International
Fertilizer Industry
Association

IFA Technical Conference

Amman, Jordan

2-6 October 1994

International Fertilizer Industry Association - Secretariat: 28 rue Marbeuf - 75008 Paris - France
Tel. +33 1 53 93 05 00 - Fax +33 1 53 93 05 45/47 - ifa@fertilizer.org - www.fertilizer.org

IMPROVING PRODUCTIVITY IN AMMONIA AND UREA PLANTS AT GNFC

G.K. Parikh and K.M. Jani
Gujarat Narmada Valley Fertilizers Company Limited, India

RESUME

Le taux d'utilisation de la capacité des unités de GNFC était faible à l'origine pour plusieurs raisons. Les obstacles limitant la productivite de l'unité ont été identifiés et des modifications ont été apportées au cours des années. Différentes mesures ont été mises en oeuvre pour améliorer la performance de l'unité. Les mesures suivantes ont été prises pour améliorer la productivité de l'unité.

◆ ◆ ◆ ◆

INTRODUCTION

Productivity has earned a very reputable place in almost all the industries because it determines the growth of an enterprise. This has become more relevant in present day context in view of tough competition in the market, changing government policies and rapid change in the technology.

The multi-dimensional view of productivity can be related to achievement in the following areas:

- Production levels
- Reduction in downtime hours/delays
- Start-up and shut-down frequency-
- Reliability of equipment and machinery
- Energy conservation measures
- Safety and pollution control
- Training of personnel
- Reduction of input costs
- Import substitution
- Development and alertness in regard to internal and external service
- Manpower development and improving effectiveness and motivation of people

Gujarat Narmada Valley Fertilizers Co. Ltd (GNFC) has the world's largest single stream ammonia-urea complex located in the industrially backward district of Bharuch. The company was established in 1976 and trial runs of ammonia-urea plants started in January 1982. Commercial production commenced from 1st July 1982. From a fertilizer company, it has diversified into chemical, electronics and engineering industry. It has captive power plants for uninterrupted power supply.

It is a well known fact that the technology of producing ammonia from fuel oil is a difficult process involving quite a different approach compared to steam reforming plants. We had selected Ms. Texaco's basic process of partial oxidation of fuel oil for the production of ammonia. During the initial phase, the capacity utilization of plant remained low because of various reasons like failure of tubes in the cooling water heat exchangers, failure of gasifier quench ring, failure of ammonia condenser tubes, frequent tripping of plants on account of spurious trips, power failure, etc. Plant reliability increased after 1985, registering high on-stream factor and increased capacity utilization.

The average daily production of ammonia started registering 105% after 1985 and slowly we could identify the bottlenecks in the plant. With the built-up of reliability, we carried out certain modifications which boosted the plant capacity further to 120% and even more thereby registering the ever highest daily production of 1718 MT/day on 23.01.1993 equivalent to the plant load of 127%. The high daily average product compensated for various interruptions caused by basic design problems of short quench ring life and high pressure drop across the 1st CO shift reactor.

There was also a marked improvement in yearly production figures and we achieved in 1991-92 ever highest yearly capacity utilization of 114.48% of ammonia and 121% of urea.

With higher capacity utilization year-by-year, the energy conservation per MT of ammonia was also reduced from 14 MKCal to 12.08 MKCal in the ammonia plant and in the urea plant from 9 MKCal to 8 MKCal.

To cultivate good habits and to inculcate creativity and innovation amongst the personnel, various techniques such as management by objective (MBO) were followed along with effective suggestion scheme to improve the plant productivity.

PRODUCTION: STEP BY STEP APPROACH

In order to understand productivity improvement in ammonia plant step by step, the entire operation period is divided into the following three major segments:

1. 1982-85 : initial stabilization process
2. 1985-88 : reliability build-up
3. 1989 onwards : performance improvement

A. 1982-85: Initial stabilization process:

Initial few years were full of anxieties because this was a unique plant operation with high pressure gasification system, having its own problems. The process stabilisation took its own time because there were certain design deficiencies. Various failures of equipment were also encountered. We faced tube leakage of cooling water exchangers of synthesis loop which in turn was responsible for creating many other exchangers to fail because of two phase flow on cooling water side. Air-compressor rotor failure, gasifier quench ring and refractory failure were also quite prematured resulting into heavy production loss during initial phase of plant operation.

We had to order new cooling water exchanger in synthesis loop with modified design and had consciously taken the decision to replace tubes of various heat exchanges to stainless steel. This has improved the plant reliability for the years to come and paid us rich dividends thereby increasing plant production outputs on annual basis.

B. 1985-89: Reliability build-up:

Phase-wise replacement of cooling water heat exchangers' tubes continued. We retubed our surface condensers also with stainless steel which gave us advantage on reliability and improving the energy efficiency of the plant. The slogan for the period "100% ammonia and 100% urea" boosted us to improve plant reliability and identify to keep both the gasifiers in line and various innovative maintenance techniques were adopted. On-line maintenance techniques for avoiding shutdowns were adopted and maximum clubbing of jobs were done which resulted in reduction in number of NIL production days.

Clear production targets and goals were fixed during the start of the year including fixation of specific consumption norms. Various action plans were targetted and day-to-day monitoring of plant parameters was actively carried out by a separate performance monitoring group.

With this approach, plant productivity improved greatly and the plant started running at maximum loads indicated by maximum loop pressure in synthesis section. At this stage itself, the average daily production of ammonia plant reached about 108%. Various techniques were tried to reduce synthesis loop pressure and this further improved the plant productivity. The details for which are separately discussed later.

C. 1988 onwards: Performance improvement:

The major limitations of the plant to achieve capacity utilization above 105% were identified and are discussed as follows:

1. Quench ring failure:

Right from the beginning, we experienced short life span of quench ring used inside the gasifiers. Problems were referred to M/s. Texaco and they had also shown their concern for trying various methods to improve the life span of the quench ring. Almost all other plants have the life span of quench ring varying from 4000 to 6000 hours. The main reason for the failure of quench ring was attributed to nickel sulphiding attack due to high temperature. M/s. Texaco has put forth several reasons for short life of quench ring at our plant and suggested various designs to be tried in case of our plant. The maximum life obtained with their suggestions was 2500 to 3000 hrs. However, we could not get better solutions for improving life of quench ring.

Based on our operating experience and data collected from other operating plants, we came to the conclusion that short life span of our quench ring was mainly due to high gas mass velocity at the quench ring area. We worked out new gasifier quench ring diameter and suggested to M/s. Texaco for incorporating a modified larger diameter quench ring. After going into the details, they suggested new shape of quench ring which was tried in their Louisiana plant. The new quench ring is of half circular shape having uniform cross section throughout. The modified design quench ring was tried in 1991 in one of our gasifiers and we could extend the operating life to more than 9 months. Details of the modified quench ring is shown in schematic diagram at **Annexure I.**

2. High synthesis loop pressure:

Ammonia plant synthesis loop had limitation beyond 105% load. Right from the beginning, the loop pressure remained as high as 240 bar as against 220 bar with this plant load. Our studies were basically concentrated on the following areas:

a) Pressure drop survey in synthesis compressor and ammonia synthesis loop

b) Possibility of intermixing of gas in hot heat exchanger of synthesis loop.

c) Temperature drop across reactor remained high indicating lower circulation. Installation of modified larger circulator and improvement in condensation of ammonia by cooling water and ammonia refrigeration system.

d) Close monitoring of H_2/N_2 ration in the loop by switching over the electronic control loop.

e) Installation of S-200 series basket.

(A detailed chronology is as per **Annexure II**)

a) Pressure drop survey/removal of recycle stage strainer:

Pressure drop of 3 bar across recycle stage (3rd stage of synthesis machine) was identified. In consultation with M/s. Bhel and our maintenance department, we removed the internals of the strainer which resulted into lower pressure drop of the loop. This resulted in increase in plant production from 1440 MTPD to 1500 MTPD on continuous basis.

b) Short-circuit in heat exchanger:

Internal passing of gas from shell to tube side of heat exchanger was identified with various analysis of gas stream. The heat exchange was having a typical design having gland system between shell and tube. This leaking gland resulted into by-passing of gas. We adopted a method of injecting furmanite compound in the gland. This resulted in sustained production of more than 1500 MTPD on prolonged basis. We still are on the look-out for permanent solution to this exchanger.

c) Larger recycle stage impeller:

Ammonia synthesis loop was designed for circulation load of 6,94,000 NM^3 of syngas. To get maximum advantage, we had locked open the guide vanes of circulation. However, circulation remained low. Instead of overall temperature difference of 189°C across converter, a difference of 210°C was obtained. Possibility of increasing circulation flow was examined in consultation with M/s. Bhel. After detailed study, it was found that existing impeller vane diameter could be increased to 410 mm as against existing 370 mm. A larger size impeller was installed in the circulator stage during September 1989. This gave us an increase throughput of about 80,000 NM^3/hour in the circulation flow. Synthesis loop pressure came down from 240 bar to 225 bar. The plant load was increased to as high as 112% with this innovative solution. These details are as per **Annexure III.**

d) Electronic control loop to H_2/N_2 ratio:

Variation in the H_2/N_2 ratio in the syngas loop is overcome very accurately by installation of electronic control loop. With this, loop pressure is being maintained fairly constant which improved operations of the synthesis loop on continuous basis.

e) Revamping of synthesis converter:

In order to optimize further production level at the highest possible level with minimum energy input, detailed studies were carried out for synthesis reactor basket revamp to S-200 series. S-200 basket was ordered and installation was planned during April 1992 shut-down. Exhaustive planning and effective execution led us to complete total revamping job in 20 days.

With the installation of S-200 basket and successful commissioning, the loop pressure problem in our ammonia plant got entirely eliminated. The loop pressure greatly came down, thereby saving in energy and improving plant load. We produced ever highest daily production of 1718 MT on 23.01.1993 after installation of S-200 basket. The detail drawings and comparison of S-100 and S-200 series baskets are as per **Annexures IV and V.**

3. <u>Increased CO_2 production as per increased demand by other plants</u>:

Our rectisol wash unit is designed to produce approx. 30,000 NM^3/hour of pure CO_2 corresponding to design requirement of urea plant. When only one gasifier is in operation, provision is made to run urea plant at 100% load.

With the increase in ammonia production, CO_2 production also increased. However, urea plant load was unable to increase beyond 100% because of CO_2 compressor drive turbine limitation. Steam pressure was increased but that gave us marginal improvement.

The design pressure for CO_2 compressor suction at urea plant was 1.55 bar. We increased CO_2 compression suction pressure in order to increase the load. The suction pressure was gradually increased to 2.4 bar g which increased urea plant load to 134%. This rise in pressure of CO_2 had not affected the operation of CO_2 tower. The total CO_2 production increased to 40,000 NM^3/hour by making small innovative changes in the rectisol wash unit.

Higher plant loads with minimum methanol circulation

a) Rectisol wash unit is designed to circulate 270 m^3/hour of methanol for 100% plant load. With close monitoring and study, we decreased the total circulation to 230 m^3/hour at 123% of plant load.

b) **Demand of CO_2 tail gas by other plants:**

Nitrophosphate plant methanol-II plant were commissioned in 1990. Both these plants utilize pure CO_2 and tail-gas for producing nitrophosphate and higher methanol production (NPU = 3500 NM^3/Hour at 125% load, methanol = 1800 NM^3/hour. We are always running short of CO_2 because of higher demands. We have future plans to increase CO_2 production from tail-gas. Presently, all the above plants are running at maximum load based on high production of CO_2 from ammonia plant.

4. **High pressure drop on shell side of spiral wound heat exchangers:**

There are many spiral wound heat exchangers in rectisol wash unit. The heat exchangers are of Linde design. During 1986-87, we noticed severe fouling in three exchangers on shell side. The pressure drop was measured which was high enough for creating problems in methanol circulation. Following actions were taken for sustaining high load operations till new heat exchangers were ordered:

1. Some of the pump impellers were replaced with higher diameter impellers depending upon the margin in the motor/pump casing.

2. Chemical cleaning with 5% HCl was carried out during shutdown opportunity.

3. Smaller sized strainers including one micro-filter was placed in methanol circuit for arresting deposits.

All above measures worked well till we got new heat exchangers.

5. **Formic acid plant based on tail-gas from LNW:**

We have put up 5000 TPA of formic acid manufacturing facility as a part of our innovative corporate philosophy. The tail-gas before going to steam superheater is routed to formic acid plant. The profitability of this plant is good in view of the market demands. With increase in ammonia plant load, formic acid plant capacity was also increased to 170% with many innovations and modifications.

6. **High pressure drop across co shift converter:**

Right from the beginning of plant operations, we have been facing problem of high pressure drop across 1st CO shift reactor R-401. With the increase in plant load, the problem was further aggravated since that the frequency of catalyst removal increased because of high throughput of gas.

Because of various metallic impurities in oil and carry-over of soot, the reactor pressure drop increased with 4-6 months of operations. Catalysts are pyrophoric in nature and they have to be kept under nitrogen atmosphere once they are in sulphide stage. We have worked out revamp scheme to solve this problem on permanent basis but till then we have been employing an innovative technique of reuse of the catalyst.

The catalyst used is of BASF K8-11 CoMo which is very expensive. We remove the catalyst from the reactor, screen it and reuse. We make-up balance quantity with fresh catalyst. This give us lot of saving on catalyst cost.

ENERGY CONSERVATION SCHEMES

As a part of our innovative approach with the increase of production and by introducing various modifications in the plant, the energy conservation per tonne of ammonia has substantially reduced. These figures are even comparable with some of the older steam reforming plants so far as the energy consumption is concerned. Some of the energy conservation schemes are described hereunder briefly. A summary of schemes and their savings is as per **Annexure VI.**

1. Ratio control of gasifier:

   In order to have better combustion efficiency in the gasifiers, we had introduced an automatic O_2 : oil ratio controller which improves combustion thereby saving in O_2 consumption. Scheme saves about 1.4 MT steam/hour which is equivalent to 0.35 MKcal/hour.

2. Use of LSHS along with FO:

   We had been using furnace oil which contained 3.5% sulphur. On account of high sulphur, we were consuming more than 0.75 T of oil per tonne of ammonia. After 1983, we started using low sulphur oil which reduced our specific consumption to 0.73 T/T. This saved about 0.2 MKcal/tonne of ammonia.

3. High conversion in CO shift reactor by water injection:

   We have three units of CO shift conversion. The design CO slippage from 3rd bed is 1.5% 3rd bed inlet temperature was reduced from 322°C to 280°C and CO slippage was reduced to 0.6% by water injection upstream of 3rd reactor. The energy was 0.08 M Cal/tonne.

4. Energy saving in synthesis loop:

   As described earlier, the saving of energy is at the synthesis loop on account of the following three schemes:

- Increase of vane diameter of recirculator	0.6 M Kcal/hour
- By-passing in gas-gas exchanger	0.12 M Kcal/hour
- Higher pressure drop across strainer i.e. steam saving	0.5 Tonne/hour
- S-200 synthesis catalyst basket	0.02 M Kcal/Tonne

5. Conversion of ammonia condenser from 4 to 2 tube passes:

   The condenser performance was not satisfactory and its cooling water approach temperature remained at 12°C instead of 10°C design. This was one of the major bottlenecks to achieve 100% capacity utilization. These exchangers were having 4 tube passes on cooling water side which was converted to 2 tube passes thereby reducing pressure drop and increase in cooling flow rate. Heat transfer increased by about 25%. Final discharge pressure of refrigeration machine also reduced by 2 bar and hence saving in steam consumption.

6. Saving in electrical energy:

   a. Trimming of impellers:

      We had identified some of the pumps delivering higher head than required. We trimmed all these impellers and reduced power consumption. Details are given in Annexure VII.

   b. Cooling water pumps:

      Initially, six cooling water pumps were required to be run as per design. This was changed to five without affecting the performance.

c. Turbine condensate recycling to dearator:

This was reviewed at length and after change of condenser tubes to SS, turbine condensate was diverted to dearator saving polishing and pumping costs.

PRODUCTIVITY IMPROVEMENT IN UREA PLANT

1. Reliability and operational improvements in CO_2 compressor:

a. Increase in CO_2 suction pressure:

The load of urea plant was limited on account of limitation in throughput of compressor at the design suction pressure which was increased by operating CO_2 stripping column T-502 in ammonia plant at a higher pressure.

Also, we found that the final discharge line NRV was offering very high pressure drop of 5 kg/cm^2 g due to mechanical limitation. The same was properly attended and the final discharge pressure reduced by 4.5 kg/cm^2 thereby improving the performance of CO_2 compressor at higher loading.

b. Single antisurge system:

Earlier between MCL and BCL stage, there was a sulphur/methanol removal section and hence two different antisurge valves were provided in the original machine. This created a lot of problems during abnormalities/trips as well as start-up of the machine as it was difficult to synchronize both the antisurge valve operations. On a number of occasions, the machine surged leading to heavy down time and production loss. The two antisurge system was replaced by a single antisurge system (from 4th stage discharge to 1st stage suction) and after that the problem of surging has been virtually overcome. **(Annexure VIII)**.

c. Replacement of interstage coolers:

Earlier the interstage coolers were of carbon steel tubes which were leaking frequently resulting in loss of urea production. We replaced all the tube bundles in the three interstage coolers with SS 304 tubes.

2. Modified urea stripper:

Our original stripper was reversed during September 1986 shutdown to extend its life. The steam entry was changed from top to bottom but this resulted in poor performance of stripper beyond 80% load. This was because the condensate removal from the bottom of shell side became reduced and the MP process steam could be introduced at a rate not more than 60 MT/hour as against 76 MT/hour design value. A 4" line was laid from the stripper top to the down stream equipment E-2 taking the condensate out from the top of stripper, thereby regaining the original urea plant load.

During April 1990 shutdown, we replaced the stripper with a new reversible type stripper. However, even with new stripper, we faced the same problem of condensate removal from the bottom. Again the 4" line from the top was useful in increasing the load on stripper while maintaining the process outlet temperature at 208°C.

3. Replacement of carbamate condensers:

Four of our Snamprogetti urea plants had two carbamate condensers with a total area of 3400 m^2 and with SS 316L tubes. We encountered tube leakage problems in the first carbamate condenser. The flow reversal on the process side was tried to increase the life.

The same was repeated in the second condenser as well. In the meantime, we ordered a new carbamate condenser with 2 RE 69 material for the tubes. A single condenser was installed during April 1992 shut-down.

4. Scaling problem in MP condenser E-7:

Our cooling water treatment was based on SHMP/$ZnSO_4$/organophosphonate in the form of a formulation prepared by our own laboratory. Due to the higher temperature in this exchanger, the rate of scale formation was high and the performance deteriorated after 2-3 months of operation leading to more load on LP section. Of late, we have started using M/s. Chembond treatment for urea cooling tower. The treatment chemicals are based on glassy phosphates and the performance of MP decomposer has improved to a great extent. We are able to maintain the process outlet temperature at about 88°C even after six months of operation. The approach temperature is also steady indicating reduced scaling in this exchanger.

5. Utilization of flash steam of condensate tank:

All the condensate in urea plant is returned to condensate tank, V-2 and flashes into steam at low pressure. The steam becomes condensed by the water cooler which was of C.S. material and this exchanger leaked. It was decided to isolate and later on removed this exchanger so that the flash steam of about 7 MT/hour could be utilized for heating DM water. A 4" line was laid from condensate tank to a drum in ammonia plant so that this steam could be utilized for heating DM + ACT in the heat exchanger E-1312 of ammonia plant deaerator system. The energy saving on account of this modification is 4.5 MKCal/hour which on annual basis works out to Rs. 1,280,000. The cost of the scheme was Rs. 500,000 and hence, payout period is nearly 5 months (**Annexure IX**).

6. Additional cooling tower cell:

To compensate for the heat removal at increased loads of plant, it was identified that one additional cooling tower cell was a must. This was executed and results are encouraging.

FUTURE PLANNING AND CONCLUSION

We believe that only dynamic management system can produce good results whether it is related to financial management or productivity management. To give great emphasis towards plant reliability and improving the productivity, we have considered following revamp programmes:

a. Revamping of air separation unit for increasing O_2 output and installation of molecular sieve instead of present revex system for plant reliability.

b. Installation of vertical decanter for better separation efficiency to remove heavy metals from soot water.

c. Rearrangement of 3rd bed CO shift converter and installing it as a parallel reactor to 1st bed reactor for decreasing problem of higher pressure drop.

d. Phase-wise replacement of spiral wound exchangers of rectisol wash unit.

e. Phase-wise replacement of pneumatical control system to electronic control system to improve reliability of operations.

f. Revamping of cooling tower system by introducing one more cell to bring down cooling water temperature which is a major limitation for increased plant load during summer time.

g. Introducing scheme in rectisol wash unit to increase the production of carbon dioxide.

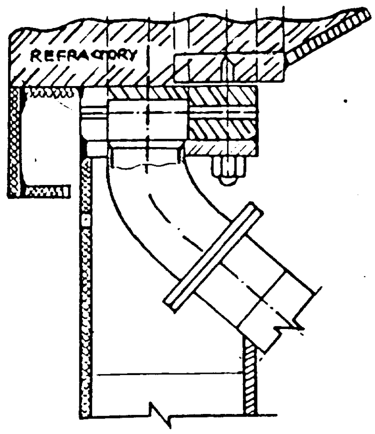
All above revamping schemes are actively of implementation stage.

CONCLUSION

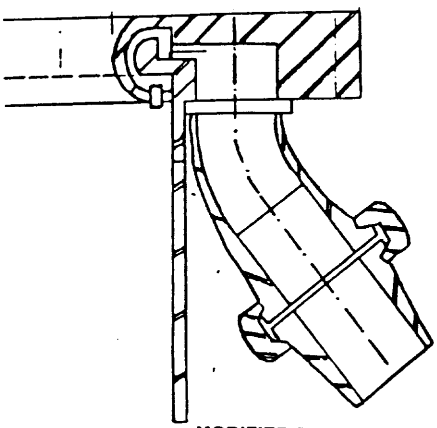
We have achieved continuous running of ammonia plant with increased capacity from 1350 to 1650 MTPD. Similarly, the urea plant has also been stabilized to operate at 125% load producing 2250 MTPD of urea. Achievements in such complex plants have been effected with great care so as not to exceed safe limits of any equipment or operating machineries. This has imparted lot of confidence in us about operations and has taught us to analyze for better productivity continuously.

ANNEXURE I

QUENCH RING



ORIGINAL DESIGN



MODIFIED DESIGN

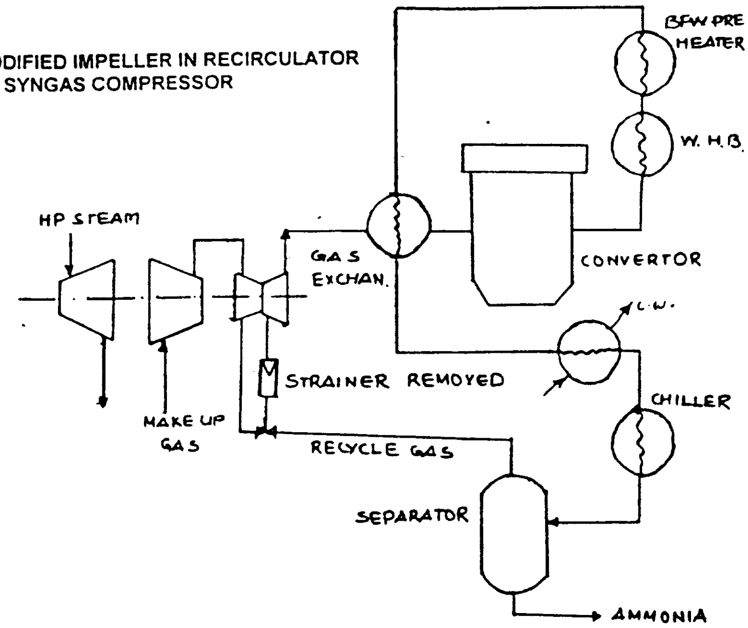
ANNEXURE II

CHRONOLOGY OF SYNTHESIS LOOP MODIFICATION

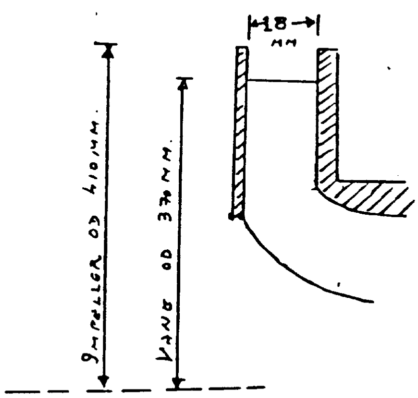
1.	Start-up heater (E-710) replaced since old one failed (coils)	February	'84
2.	New interstage cooler (E-709) with single pass on cooling water side installed.	June	'84
3.	New interstage cooler (E-701) tube bundle replaced.	July	'84
4.	Water cooler (E-706) changed over to new desiegn (split form).	July	'84
5.	W.H.B.E.-703 dome cover repaired and new tube bundle provided.	June	'85
6.	HP BFW preheater (E-704) replaced to new type.	June	'85
7.	Suction strainer of circulator (C-701) 3rd stage removed.	April	'85
8.	Furmaniting of hot gas/gas exchanger (E-701) adopted.	June	'88
9.	Modified circulator installed.	September	'89
10.	New interstage cooler (E-702) tube bundle replaced.	December	'92
11.	S-200 series basket/new catalyst installation.	April	'93

ANNEXURE III

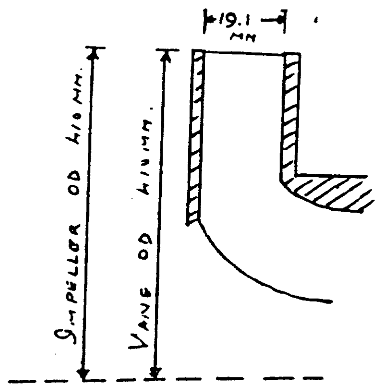
MODIFIED IMPELLER IN RECIRCULATOR OF SYNGAS COMPRESSOR



ORIGINAL RECYCLE IMPELLER



MODIFIED RECYCLE IMPELLER



ANNEXURE V

ENERGY SAVING AND COST BENEFITS OF S-200 ENERGY SAVINGS WITH DESIGN CAPACITY OF 1350 MT/DAY

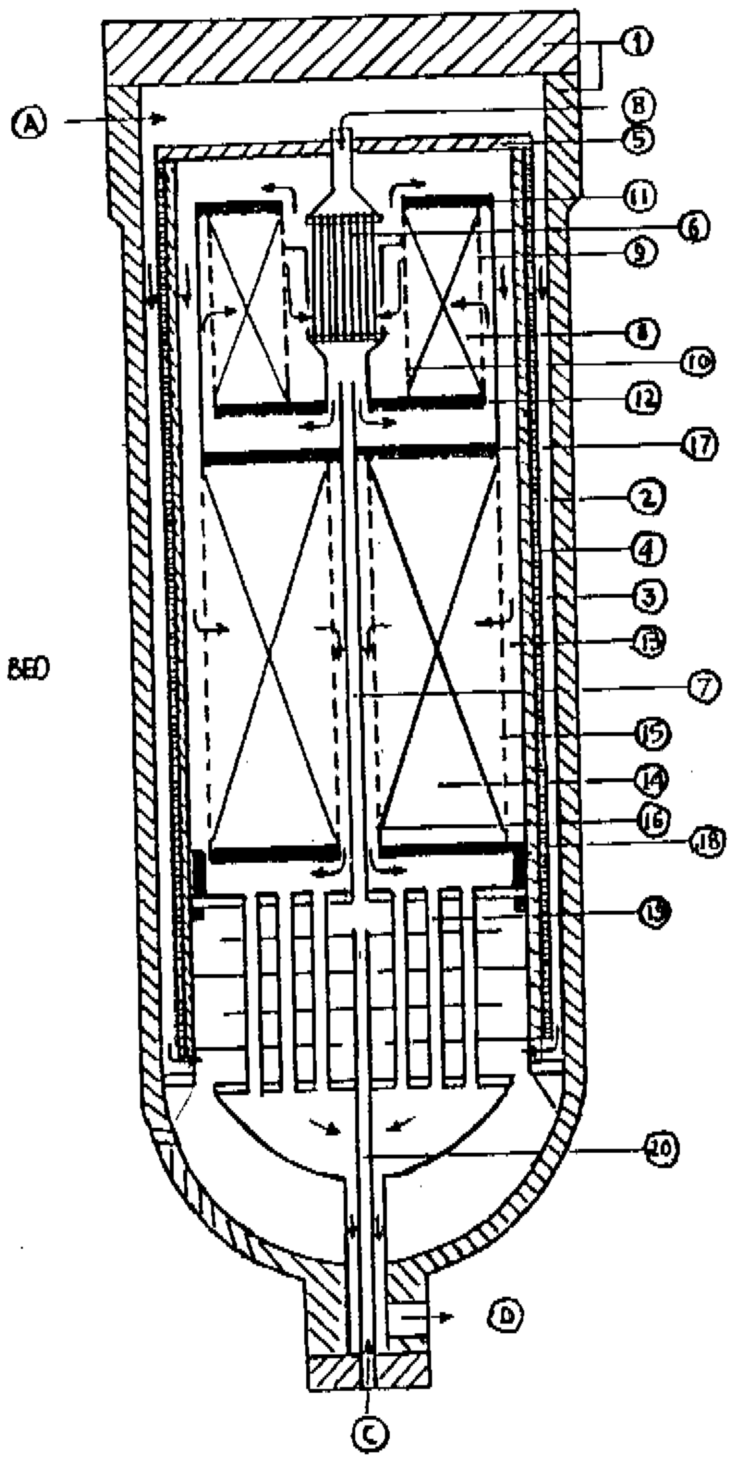
CONVERTER CONDITIONS	S-100	S-200
Inlet pressure kg/cm²	229.5	214
Inlet temperature °C	240	228
Outlet temperature °C	429	430
Inlet flow NM³/hour	694,436	638,612
Inlet NH_3 concn. %	3.85	4.37
Outlet NH_3 concn. %	16.24	18.11
Loop pressure drop kg/cm²	13.2	11.9
Compression power - KW	10,524	9,730
Refrigeration power - KW	2,921	2,122
Waste heat recovery equivalent KW	11,448	11,434
Net power consumption - KW	1,997	418

Energy saving = 1579 KW = 37.896 NWH/Day = 1,56,32,100 Rs./Annum (Rs. 1.25/KWH x 330 days) Pay off period is approx. three years.

SERIES 200 AMMONIA SYNTHESIS CONVERTER
TYPE WITH LOWER HEAT EXCHANGER

A : SHELL COOLING INLET
B : INLET FOR GAS TO INTERBED EXCHANGER TUBE SIDE
C : COLD SHOT INLET
D : GAS OUTLET

1 : PRESSURE SHELL
2 : OUTER ANNULUS
3 : OUTER BASKET SHELL
4 : BASKET INSULATION
5 : BASKET TOP COVER
6 : INTERBED HEAT EXCHANGER
7 : TRANSFER PIPE
8 : 1ST CATALYST BED
9 : OUTER CATALYST BED WALL
10 : CENTER TUBE 1ST CATALYST BED
11 : COVER PLATE
12 : CATALYST SUPPORT
13 : ANNULUS AROUND THE CATALYST BED
14 : 2ND CATALYST BED
15 : OUTER CATALYST BED WALL
16 : CENTER TUBE 2ND CATALYST BED
17 : COVER PLATE
18 : CATALYST SUPPORT
19 : LOWER HEAT EXCHANGER
20 : COLD BY-PASS PIPE



ANNEXURE VI

BRIEF SUMMARY OF ENERGY CONSERVATION SCHEMES

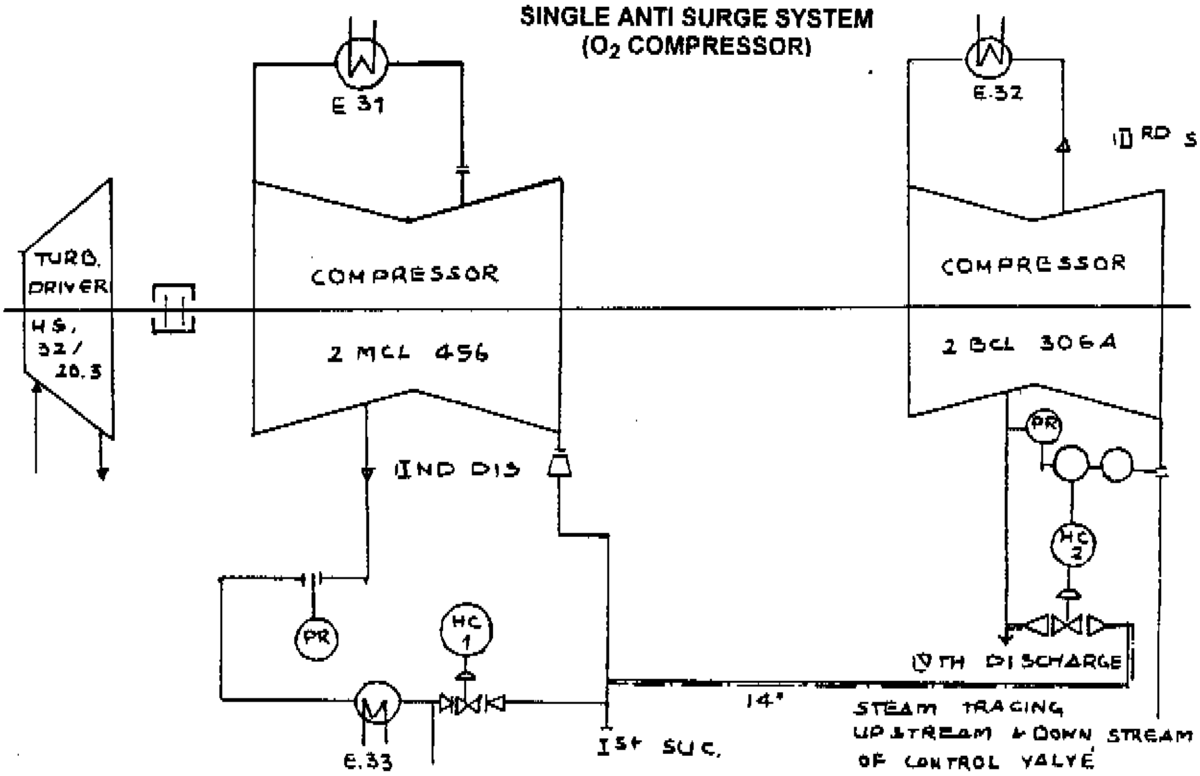
Sr. No.	Description		Energy MKCal/Hr.	Saving MKCal/Hr.	Annual Saving Rs.lacs	Cost Benefit Ratio
1.	Oxygen oil ratio controller	(1.4 MT/Hr HP steam)	0.350	0.006	16.800	1 month
2.	Modification in Soot Water Cooler	(4 MT/Hr LP steam)	2.000	0.035	16.800	1 month
3.	Higher Conversion in CO-Shift reactor by water injection		4.500	0.080	56.000	Cost of scheme is negligible
4.	High pressure drop across Strainer	(0.5 MT/Hr. HP steam)	0.125	0.002	6.600	Cost of scheme is negligible
5.	Series 200 Catalyst Basket (Advance due to fresh catalyst not considered)	(4.5 MT/Hr. HP steam)	1.125	0.020	56.000	
6.	E-704 (old) for catalyst heating of CO shift converter	LPG fuel saved Reduce start-up time	-.-	--	5.000	1 year
7.	Use of flash steam of Urea plant in Deaerator	(7 MT/Hr. Flash steam)	3.000	0.050	12.800	5 months
8.	Reduction in Electric Power consumption		1.032 (1200 KW)	0.018	96.000	Cost of scheme is negligible
9.	Various Reliability & Continuity of Operation Measures			2.279		
10	TOTAL ENERGY SAVING COMPARED TO FIRST THREE YEARS OF OPERATION.	(14 MKCAL→11.51 MKCAL)		2.490 MKCAL/MT		

ANNEXURE VII SAVING IN ELECTRICAL ENERGY

Description of pump	Original Data Flow M^3/Hr	Impeller Dia MM	Power Cons. KW	Head MLC	After Modification Flow M^3/Hr	Impeller Dia. MM	Power Cons. KW	Head MLC	Power Saving KW	Remarks
Condensate	150	321	159	220	150	291	130	180	29	
Hot condensate	150	348	83	175	150	320	70.2	150	12.8	Pump replaced by new one of 40KWH Poweer Cons.
C/X-102 Cond.	82.1	258	57	155.3	82.1	202	35	146	22	
C/X-701 Cond.	60.2	250	45	158	60.2	203	27	95	18	
C/X-101 Cond.	102	280	63	158	102	NA	38	95	25	Three impellers were removed.
C/X-1101 Cond.	48	280	49.2	158	48	NA	29	95	20.2	-do-
C.W. Supply	600	310	105	35	600	300	98.3	33	6.7 x 2	
D.M. Transfer	125	306	60	128	125	294	55.4	118	4.4 x 2	
Treated Amm.Cond.	186	350	125	102.6	186	320	104.5	86	20.5 x 2	
Treated Urea Cond.	36.3	243	25	82.9	36.3	233	23	76	2 x 2	
Amm.C.W. Pump									960	Stoppage of one C.W. pump
General Service (CPP)									80	Stoppage of GSW Pump
TOTAL POWER SAVING									1,234.6	

ANNEXURE VIII

SINGLE ANTI SURGE SYSTEM
(O_2 COMPRESSOR)



ANNEXURE IX

UTILISATION OF FLASH STEAM OF CONDENSATE TANK IN UREA PLANT