Ammonia C stream	190 tpd	1976
Urea A&B streams	2x300 tpd	1968
Urea C stream	350 tpd	1976
Steam Generation Plant	3x55 te/hr	1968

In order to assess the benefits of efficiency improvements the following variable costs were used.

Naphtha (process)	2344 rupee/te
Naphtha (fuel)	5505 rupee/te
Fuel oil	2000 rupee/te
Coal	550 rupee/te
Power	950 rupee/mwh

41 ata steam (sup)	120 rupee/te
14 ata steam (sat)	120 rupee/te
4.5 ata steam (sat)	105 rupee/te
Ammonia	4150 rupee/te
Urea	3316 rupee/te

3. AMMONIA PLANT

The ammonia plant was designed and installed by TEC Japan to produce 350 tes/day of ammonia in two 50% streams and to achieve 330 stream days/year. Erection work was completed in 1968 and commercial production started on 1/1/69. Later in 1969 it was decided to uprate and expand the plant capacity utilising existing installed spare machines and adding other equipment. The existing streams were generally uprated to 190 te/day each and a third stream added to give a total capacity of 570 tes/day and 300 stream days/year. The third stream came into commercial production on 1/4/76.

The process route for production is naphtha partial oxidation using Shell Gasification, HT Shift conversion, CO_2 removal by hot potash followed by MEA and caustic soda wash, nitrogen wash and ammonia synthesis at 365 kg/Cm^2. The third stream is similar to the first two except that Benfield is used instead of hot potash.

Since the addition of the third stream capacity utilisation has been low and average efficiencies have been poor.

	Design (flowsheet)	FICC norm	Achieved Performances Good day	Good Month	*Annual 1986-7
Capacity te/day	570	505	439	380	
Naphtha te/te	0.773	0.825	0.766	0.808	0.855
Power kwh/te	1450	1765	1580	1705	2059
Steam 41 ata sup import	2.53				
41 ata sat export	1.45				
14 ata import	0.15				
Net MP steam	1.23	1.59	1.50	1.68	1.93
4.5 ata steam export	0.43	N/A	N/A	N/A	N/A
Variable cost rupee/te	3300				4150

* To February, 1987

The efficiency losses and improvements are divided into four main sections :

3.1 Loss Accounting

3.2 Equipment Problems

3.3 Increase of Plant Capacity

3.4 Other Efficiency Improvements

3.1 Loss Accounting

Loss accounting for production and efficiency losses is done in detail each month. However, the figure used as the base for loss accounting is the 'FICC norm' figure which is set by FICC as an average efficiency that should be achieved by the plant. In all loss accounting exercises, the flowsheet figures must be used as the base if all the reasons for losses are to be identified. Once all the reasons for losses have been identified then ways of reducing or eliminating them can be evaluated.

On a plant such as this where naphtha is used only for feed, naphtha efficiency accounting should be very easy. Losses can only occur due to poor carbon efficiency in gasification, flaring gas, high CO slip or ammonia losses.

3.2 Equipment Problems

3.2.1. Gasifier Waste Heat Boiler

The coil of the Waste Heat Boiler has to be replaced often due to erosion. It is unlikely that any significant benefit can be achieved by changing the material of the coil. The only remedy for erosion is a redesign of the boiler either by putting in baffles or changing the geometry of the coil in order to reduce gas velocities.

3.2.2. CO_2 Regenerator

There is a high fluctuating pressure drop in the regenerator. This problem has arisen since Ucon addition was stopped and is due to foaming. Ucon addition should be resumed.

It was also stated that the pressure drop in the regenerator of A&B streams was high (0.6kg/cm^2 at 90% load, instead of 0.3 kg/cm^2 design) before the Ucon addition was stopped. This is probably due to deterioration of the packing in the regenerator. It is recommended that :

1) The time period over which the pressure drop increased is determined from past records.
2) If the pressure drop has increased over a period of time, it can be assumed that the packing has deteriorated and should be replaced.

3.2.3. CO_2 Overhead Condenser

The performance of this exchanger is poor and the exit gas temperature rises to 70°C instead of the design of 50°C. This results in higher temperature gas going to CO_2 compression. The condenser is cleaned regularly on both sides. The fouling is probably due to sludge deposition on the cooling water side of the exchanger due to low cooling water velocity and high metal temperature. The maximum design tube metal temperature should be 6oC for organic-phosphate cooling water treatment in order to prevent sludge deposition. In the overhead condenser the process side heat transfer coefficient is likely to be high due to condensing. This will give high metal temperatures.

It is recommended that the performance of the exchanger be monitored and the performance compared before and after tube side cleaning and shell side cleaning. If the fouling is mainly on the tube side, then the exchanger should be redesigned with a higher cooling water velocity, or should be cleaned regularly. Alternatively the heat load on the cooler can be reduced. (See section 5.4)

3.2.4. ASU Air Compressor Intercooler

The delivery temperatures of the air compressor stages get high due to cooling water fouling of the intercoolers. In order to enable easy cleaning of the tubes, the top end coolers are left off. The water flow is gagged in order to stop water spilling over the channel end. The level of water in the channel end is about 5 inches.

A level of 5 inches in the channel end will give a low water velocity and low water flowrate in the heat exchanger. This is bound to cause a high suction temperature to the next stage of the compressor, increasing the power requirement. It is recommended that the end cover be put back, between cleaning periods and that the air temperature exit the cooler is monitored in order to determine when cleaning is necessary.

3.2.5. Moisture in Nitrogen Compressors

When any of the three nitrogen compressors is restarted after being shut down, the level of water in the nitrogen takes about 6 hours to fall to the non-detectable limit on the analyser. This causes wasted power worth 8500 rupees on each compressor restart in addition to other consequential losses. It is assumed that the water is from an intercooler leak. It would seem unlikely that each of the three compressors has leaking intercoolers and some common cause of the problem should be sought. The problem should be solved by careful detective work by site personnel.

3.2.6. Moisture in Nitrogen Wash

'C' stream nitrogen wash unit has to be shut down every 15 days to be derimed after high pressure drop is observed. The source of moisture is assumed to be due to a leak in one of the reversing heat exchangers in 'C' stream ASU. Again this problem will need to be solved by careful detective work. There must be a suspicion that this problem is related to the problem with the nitrogen compressors.

3.2.7. Ammonia Refrigeration Compressor Performance

The ammonia refrigeration condensers are performing badly and consequently high compressor delivery pressures are needed in order to ensure condensation. In an attempt to reduce the delivery pressure, and machine power the suction of the machine is throttled from about 4 kg/cm^2g to 2 kg/cm^2g. This will cause the machine to be inefficient. Also the higher saturation temperature at 4 kg/cm^2g will cause a higher temperature of synthesis gas exit the synthesis gas cooler.

There are suggestions that the refrigeration compressor motors are underdesigned or that a new compressor is needed to cope with the load. These are incorrect. The problem lies with the performance of the refrigeration condensers and this is where the problem must be solved. Additional condensers have already been installed. However, it is apparent that in this plant cooling water system a majority of the pressure drop is through pipework rather than the coolers themselves. In such a system, addition of an extra cooler may not significantly increase the total water flow rate to the coolers. In fact the velocity in each cooler will be reduced causing a higher rate of cooling water fouling. It is therefore important that the reason for poor performance is diagnosed before spending money on modifications. It is recommended that :-

1) The performance of the condensers be monitored on a regular basis.
2) The actual performance is compared with the design performance.
3) If actual performance is not meeting design performance, the design of the condenser is checked.
4) If the original design area is satisfactory the condensers be cleaned. Ideally some condensers should be cleaned on the tube side, others on the shell side and others both sides, so that it can be identified which side is getting fouled.
5) If the condensers are fouling quickly then they should be redesigned to give a higher cooling water velocity to prevent fouling.

3.2.8. Leaky Steam Traps

Steam traps do not have an infinite life. As they get worn and start to leak significant quantities of steam, they must be replaced.

3.2.9. Gland Leakage from HPC Pumps

Potassium carbonate tends to crystallise in cold spots in the HPC system e.g. in the glands of pumps. If the pump is started with crystals in the gland, the gland will get scored and start to leak. Glands must be flushed thoroughly with clean water before starting in order to prevent crystallisation.

3.2.10 Cooling Water System

The cooling water system is a low pressure drop system with some parts using open coolers. Sludge conditioning was originally done using chromate treatment, but this has been changed to an organic-phosphate treatment because of pollution problems. At present there is no biocide treatment of the ammonia plant cooling water system.

With chromate sludge conditioning tube metal temperatures should be below 60°C whereas with organic-phosphate treatment they should be kept below 50°C. This means careful design of heat exchangers to ensure high cooling water velocities. Biocide treatment is very important in order to stop organic matter binding together mineral particles on the tube walls.

There is no flowsheet which shows how much cooling water goes to each cooler. When it is evident that some coolers require more water, a new line has been installed from a high pressure point in the cooling water system to the cooler. There is a danger that in doing this sort of modification, other parts of the cooling system may be starved of water.

Recommendations:

1. The cooling water flowrate to each heat exchanger should be determined. This can be done by injecting a radioactive tracer pulse at the pump suction and determining how long it takes the pulse to reach different points in each leg of the system. The flowrate can be determined from the volume of the system between measuring points.

2. It should be assessed what is the optimum distribution of cooling water and modifications made to improve the distribution. For example compressor intercoolers will perform quite well with a cooling water temperature rise of 15°C as long as the velocities are adequate whereas in condensers, the cooling water temperature rise must be kept low. Modifications would include installing orifice plates, installing larger lines, putting some exchangers in series instead of parallel and redesigning heat exchangers.
3. Control biological growth by shock doping with chlorine. Chlorine should be added once per shift over the period that the cooling water takes to go round the system and at a level to ensure 1 ppm free chlorine in the return cooling water.
4. Install on line cleaning eg Taproge in turbine condensers and air rousing nozzles between baffles in exchangers where cooling water is on the shell side.
5. Monitor cooling tower performance and check water distribution in the towers.

A five degree centigrade reduction in cooling water temperature is worth about 9×10^6 rupees/year.

3.3 Increase of Plant Capacity

The ammonia plant capacity is limited by some machines at a level below flowsheet rate. This causes an efficiency loss because most of the electric driven machines do not have any means of control to reduce power at part rate. If the plant rate can be increased to the flowsheet value, the extra power needed will be very small, so the electrical efficiency will improve. Proposals for increasing plant output are :-

New nitrogen compressor

New oxygen compressor

New refrigeration compressor

3.3.1 New Nitrogen Compressor

The three original nitrogen compressors were designed for a rate of 6300 NM^3/hr with a guarantee rate of 5900 NM^3/hr. On the guarantee test, they only achieved 5400 NM^3/hr and now achieve about 5300 NM^3/hr. A fourth compressor was installed during the expansion for a rate of 7150 NM^3/hr but only achieves 6900 NM^3/hr. The nitrogen requirement is 18810/hr at 100% rate. The plant cannot therefore be run above 93% rate without running all four nitrogen compressors. It would be worth running the fourth compressor in order to get an extra 7% production rate but if other limits prevent a rate much higher than 93%, it would be too inefficient to run the fourth nitrogen compressor.

It is proposed to install a new nitrogen compressor for 2100 NM^3/hr nitrogen. This can be easily justified compared to running the fourth large machine.

3.3.2 New Oxygen Compressor

There are three installed oxygen compressors each designed for a rate of 5300 NM^3/hr. The achieved rates are 4900 NM^3/hr. The oxygen requirement is 15675 NM^3/hr at 100% rate. The plant is therefore limited by oxygen compression at 94% rate. The

compressors can be modified to increase the capacity. This should be done in order to improve the plant efficiency and output.
There is a proposal to install a fourth spare oxygen compressor, so that the plant can continue to be run at full rate when an oxygen compressor is being inspected or overhauled. The fourth oxygen compressor will not significantly affect the efficiency and must be justified on extra plant output alone.

3.3.3 New Refrigeration Compressor

It has been proposed that a new refrigeration compressor should be installed because the synthesis gas temperature exit the cooler is high and there is a refrigeration limit. The synthesis loop pressure operates below the design operating pressure of 365 kg/cm^2 and well below a maximum operating pressure of about 380-390 kg/cm^2. Also the suction of the syn gas compressors is throttled. The production is not therefore limited by the synthesis loop and there is no need to put in an extra compressor. The synthesis gas temperature exit the cooler condenser could be high because the suction of the refrig compressor is throttled or simply because the loop is operating at lower than design pressure. This puts a higher proportion of the ammonia condensation load on the cooler/condenser rather than the water cooler. Additional compression need only be considered when the loop is operating at the maximum operating pressure for a significant proportion of the year.

Even then it would be cheaper to modify existing equipment either on the synthesis gas compressor, recycle compressor, rather than installing a new refrigeration compressor.

3.4 Other Efficiency Improvements

3.4.1 Synthesis Gas Compressor

Each stream normally operates below design rate, so the suction valves to the synthesis gas compressors are throttled in order to keep the make gas stream at design operating pressure. The synthesis gas compressors have clearance pockets which are always kept closed. The clearance pockets must be opened rather than throttling the suction of the compressor in order to save power on the compressor. When the clearance pocket is opened, the make gas pressure should be operated below the design value rather than throttling the suction to the machine. This will save power on the oxygen and nitrogen compressors.

3.4.2 Air Separation Units

When one stream is shut down due to failure of equipment in the stream, the ASU is kept running unless the stream is going to be off for some time. This is so that the cold box will not heat up and therefore delay restart of the stream. The time needed to recool the cold box can be greatly reduced by providing a storage of liquid nitrogen to cool the cold box. This will mean that the ASU can be economically shut down during shorter stream stoppages than at present and give a reduced power usage. Liquid nitrogen is produced in the ASU so it should be possible to bleed some of it to a liquid storage tank during normal operation, so that the tank can be refilled. It is recommended that the feasibility and economics of doing this should be examined.

3.4.3 Gas Turbines

It was proposed that gas turbines be used instead of electric motors to drive the largest machines. The cost of power from a gas turbine fired with naphtha will be more than double the cost of imported power. If fuel oil is used on the gas turbine, the cost of power will still be greater than the cost of imported power, so there is no case for spending the capital for gas turbine drives.

4. UREA PLANT

The Urea Plant was designed and installed by TEC Japan to produce 543.5 tes/day of urea in two 50% streams. Commercial production started on 1.1.69. The plant capacity was increased at the same time as the ammonia plant expansion. The existing streams were uprated from 272 tes/day to 300 tes/day and a third stream of 350 tes/day was added to give a total capacity of 950 tes/day.
The initial plant was the Mitsui Toatsu Recycle B process while the third stream was the MTC Recycle C process.
The efficiency of the Urea Plant is poor compared to design in the usage of ammonia, steam and power.

	Design	Achieved Performance		
		Good day	Good Month	*1986-7
Capacity te/day	950	808	671	570
NH_3 te/te	0.601	0.600	0.639	0.666
14 ATA Steam te/te	1.5	1.65	1.79	2.10
Power kwh/te	237	265	296	334
Variable Cost rupee/te	2887		3316	

* To February, 1987

The poor ammonia efficiency to urea represents an alarming loss of ammonia to the environment in gaseous and liquid effluents, and is equivalent to a loss of 58 x 10^6 rupees/year in terms of lost urea.

The main losses are due to losses in the ammonia absorber and in fume from the prilling towers. These problems can both be solved and must be progressed with greater urgency than is presently apparent.

4.1 Ammonia Recovery Column

The level of ammonia in the vent gas is 2.4% instead of the design figure of 1.9%. The temperature measured on the outside of the pipe feeding water to the top of the column indicates that the temperature is not significantly above design. The high ammonia slip is probably due to poor mass transfer performance of the column. This can be improved by changing to a more efficient packing and/or by changing to a better liquid distributor.
It is recommended that a thermocouple is installed in the

water feed line to the column and if the cooling water temperature is at design then a tower packing vendor eg Norton, be asked to recommend replacement packing and distributor for the column.

4.2 Ammonia Absorber

The ammonia absorber is not working as design and a large amount of ammonia is not absorbed. The inlet cooler has an exit temperature of 70°C instead of 50°C and this must be at least a partial cause of the problem. The cooler is co-current and is of an open type with water sprayed over the cooler tubes. The area of the cooler has been increased, but this did not significantly improve the cooling performance. Because the present cooler is co-current, the poor performance could be due to a low cooling water flowrate rather than a low surface area. A change to counter-current flow will improve the performance.

It is recommended that -

1) The present design of cooler is checked thermally.
2) The design is modified to eliminate the cooling problem. eg change to counter-current or use more cooling water or install more area.
3) If the ammonia absorber still does not perform satisfactory, the design of the absorber must be checked to determine the amount of mass transfer area required.

4.3 Fume from Prilling Towers

The fume scrubbing system on the prilling towers did not work well and has been removed. There is no plan for installing a new or modified system.

It is recommended that TEC be asked to design a new scrubbing system.

4.4 Conversion to ACES

It has been proposed that the existing A and B streams of the Urea Plant be converted from the MTC Recycle B process to the MTC ACES process. This would enable a higher urea production because of the better ammonia efficiency of the ACES process and would reduce the steam consumption of the plant. The design efficiency of the present A & B streams and the quoted efficiencies of some MTC processes are compared below.

	A&B Stream Design	MTC Process Quoted Figures		
		Recycle B	Recycle D	ACES
Ammonia te/te	0.61		0.57	0.57
Steam Import te/te				
26 ata	-			0.63
14 ata	1.68	1.5	0.78	-
Steam Export te/te				
6 ata	-	-	-	0.07
Power kwh/te	240	165	137	121

The design utility consumption figures for a particular plant are normally worse than the figures quoted by licensees. This is because the power usage does not include power for cooling water circulation and the steam consumption figure may not include steam needed for effluent treatment.

Differences between design figures and operating efficiencies are a function of solving technical problems on the plant, the rate at which the plant is run and how well the plant is operated and maintained. The differences are not likely to be affected by the process used.

The cost of treating the effluent from the Urea Plant will be more or less independent of whether the effluent treatment is added to the existing plant or a modified process.

Although the steam consumption of the ACES process is less than the Recycle D process, the pressure of steam needed is higher. About an extra 1.0 MW could be obtained by letting down 41 ata steam to 14 ata is a turboalternator for the Recycle D process, whereas it would not be worth putting steam through a turboalternator to let it down from 41 ata to 26 ata for the ACES process. The steam export from the ACES process of 6 ata steam is not worth anything if other mods are made to the steam system to supply the steam for the SGP deaerator.

Conversion of a Urea Plant from the Recycle B process to Recycle D has been done for the 200 te/day plant in the MTC Chiba factory. However, it is not clear that this conversion was entirely successful. Conversion of a Urea Plant from the Recycle C process to ACES has been done for the Korea Fertilizer Corporation 1000 te/day Urea Plant. Less work has to be done to convert a Recycle C process to Recycle D than convert it to ACES. Estimated costs for the ICI Billingham 1000 te/day MTC Recycle C Urea Plant were £2m to convert to Recycle D and £6m to convert to ACES.

It is recommended that :-

1) When considering whether to convert the current A&B streams to a new process, the utility figures that will actually be achieved on the modified plant must be used rather than figures quoted in publicity literature.
2) Whether to convert to the Recycle D process or to ACES must be considered carefully taking into account the capital cost, the value of energy improvements, the

import steam pressure required and how well proven the processes are.
The value of converting to a new process using process design figures and accounting for increased urea output and reduced steam and power consumption is about 47×10^6 rupees/year.

5. STEAM GENERATION PLANT

The boiler plant has 3 boilers each with a design capacity of 55 te/hr. The boilers never met the design guarantee and compensation was given by the boiler vendors.

	Guaranteed Design	Guarantee Test	Achieved 1986-87
Steam rate per boiler te/hr	45	35	
Coal te/te	0.144	0.149	0.185
Coal NCV kcal/kg	5320	5900	5000
Coal at 5320 kcal/kg	0.144	0.165	0.174
Efficiency LHV	88.5	77.2	73.2
Power kwh/te	12		17.3
LP steam	0.12-0.13		0.16-0.20
Variable cost rupee/te	96.8		120

The achieved coal efficiency is very poor. It is recommended that the boiler is monitored to find the cause of the poor efficiency with a view to making modifications to improve it.

5.1 Loss Accounting

The difference between the actual and design LP steam and power usages are not accounted for by the plant operating team. It is recommended that this be done.

5.2 Operation on two boilers instead of three

The power usage is high compared to design because three boilers are normally run instead of two. The normal steam rate of 90 te/hr is well within the design capacity of two boilers of £110 te/hr. There are problems with ash disposal at rates greater than 40 te/hr per boiler. Also the steam temperature rises as the rate on the boilers is increased. Normal operation of two boilers instead of three would save about 1.9×10^6 rupees/year. It is recommended that :-

1) The ash disposal system is modified to allow for higher ash rate at higher boiler rate.
2) The maximum allowable superheat temperature should be determined. This could be significantly higher than the design operating temperature of 413°C.
3) The reason for the higher than design superheat temperature at high rates should be investigated. For example, it could simply be that the boiler coils are fouled and need cleaning.

5.3 LP Steam Usage

The LP steam usage for dearation is much greater than the theoretical value. It is recommended that checks are made on the deaerator to see whether excess steam is being vented and on the steam line lagging to see whether the insulation is intact.

5.4 Supply of heat for deaeration.

Steam to supply some of the deaeration heat is let down directly from the HP level. This is inefficient. There are two ways of improving this situation. It may be worth doing just one or both of them.

1. Use steam turbine driven BFW pump.
   There are two electric BFW pumps and one emergency steam turbine pump each with a design rate of 130 te/hr. An electric BFW pump is normally run at a power of 350 kw. It is recommended that either the emergency steam turbine pump is run continuously or if this is too inefficient or too unreliable a new steam turbine is installed to operate from 41 ata to 3.5 ata. The annual saving from using a steam turbine will be 2.7×10^6 rupees/year.

2. Send Urea Plant Condensate to SGP
   Hot process condensate from the Urea Plant is sent to the ammonia plants. This means that no demin water heating has to be done on the ammonia plants. The SGP plant is supplied with cold demin water, which has to be heated by steam in the deaerator. It would be more efficient to send the hot process condensate from the Urea Plant to the SGP and to preheat demin water on the ammonia plant with low grade heat. For example, the regenerator overheads could be used to heat demin water, thus reducing the heat load on the overheads condenser.

5.5 Installation of Turboalternator

About 60 te/hr of steam is let down from 41 ata to the 14 ata steam main. About 3.3 MW power could be recovered if the steam from the gasification waste heat boiler were superheated and all the let down steam put through a turboalternator.

If all three boilers were run at 45 te/hr, there would be about 45 te/hr of excess steam that could be used to generate power in a turboalternator. At a condensing pressure of 140 mbar about 10 MW could be generated from this steam.

The rough economics of the scheme would be :-

	10^6 rupees/year
Saving in import power of 13.3 mw	100
Cost of extra steam raising on boiler	40
LSHS firing for superheater	8
Total variable cost saving	52
Capital cost of turboalternator and superheater	approx. 90 x 10^6 rupees

This would give a payback time of less than two years. The economics will be affected by boiler availability. However, if boiler maintenance is done when one ammonia/urea stream is shut down and the two remaining boilers fired up to 50 te/hr each, then the condensing rate only falls to 40 te/hr, so the power reduction will be small. If the urea plant A&B streams are converted to MTC Recycle D process the available power from a turboalternator will increase to 17 MW.

If the Urea Plant A&B streams are converted to MTC ACES process, it will not be worthwhile having a steam passout at 26 ata, so the available power from the alternator will be 16.8 MW, with a condensing rate of 71 te/hr. However a new superheater should not be necessary.

It is recommended that the case for installing a turboalternator is looked at in more detail, whatever is decided about the Urea Plant improvements.

VS21 12th May, 1987