

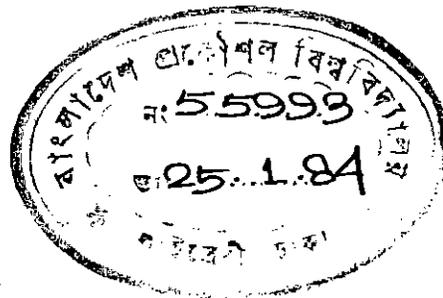
TECHNOLOGICAL CONSTRAINTS IN  
UREA PRODUCTION IN BANGLADESH

A Thesis

by

MD. MIZANUR RAHMAN

Submitted to the Department of Industrial & Production  
Engineering, Bangladesh University of Engineering and  
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ments for the degree of  
MASTER OF SCIENCE IN INDUSTRIAL & PRODUCTION ENGINEERING



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BANGLADESH UNIVERSITY OF ENGINEERING AND TECHNOLOGY, DHAKA

CERTIFICATE

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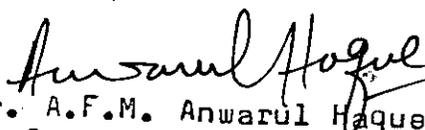
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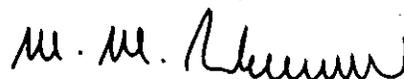
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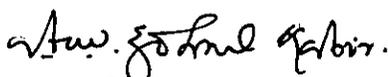
Dr. M. Anwarul Azim  
Professor & Dean,  
Faculty of Mech. Engg., BUET, Dhaka,  
Chairman of the Committee.



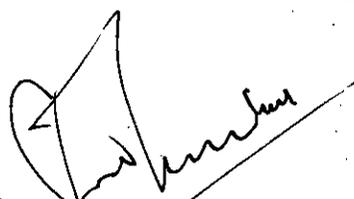
Dr. A.F.M. Anwarul Haque  
Professor & Head,  
Dept. of IPE, BUET, Dhaka,  
Member of the Committee.



Dr. M. Mizanur Rahman  
Associate Professor,  
Dept. of IPE, BUET, Dhaka,  
Member of the Committee.



Dr. A.B.M. Zohrul Kabir  
Assistant Professor,  
Dept. of IPE, BUET, Dhaka,  
Member of the Committee



Mr. S. M. Nuruzzaman  
Director (Production &  
Research), BCIC, Dhaka,  
External Member of the  
Committee.

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## ABSTRACT

Agriculture, contributing more than 50% to the GDP, is vitally important to the national economy. To boost up the agricultural sector, a huge amount of money has been invested in the agro-support and agro-based industries. The government is giving priority to build up the fertilizer industry because its marginal contribution is quite high. Urea fertilizer is highly useful to boost farm production. Bangladesh has, at present three urea fertilizer factories: Natural Gas Fertilizer Factory (NGFF) at Fenchuganj, Urea Fertilizer Factory at Ghorashal (UFFG) and Ashuganj Fertilizer and Chemical Company Limited (AFCCCL). A urea fertilizer factory at Polash, Dhaka is also under construction under Chinese assistance program.

One of the major causes of the poor contribution by these industries is the low capacity utilisation of its machineries and equipments. If these factories are to operate at the designed economic level, the down-time of various plants, machines and equipments will have to be kept to a minimum. An analysis of down-time will help to identify the causes of failure of machines and equipments and thus help to remove these causes systematically.

Due to time and other limitations the down-time of  $\text{NH}_3$ -plants of UFFG and NGFF were studied here. The period

of analysis for each factory was four years at the same age level.

It has been found that the processes differ mainly in sulphur removal from natural gas, reformation of natural gas,  $\text{CO}_2$  and  $\text{CO}$  separation from process gas, synthesis pressure, conversion rate of  $\text{N}_2$  and  $\text{H}_2$  into  $\text{NH}_3$  etc.

Down-time of  $\text{NH}_3$ -plant of NGFF is 3244 hours (5.89% of total productive time). Units which caused this down-time are: Main gas compressor (35.50%), Primary Reformer (22%), Secondary Reformer (12%) and  $\text{NH}_3$ -Cooled Copper Solution Cooler (8.65%). Main causes of the failure of the units are: high temperature (33.78%), leakage (24.27%), breakage (12.05%), damage (11.54%) and excessive vibration (10.18%).

The down-time of the  $\text{NH}_3$ -plant of UFFG is 3010 (10.14% of total productive time) hours. In UFFG, Synthesis Compressor caused 16% of the down-time and there was no down-time due to reformers. The share of other units is: Methanator Effluent Economizer (17.45%), Synthesis Compressor (15.87%), Air Filter (13%), Methanator (9.36%) and GV solution Pump (7.75%) of UFFG. In UFFG, damage, leakage, high temperature and blockage contribute 35.99%, 35%, 15.88% and 7.69% respectively.

The down-time of Main Gas Compressor of NGFF is plotted against the output pressure of the respective stages of the

(22)

iv(3)

compressor.. The points lie on straight line. The equation of the straight line is,

$$X - 0.947Y - 9.478 = 0$$

where X = Down-time of the Compressor in hour

Y = Output pressure of the Compressor (Kg/cm<sup>2</sup>)

Astonishingly the same point of the synthesis compressor of UFFG fits this line well. It was not possible in this thesis to identify the reasons.

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## A B B R E V I A T I O N S

ATB	Air-Turbo-Blower
BTU	British-Thermal-Unit
BD	Bangladesh
DSV	Desulphurised Vessel
FG	Flue Gas
FSS	Feed Stock Steam
FGWHB	Flue Gas Waste-Heat Boiler
FAD/F.A.O	Food & Agriculture Organisation
FNG	Fuel Natural Gas
GDP	Gross Domestic Product
GNP	Gross National Product
GPL, g/l	Gram per litre
GV	Giammarco-Vetrocoke
HTS	High Temperature Shift
H.P.A./HPA	High Pressure Absorber
IDF	Induced Draft Fan
K.O.	Knock Out
K.Cal/K.Cal	Kilo-Calorie
Kg/cm <sup>2</sup> A	Kg/cm <sup>2</sup> Absolute
Kg/cm <sup>2</sup> G	Kg/cm <sup>2</sup> Gauge
LTS	Low Temperature Shift
Mm <sup>3</sup> /mm <sup>3</sup>	(Milimeter) <sup>3</sup>
m <sup>3</sup>	(Meter) <sup>3</sup>

Mea/MEA	Mono-ethanol-amine
MGC	Main Gas Compressor
MT	Metric Ton
MP	Muriate of Potash
NGFF	Natural Gas Fertilizer Factory
NNP	Net National Product
NM <sup>3</sup>	Normal metre-Cube
DSR	Organic-Sulphur-Remover
PNG	Process Natural Gas
PSIG	Pound per square inch gauge
PR/IRF	Primary Reformer
PGHEX	Process Gas Heat Exchanger
SSP	Single Super Phosphate
SCF	Standard Cubic feet
t/hr	Ton/hour
TSP	Triple-Super Phosphate
UFFG	Urea Fertilizer Factory, Ghorashal
1RF	Primary Reformer / Ist Reformer
2RF	Secondary Reformer

CHAPTER I  
INTRODUCTION.

## 1.1 Identification of the problem



One of the challenging problems for the economy of developing countries like Bangladesh is its over population. It has an area of 141,100 sq.km. and population of about 90 million making it densely populated country in the world. About 80 percent of people are dependent on agriculture. Thus pressure on land is very high. Though the diffused rays of development have started reaching the country side scientific knowledge could not be fully utilised to improve the economy of the country. One of the fields where immense effort is necessary, is agriculture.

The per capita income of Bangladesh is the fifth lowest in the world. The contribution of agriculture towards the Gross Domestic Product (GDP) in 1980-81 was 53% (A-2) and that of industry only 7.3%. Hence priority projects for the development should be in agriculture. The policy of the government is rightly directed toward this goal. A huge sum of money has been invested for the agro-support and the agro-based industries. The government is giving priority to build up the fertilizer industry because the marginal contribution of the fertilizers is quite higher.

There are many ways of classifying the fertilizers<sup>1</sup>. These are:

1. According to the nature of source
  - a) Natural fertilizers
  - b) Artificial fertilizers
  - c) Mineral fertilizers

a. Natural fertilizers

These materials are naturally available in a firm or locality and are constituted mainly of the remains of plants and animals. Example: farmyard fertilizer, compost etc.

b. Artificial fertilizers

The materials included here are found either as deposits like the phosphate rocks and potash minerals or artificially manufactured products like ammonium sulphate, superphosphate, urea etc. Mainly all these are of inorganic nature.

c. Mineral fertilizers

These are found in the ash of the wood, the phosphates, the chlorides, sulphates of the alkaline earth-metals.

2. According to nutrient content

a. Single fertilizers

b. Compound fertilizers

a. Single fertilizers

The fertilizers having only one nutrient element are included in this group, for instance, ammonium sulphate, urea etc.

b. Compound fertilizers

These materials which are capable of supplying more than one nutrient elements. For example (1) Ammonium phosphate

supplies both nitrogen and phosphorus and (2) Nitrophoska supplies both nitrogen, phosphorus and potassium.

3. According to the nature of action

- a. General fertilizers
- b. Special fertilizers
- c. Stimulated substances
- d. Soil amendments or soil conditioners

a. General fertilizers

When added to the soil, these undergo a series of changes and release many nutrient elements which are required for crop growth. Examples; farmyard manure, compost, green fertilizer, sewage sludge and guano.

b. Special fertilizers

When added to the soil, these undergo only few changes and release one or more plant nutrient elements. All the nitrogenous phosphatic and potassium fertilizers are included in the group.

c. Stimulated substances

These are the carriers of trace elements. They are said to be stimulated in their action on plant growth.

d. Soil amendments or soil conditioners

These are capable of increasing the crop growth by suitably modifying the physical, chemical and biological properties of the soil, for instance, gypsum and lime.

Another classification of fertilizer is listed in Annexure A-3.<sup>1</sup>

Nitrogen, among the plant nutrient, has the most profound influence on plant growth. Insufficiency of this nutrient as well as excess is liable to reduce the yields and impair the quality of the produce. It is a constituent of protoplasm and is closely associated in the activity of every living cell. Consequently, excess or deficiency of this nutrient will immediately affect the plant growth.

Though young plant require more nitrogen than those approaching maturity, the plants continue to absorb nitrogen throughout their life. Hence an adequate supply of nitrogen is essential in the soil for good plant growth.<sup>2</sup>

Nitrogen is available at various percentage in different types of fertilizers (table 1.1)<sup>2</sup>

The nitrogen content (46%) in urea fertilizer is the maximum of all nitrogenous fertilizers. So urea is the main fertilizer for supplying maximum nitrogen in the soil.

Methane gas is the main raw material for the production of urea fertilizer. In Bangladesh, NG contain about 99.5% methane. It is available at low cost and for this reason NG based fertilizer is suitable for our country. Government has invested huge amount of money in urea fertilizer production and also decided to install more fertilizer factories in future for the fulfilment of country's demand as well as to earn foreign currency through export.

Table 1.1 Different types of nitrogenous fertilizers showing the nitrogen content

Name of fertilizer	Nitrogen content %	Form of Nitrogen
1. Sodium nitrate ( $\text{NaNO}_3$ )	16	Nitrate
2. Potassium nitrate ( $\text{KNO}_3$ )	12.5-13.5	Nitrate
3. Ammonium sulphate ( $(\text{NH}_4)_2\text{SO}_4$ )	20.6	Ammoniacal
4. Ammonium Chloride ( $\text{NH}_4\text{Cl}$ )	26.0	Ammoniacal
5.* Ammophos-A	11.0	Ammoniacal
6.* Ammophos-B	16.0	Ammoniacal
7. Ammonium nitrate ( $\text{NH}_4\text{NO}_3$ )	33.0	Ammoniacal 16.5 + Nitrate 16.5
8. Ammonium sulphate nitrate	25.6	Ammoniacal 19.0 + Nitrate 6.6
9. Urea $\text{CO}(\text{NH}_2)_2$	46.0	Amide
10. Calcium Cyanamide ( $\text{CaCN}_2$ )	20.6	Amide
11. Dried Blood	3-14	Protein(organic)
12. Ground nut cake	8.0	Protein(organic)
13. Meat meal	9.0-11.0	Protein(organic)
14. Guano	10.0	Protein(organic)

\* Ammophos-A — Commercial monoammonium phosphate  
 Ammophos-B — Commercial grade of ammonium phosphate  
 containing 16% nitrogen and 20%  $\text{P}_2\text{O}_5$

NGFF (Natural Gas Fertilizer Factory) is the first urea fertilizer factory in the country. Its commercial production of urea started in 1962 with an annual rated capacity of 1,06,000 MT (353 MT/day). UFFG (Urea Fertilizer Factory, Ghorashal) started in 1970 with an annual rated capacity of 3,40,000 MT (1133 MT/day). AFCCL (Ashuganj Fertilizer and Chemical Company Limited) at Ashuganj with a capacity of 1600 MT per day is giving commercial production since 1962. The fourth factory with a capacity of 1730 MT of urea per day is under implementation. It is expert oriented and will be erected at Chittagong. A Chinese assisted urea fertilizer factory at Polash, Dhaka having an annual rated capacity of 100,000 MT (333 MT/day) is also under construction. Unfortunately, these factories have failed to produce the promised return. One of the major causes of the poor contribution by these industries is intolerably low capacity utilisation of its machineries.

Bangladesh imports machinery and equipment by using its hard-earned foreign currency. Therefore full utilization of these machineries is necessary. In spite of gaining experience in urea production, troubles and failures of machines, equipments and plants repeat in these industries.

If these factories are to operate at the designed economic level, the down-time of various plants, machines

and equipments will have to be kept to a minimum. An analysis of the down-time will help to keep these bottle-neck points in operation and also in selection of process for the future urea fertilizer factories.

## 1.2 Objectives

In the light of the above discussion two pioneer urea fertilizer industries, namely Natural Gas Fertilizer Factory, Fenchuganj (NGFF) and Urea Fertilizer Factory, Ghorashal (UFFG) will be studied in this thesis. The analysis was based on the down-time of the machineries and equipments. It will not be correct to compare the data of the same year, because the age of the plant has an influence on the down-time. Necessary corrections will be accordingly made.

Due to time limitation of this thesis, the down-time analysis will be limited to the ammonia plant of the industries. The objective of the analysis will be:

1. To identify machineries and equipments, which caused the plant shut-down and the influence on urea production.
2. To identify the technological constraints; such as temperature, leakage, pressure, which caused the failure of the machineries and equipments. The relative weightage of these factors will also be calculated.
3. To compare the processes used in both the fertilizer factories.
4. To recommend suitability of various machines and equipments used in the ammonia plant.

CHAPTER - 2  
NATURAL GAS FERTILIZER FACTORY ( NGFF )

## 2.1 History of NGFF<sup>3</sup>

Located at Fenchuganj in the district of Sylhet Natural Gas Fertilizer Factory (NGFF) is the pioneer fertilizer project of the country and consists of Ammonia plant and Urea Plant including auxiliary facilities like Power Station, Water Treatment Plant etc. The Urea and Ammonia Plant including auxiliary facilities like Water Treatment Plant, Maintenance workshops, Laboratory etc. was erected by KOBE STEEL LTD. of Japan. The factory has its own Power Plant having three turbo-generators of 12 MW each which was erected by Mitsubishi Heavy Industries Organised Ltd. Later in 1969, one Ammonium sulphate plant was set up to meet the demand of the neighbouring Tea Estates. The Ammonium Sulphate Plant was constructed by MITSUBISHI SHOJI KAISHA LTD. of Japan.

The chief raw materials for the factory are Natural Gas and Sulphur. Natural Gas as feed stock and fuel is received from Haripur gas field 28 miles away from the factory. Sulphur is imported raw materials.

1. Date of commencement of trial production, Dec, 1961
2. Date of commencement of commercial production, July, 1962
3. Capacity of production of urea, 106,000 MT/Yr.
4. Capacity of production of Ammonium Sulphate, 12000 MT/yr.
5. Consumption of utilities;
  - a) Make up water 1700 MT/hr.

- b) Cooling Tower of Power Station 550 MT/hr.
- c) Sanitary Water 200 MT/hr.
- d) Pure Water 150 MT/hr.
- e) The power plant has three generating units with a capacity of 12 MW each. Normally two units are in continuous operation and the other one is kept stand-by alternately to facilitate maintenance works.

## 2.2 Description of NGFF<sup>3</sup>

Main raw materials for the production of urea fertilizer are liquid ammonia, liquid carbon-di-oxide and air. The former two components are manufactured from the ammonia plant from the Natural Gas. The natural gas is received from Haripur Gas field whose chemical analysis is given in the Annexure- A- 8.

### AMMONIA UNIT

The raw materials used in NGFF for the synthesis of Ammonia are Natural gas(NG), air and water.

#### 2.2.1 Desulphurisation process

##### 2.2.1.1 Process Description

The natural gas obtained from gas field is kept at a pressure  $10 \text{ kg/cm}^2$  and is received in Main Gas receiver. From the Main gas receiver, a portion of natural gas used for processing which is called as Process Natural gas(PNG), another portion used as fuel which is called Fuel Natural Gas (FNG).

The PNG then passes through the Organic Sulphur Removal (OSR) vessel containing catalyst which removes hydrogen sulphides and organic sulphur. The catalyst is activated charcoal. The sulphur content of the gases leaving the OSR vessel should be held to 1 or 2 ppm maximum as sulphur will poison the primary reformer catalyst.

### 2.2.2 Reforming process

#### 2.2.2.1 Primary Reformer

In the primary reformer, the catalyst is contained in the long vertical special alloy stainless steel tubes (25 cr, 20 Ni). These tubes are arranged in the cylindrical refractory-lined combustion section of the furnace in two vertical planes on the diameter of the heater and intersecting at right angles. Natural gas is used to fire the furnace.

#### 2.2.2.2 Secondary Reformer

This vessel contains similar catalyst to that in the primary reformer but is arranged in a single bed.

#### 2.2.2.3 Catalyst

Reforming Catalyst contains 22-25% of Nickel deposited on the refractory of  $MgO$ ,  $Al_2O_3$ ,  $Ca(AlO_2)_2$  and  $CaSiO_3$ . It has the colour of straw due to oxidised Nickel.

The Catalyst is insoluble in water, but loses activity when the water contains sulphur compounds or dissolved salts.

When to start the process with this catalyst wet with water, the temperature should be raised at the rate lower than  $38^{\circ}\text{C/hr}$ . because at the rate higher than  $38^{\circ}\text{C/hr}$ , the water in the catalyst vaporizes very quickly and result in the expansion of the volume of catalyst. Thus the strength of catalyst is reduced.

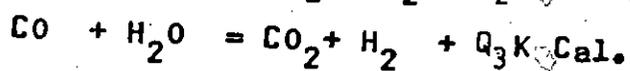
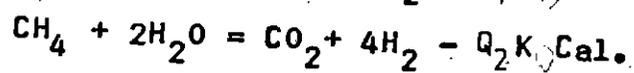
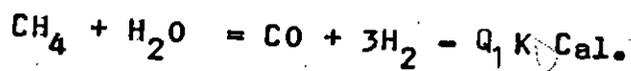
When Ni in this catalyst is in oxidised state, the catalyst may be heated to  $980^{\circ}\text{C}$  in  $\text{N}_2$ , steam or air.

The catalyst is strong against temperature change but the temperature rise lower than  $90^{\circ}/\text{hr}$  is desirable.

#### 2.2.2.4 Process Description

The Process Natural Gas (PNG) from the OSR vessel is mixed with steam and then passes through the Mixed Gas Preheater where it is heated to  $350^{\circ}\text{C}$  by heat exchange with hot converted gas from the CO-converter.

The resultant mixture passes over catalyst in the primary reformer tubes and the following basic reactions occur.



Leaving the tubes the reformed gas has a temperature of  $620^{\circ}\text{C}$ , while the flue gases which surround the tubes leave the furnace at about  $800^{\circ}\text{C}$ .

Reformed gas and steam mixture leaving the furnace enter the Secondary Reformer. At the inlet of the secondary reformer sufficient air is mixed with the incoming hot gas from the primary reformer to give the desired ratio of hydrogen to nitrogen in the product gas after CO conversion. A portion of the gas burns with air in the upper part of the 2RF, resulting in a temperature of about  $930^{\circ}\text{C}$ . As the gases pass through the catalyst, bed reactions similar to those occurring in the primary occur and result in lowering the  $\text{CH}_4$  content of the exit gas to about 0.3%.

Leaving the 2RF, the reformed gas is cooled to CO conversion temperature in the Reformed gas Waste Heat Boiler. The temperature  $410^{\circ}\text{C}$  at the inlet of the converter is governed by amounts of hot reformed gas passed through and by passed around the boiler.

The CO-converter is charged with catalyst  $\text{Fe}_3\text{O}_4$  which promotes the water gas shift reaction, removing most of the CO and producing additional hydrogen. This reaction is exothermic and generates heat. The reformed gas from the Waste Heat Boiler enters into the CO-converter at  $410^{\circ}$  and leaves as converted gas containing about 2% CO at about  $440^{\circ}\text{C}$ .

Leaving the CO-converter the converted gas is partially cooled by heat exchange with the incoming reaction gas and steam in the Mixed Gas Preheater.

The gas passes through the tube side of the carbonate reboilers giving up heat for carbonate solution regeneration and then flows to the Gas Scrubber Cooler for final cooling before compression. A Knock-Out-Drum affords production to the compression section in case of carryover from the Gas Scrubber Cooler. Water exit the Gas Scrubber Cooler is used to heat boiler feed water and then sewerred.

#### 2.2.2.5. Operating Conditions

First Reformer (1RF) outlet process gas temp.	=620°C.
Secondary Reformer(2RF) outlet process gas temp	=816°C.
CO-converter inlet process gas temperature	=410°C.
CO-converter outlet process gas temperature	=440°C.
Flue gas temp. at the outlet of 1RF	=800°C.
CH <sub>4</sub> content at the outlet gas of 1RF.	=7.7%
Steam/Drygas ratio at 1RF outlet gas	= 1.1
CH <sub>4</sub> content at the outlet gas of 2RF	= 0.3%
CH <sub>4</sub> content at the outlet gas of CO-converter	= 0.2%

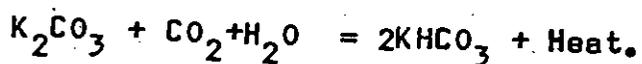
The flue gas from the Primary Reformer is added to natural gas in the Down-comer and then passed through Flue Gas Waste Heat Boiler(FGWHB). The gases burn in the Down-comer raising the resultant mixed gas temperature to 1040°C. This

gas mixture is then cooled in the Flue Gas Waste Heat Boiler to about 900°C and is discharged into the stack by Induced Draft Fan (IDF).

### 2.2.3 CO<sub>2</sub> Removal Process

#### 2.2.3.1 Process Description

Synthesis gas from the 1st stage of compression enters the CO<sub>2</sub>-absorber at about 20 kg/cm<sup>2</sup> and 107°C containing approximately 16.7% of CO<sub>2</sub>. In passing through the carbonate solution following counter currently over rasching ring made of iron CO<sub>2</sub> is absorbed according to the following reactions.



The gas leaves the CO<sub>2</sub> absorber at approximate 96°C and 20 kg/cm<sup>2</sup>. Heat is removed by indirect cooling with cooling water in the gas after cooler before final separation of condensate in the knock out drum. The temperature and pressure of the gas returning to the 2nd stage compressor are about 40°C and 19.4 Kg/cm<sup>2</sup>. Then it enters the 3rd stage compressor.

One third of the solution (equivalent 20-25% by weight K<sub>2</sub>CO<sub>3</sub>) containing approximate 15 mm<sup>3</sup> of CO<sub>2</sub> per m<sup>3</sup> enters the top of the Absorber at 93°C after being cooled in the solution cooler and diluted with condensate from the CO<sub>2</sub> and Water Separator. The remaining two third of the solution enters the Absorber between the two beds of packing at essentially the temperatures in the Reboiler (112°C).

The rich solution leaves the Absorber at 107°C containing approximate 30 mm<sup>3</sup> of CO<sub>2</sub> per m<sup>3</sup> flashes into the top of the solution Regenerator. The flashed gas mixed with the CO<sub>2</sub> rising from the packing passes to the Overhead Condenser where it is cooled to 43°C. Water vapour condensed is separated from CO<sub>2</sub> in the CO<sub>2</sub> and Water Separator. CO<sub>2</sub> is piped to Urea Plant. The condensate is pumped to the Absorber top solution feed line by the condensate pump.

Liquid from the top of the Regenerator passes down through the packing and into the Reboiler giving up CO<sub>2</sub> according to the following reaction.

$2\text{KHCO}_3 + \text{Heat} = \text{K}_2\text{CO}_3 + \text{CO}_2 + \text{H}_2\text{O}$  Heat for the reaction is supplied by hot gases from the Gas Reforming Section. The regenerated or lean solution flows to the solution pump for recycle to the absorber.

#### 2.2.4 High Pressure Purification Process

##### 2.2.4.1 Process Description

The discharge of the 3rd stage of the compressor is a gas mixture which is expected to consist on the average of H<sub>2</sub> and N<sub>2</sub> in the ratio of 3 to 1, together with CO<sub>2</sub>, CO, CH<sub>4</sub> plus argon.

All of these gases except N<sub>2</sub> & H<sub>2</sub> may be considered impurities, either because they are catalyst poisons or because, merely by their presence, they reduce the effective hydrogen-nitrogen pressure. Furthermore, carbon di-oxide may form

ammonium carbonate or carbonate, which may plug the synthesis process lines.

The third stage output of the Main compressor at a normal pressure of  $135 \text{ Kg/cm}^2$  flows through the Copper-Liquor-Scrubber where it is scrubbed with refrigerated cu-liquor.

Copper liquor scrubber consists of five layers packed with iron resching rings and reduces CO to 5 to 10 ppm and  $\text{CO}_2$  50 to 60 ppm.

Then the gas enters the Final purifier where  $\text{CO}_2$  content is further reduced to 0 to 5 ppm by Ammonia aqua.

The gas leaving the final purifier flows through the High Pressure (H.P.) Separator where any entrained liquid is removed and then to the fourth stage suction of the main compressor.

The cu-liquor leaving the scrubber flows to the top of the copper-liquor Regeneration tower which consists of 2 layers packed with 38/38 iron rasching rings. The liquor flows through the upper layer downward releasing almost all  $\text{CH}_4$ ,  $\text{H}_2$ ,  $\text{N}_2$  & CO which have been absorbed physically.

Then it enters the Copper Solution Steam Heater where it is heated to  $78^\circ\text{C}$  to release remaining CO,  $\text{CO}_2$ ,  $\text{NH}_3$  etc. and returns to the Copper-Solution-Regeneration-Tower layer, flows down to the bottom and then enters Copper-Solution-Buffer Tank where it stays a little while to release all the remaining gas. Here the regeneration is completed.

The gases released in the above mentioned process ( $\text{CO}$ ,  $\text{CO}_2$ ,  $\text{CH}_4$ ,  $\text{NH}_3$  &  $\text{N}_2$ ) enter the Copper-Solution Regeneration Tower, go up through the tower giving heat and  $\text{NH}_3$  to the liquor being regenerated and are vented from the system as off gas.

The off gas enters a steam jacketed pipe and heated to prevent ammonium carbonate formation. The off gas, containing principally  $\text{CO}$ , is burned with flue gas entering Down-Comer and then Flue Gas Waste Heat Boiler in the Gas Reforming Section.

To keep an adequate ratio of  $\text{Cu}^{++}$  air is added to the bottom of the Regeneration Tower.

The amount of  $\text{NH}_3$  in the copper-liquor is adjusted by adding  $\text{NH}_3$  to the top of the Copper Solution Buffer Tank.

Generated Copper-Liquor (about  $75^\circ\text{C}$ ) is pumped from the Buffer Tank by a low pressure copper solution pump (about  $5 \text{ Kg/cm}^2$ ), through the shell side of the water cooled  $\text{Cu}$ -solution cooler (about  $40^\circ\text{C}$  at the exit) and the tube side of the  $\text{NH}_3$  Cooled Copper Solution Cooler where it is cooled to approximate  $0^\circ\text{C}$ .

The cold regenerated copper liquor then passes through the copper solution pump suction filter to the H.P.  $\text{Cu}$ -liquor pump which pumps the liquor to the copper liquor scrubber.

A small amount of  $\text{Cu}$ -liquor leaving the Water-Cooler is diverted from the main liquor stream and passed through a  $\text{Cu}$ -liquor side back into the top of the Regeneration Tower or into

the Buffer Tank to remove solid impurities and foreign materials.

The aqua  $\text{NH}_3$  made from pure water and  $\text{NH}_3$  gas from  $\text{NH}_3$ -Recovery and Storage Section, leaving the Final Purifier flows to the Aqua Let-Down Tank and from there to the Urea Plant.

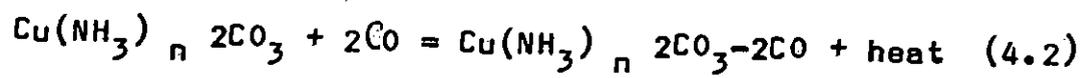
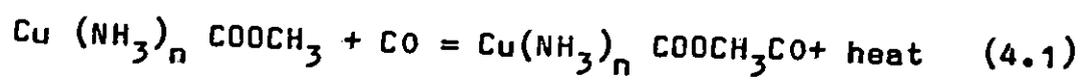
#### 2.2.4.2 Chemistry of the Process

The solution used for the absorption of CO is copper ammonium acetate-carbonate liquor containing approximate 125 Gram/Litre (GPL) of cuprous copper and 25 GPL of Cupric Copper. It is prepared as the acetate salt and initially contains little or no carbonate.

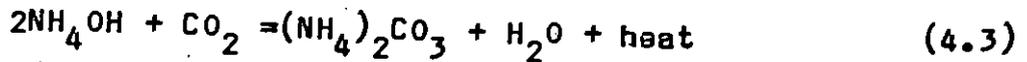
The basic reaction which take place during the absorption of CO by copper-liquor and regeneration of the liquor are outlined below:

##### 1. Copper Liquor Scrubbing

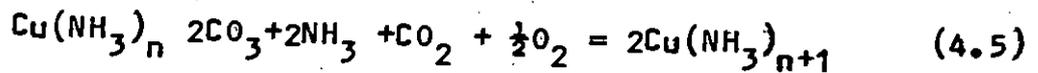
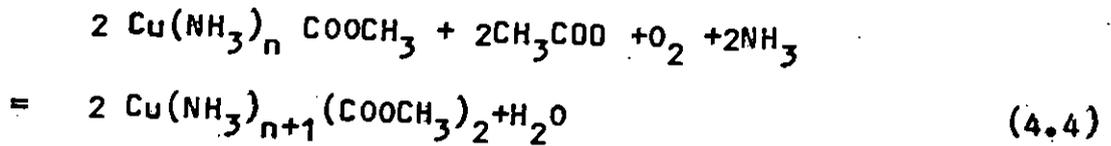
CO is absorbed by the copper liquor with the formation of the cuprous CO compound and the liberation of a considerable amount of heat.



$\text{CO}_2$  is absorbed by free  $\text{NH}_3$  to form ammonium carbonate. This reaction also liberates heat.



$\text{O}_2$ , if present is absorbed by the cuprous salt which is oxidized to the cupric form.



## 2. Reflux

The rich liquor at about  $25^\circ\text{C}$  enters the top of the reflux section which is the upper layer of the Regeneration Tower.

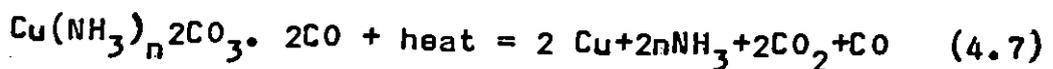
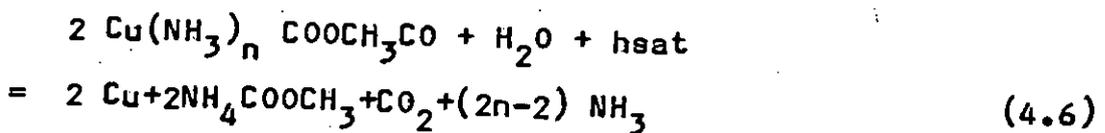
Some of the absorbed gases flash off and most of the  $\text{NH}_3$  driven off at the regeneration temperature is reabsorbed.

## 3. Regeneration

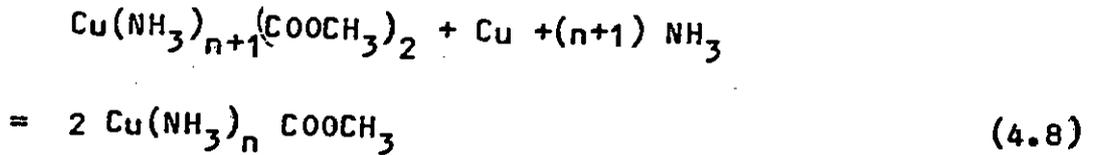
The solution is heated to  $78^\circ\text{C}$  reversing reactions (4.1) and (4.2) to liberate  $\text{CO}$ .

## 4. Reduction

In addition to the above regeneration reactions, some of the  $\text{CO}$  which is in the cuprous complex salt is known to decompose in such a way that  $\text{CO}_2$  and metallic copper are formed.



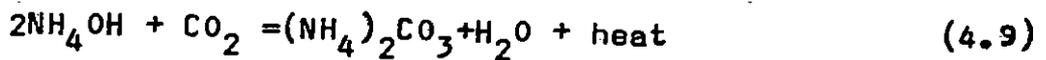
This metallic copper will react immediately with the cupric iron and reduce them to cuprous state.



The above reactions follow laws of equilibrium. For example in cold solutions with high, partial pressures of CO and CO<sub>2</sub> in the gas phase, reactions (4.1), (4.8) & (4.3) will proceed toward the right, whereas in hot solutions with high concentrations of these components in the liquid phase and low partial pressure in the gas phase, these reactions will reverse and proceed to the left.

#### 5. Aqua NH<sub>3</sub> Scrubbing

To remove all but a trace of the gas stream the gas is scrubbed in the Final Purifier with a 5% NH<sub>4</sub>OH solution.



#### 2.2.4.3 Operating Conditions

The gas contents under the normal operating conditions are as follows:

	Inlet Cu-Liquor Scrubber	Inlet Final Purifier	Exp High Pressure Separation
H <sub>2</sub>	72.2%	-	74.48%
N <sub>2</sub>	24.1%	-	24.82%
CO	2.40%	5 to 10 ppm	5 to 10 ppm
CO <sub>2</sub>	0.61%	50 to 60 ppm	0 to 5 ppm
O <sub>2</sub>	trace	0 ppm	0 ppm
CH <sub>4</sub>	0.69	-	0.70%

## 2.2.5 Synthesis Process

### 2.2.5.1 Process Description

After fourth stage compression, purified gas enters the synthesis section at 300-365 Kg/cm<sup>2</sup>. The system pressure changes according to the activity of the catalyst, gas contents converten temperature and space velocity. This gas is expected to contain about 5ppm CO<sub>2</sub>, 10ppm CO, No O<sub>2</sub>, 0.3% argon, 0.4% CH<sub>4</sub>, 74.5% hydrogen and 24.8% N<sub>2</sub>.

This gas is referred to as Make up gas. Make up gas, together with recycled gas from the circulator is passed through the filter to remove any entrained oil.

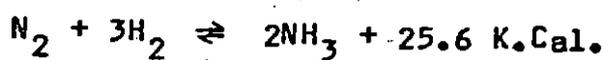
The gasses then pass through the shell of the Col Exchanger exchanging heat with the low temperature gas (about 0°C) in the tube side and the temperature of the gas is lowered down to 2-15°C and part of the NH<sub>3</sub> in the circulating gas is condensed to liquid NH<sub>3</sub>.

Then the gases enter the NH<sub>3</sub> cooled condenser and are cooled down to 0°C or below. Here the NH<sub>3</sub> content in the circulating gas will be 3 to 4%.

The mist-free gas leaving the Secondary separator where the liquid NH<sub>3</sub> is separated from the gas stream is sent through the tubes of the Cold Exchanger, passing through which the temperature of the gas rises to 15-35°C to the NH<sub>3</sub> converter.

The NH<sub>3</sub> converter consists of Heat Exchanger part and catalyst bed.

The circulating gas enters the shell side of the Heat Exchanger Part, where it exchanges heat with temperature reaction gas in the tube side and then passes through catalyst bed cooling tube absorbing reaction heat. The circulating gas temperature at the inlet of the catalyst is 420-450°C. In converter the following reactions takes place.



The reaction gas temperature is lowered down to 165°-200°C passing through the tube side of the Heat Exchanger part.

The gases leaving the converter are sent to the Water Cooled condenser, (where they are cooled down to the cooling Water temperature and 65-90% of  $\text{NH}_3$  is condensed) and following it the Primary Separator where about 50% to 60% of the total  $\text{NH}_3$  production is separated.

The gas then pass to the circulator where the gas pressure is increased from about 330  $\text{Kg/cm}^2$  to about 365  $\text{Kg/cm}^2$  or enough to overcome the pressure drop through the system.

The gas from the circulator discharge passes to the filter where additional Make up gas is introduced and the cycle repeated.

A small portion of the gas leaving the Primary Separator is purged from the system to prevent the accumulation of inerts ( $\text{CH}_4$  and Ar).

$\text{NH}_3$  which has been removed in either the Primary or the Secondary Separator is let down to 17-18  $\text{Kg/cm}^2$  and piped to  $\text{NH}_3$  Storage and Recovery Section.

#### 2.2.5.2 Operating Condition

$\text{NH}_3$  content of the gases at the inlet of the converter = 3%

$\text{NH}_3$  content of the gases at the outlet of the convertor = 15.8%

#### 2.2.6 $\text{CO}_2$ Removal Process From Flue Gas

##### 2.2.6.1 Process Description

$\text{CO}_2$  (10.6%) in the flue gas of the Power Plant is recovered as pure  $\text{CO}_2$  (99% or above) by means of Monoethanol Amine (MEA). Recovered  $\text{CO}_2$  gas is used for synthesis of Urea.

$\text{CO}_2$  content in the vent gas exit the  $\text{CO}_2$  absorber is normally below 0.4%.

Process Flue Gas (FG) is taken out just after boiler I.D. fan. FG at atmospheric pressure and  $160^\circ\text{C}$  enters FG Cooler where it is washed and cooled down to  $37^\circ\text{C}$  is separated of water passing through Separator and enters FG Boiler, suction pressure of which is  $0.95 \text{ Kg/cm}^2 \text{ A}$ .

FG leaving FG Blower is  $68.4^\circ\text{C}$  &  $1.21^k$  and enters the bottom of  $\text{CO}_2$  Absorber where  $\text{CO}_2$  in FG is absorbed by MEA Solution ( $\text{CO}_2$  lean and  $41^\circ\text{C}$ ).

FG now free of  $\text{CO}_2$  leaves the Absorber from the top, enters separator where it is separated of water and MEA carryover and is then vented to atmosphere at  $41^\circ\text{C}$ .

Mea mist carried over and separated in the Separator returns to  $\text{CO}_2$ -Absorber bottom. Temperature at the Absorber bottom is  $55^\circ\text{C}$  and almost all  $\text{CO}_2$  in FG is absorbed.

Mea solution now  $\text{CO}_2$  rich and at  $55^\circ\text{C}$  flows through rich solution pump and then heated to  $95^\circ\text{C}$  in solution Heat Exchanger, filtrated by Amine filter and enters the top of Amine Regenerator.

Pressure of  $\text{CO}_2$  rich amine solution is reduced to  $0.9 \text{ Kg/cm}^2\text{G}$  ( which is the pressure of Amine Regenerator) and a part of  $\text{CO}_2$  is set free Amine solution is regenerated by hot vapor from Reboiler while going down to the bottom.

$\text{CO}_2$  in the amine solution is almost completely evaporated. Solution temperature at the bottom of the Regenerator is  $121^\circ\text{C}$ .

Reboiler is heated by  $3.5 \text{ Kg/cm}^2$  steam.

Mea solution at  $121^\circ\text{C}$  now  $\text{CO}_2$  lean, is sent to the shell side of Solution Heat Exchanger by pressure difference, where it is cooled down to  $80^\circ\text{C}$ . Then it enters the shell side of Solution Cooler and is cooled further down to  $40^\circ\text{C}$  and it flows to lean solution pump.

Mea solution is pumped to the top of  $\text{CO}_2$  Absorber which keeps the solution flow at 74t/hour in normal operation.

Thus the circulation of Mea solution ( $\text{CO}_2$  absorption and regeneration cycle) is continued.

CO<sub>2</sub> and steam mixture leaving the top of Amine regenerator at 108°C is cooled down to 40°C passing through Overhead Condenser is separated of water in Acid Gas Separator and goes to CO<sub>2</sub> Gas Holder.

In the operation of the CO<sub>2</sub>-removal section, O<sub>2</sub> should be below 2.5% Soda ash is used as MEA detergent and 10% CuSO<sub>4</sub> aqua is added as inhibitor of corrosion.

Chemical reaction in this section is  $\text{CO}_2 + \text{H}_2\text{O} + 2\text{NH}_2\text{CH}_2\text{OH} = (\text{HOCH}_2\text{CH}_2\text{NH}_3)^{++} \text{CO}$  MEA solution contains 18% Wt. of MEA.

## UREA UNIT

### 2.2.7 Urea Synthesis and Separation

#### 2.2.7.1 Process Description

Liquid NH<sub>3</sub> is subcooled in NH<sub>3</sub> Subcooler by brine and fed to Synthesis Autoclave at pressure of 300 Kg/cm<sup>2</sup> by liquid NH<sub>3</sub> feed pump.

Liquid CO<sub>2</sub> from storage tank through CO<sub>2</sub>-Cooler in which liquid CO<sub>2</sub> is sub-cooled by brine is fed to Autoclave.

A small amount of air is sent to Synthesis Autoclave and Excess NH<sub>3</sub> separator by Process Air Compressor for protecting corrosion inside of the above vessels. Liquid NH<sub>3</sub> and CO<sub>2</sub> mixed in Synthesis Autoclave react to form the reaction mixture consisting of Urea, carbonate, water and excess NH<sub>3</sub>. Discharged mixture from Synthesis Autoclave enters to Excess

$\text{NH}_3$  Separator which is a kind of distillation tower. Excess  $\text{NH}_3$  separated as a gas exists from Excess  $\text{NH}_3$  Separator top and it is transferred to Excess  $\text{NH}_3$  condenser where  $\text{NH}_3$  gas is condensed returned to  $\text{NH}_3$  storage Tank. Reaction mixture separated from excess  $\text{NH}_3$  is transferred from Excess  $\text{NH}_3$  separator bottom to Decomposition and Evaporation section.

#### Synthesis Autoclave

Liquid  $\text{CO}_2$  Liquid  $\text{NH}_3$  and process air enter to Synthesis Autoclave bottom. These mixture reacts to form Urea, carbonate, water excess  $\text{NH}_3$ ,  $\text{H}_2$ ,  $\text{N}_2$  and  $\text{O}_2$ . Pressure and temperature is controlled by changing  $\text{NH}_3$  or  $\text{CO}_2$  feed ratio. The reaction being exothermic, no external heat is required.

#### Excess $\text{NH}_3$ Separator

This separator is designed to separate most of the excess  $\text{NH}_3$  present in reaction mixture from Synthesis Autoclave. Heating coil is prepared at the bottom to heat the autoclave products by  $10.5 \text{ Kg/cm}^2$  steam. Excess  $\text{NH}_3$  separator is equipped with a top condenser. A portion of the stripped  $\text{NH}_3$  gas is condensed by cooling water, which passes down to Tower as reflux and the rest passes as a gas.  $\text{NH}_3$  gas from Excess  $\text{NH}_3$  separator is delivered to Excess  $\text{NH}_3$  condenser where the  $\text{NH}_3$  gas is condensed and returned to  $\text{NH}_3$  storage Tank for reuse.

#### 2.2.7.2 Operating Conditions

From Excess  $\text{NH}_3$  separator, excess  $\text{NH}_3$  gas is separated from urea solution. The constituents of Gas and urea

solution is given below:

Name of Constituents	Inlet of Decomposer		Name of constituent	Inlet of Excess NH <sub>3</sub> Condenser
	Kg. mol/hr.	% of Wt.		
NH <sub>3</sub>	69.4	12.1	NH <sub>3</sub>	317
Carbamate solution	26.7	21.1	N <sub>2</sub>	1.2
Urea	80.1	48.8	H <sub>2</sub>	0.6
Water	98.7	18.0		

## 2.2.8. Decomposition and Evaporation Process

### 2.2.8.1 Process Description (Decomposer System)

Synthesis solution from the bottom of Excess NH<sub>3</sub> separator is sent to Decomposer in which unconverted carbonate is disassociated back to NH<sub>3</sub> and CO<sub>2</sub> by heating with 10.5 Kg/cm<sup>2</sup> steam. The gases and solution mixture are delivered to Decomposer Separator where the reaction mixture is separated to the mixed gas of NH<sub>3</sub>, CO<sub>2</sub>, H<sub>2</sub>O and Urea solution accompanying small quantities of dissolved NH<sub>3</sub> and CO<sub>2</sub>. The mixed gas from Decomposer separator is passed overhead to Absorber Stripper and urea solution delivered to Evaporator system.

In the Evaporator urea solution is heated up by 10.5 Kg/cm<sup>2</sup> steam and the concentration of urea solution will be brought up to more than 99.5%. In Evaporator, Vacuum is taken and air stripping is performed to avoid formation of biuret as far as possible.

Interior of Evaporator is maintained vacuum by steam ejectors.

Urea solution concentrated more than 99.5% is delivered to Prilling Tower and is formed into spherical droplets. The mixed gas of  $\text{NH}_3$ ,  $\text{CO}_2$  and  $\text{H}_2\text{O}$  from Evaporator is washed with ammonium carbonate solution and then passes into Evaporator surface condenser where some of the gases are condensed. This condensates are collected in Evaporator Hot Well. Gases not condensed then pass to Evaporator Inter-Condenser and Evaporator After Condenser for Cooling and condensation and collected in Hot Well and gases not condensed too is purged to the atmosphere. A portion of solution collected in Evaporator Hot Well is pumped into washer by Washer Spray Pump and circulated to wash evaporating gas and the rest is delivered to  $\text{NH}_3$  Stripper System by Evaporator Condensate Pump.

#### 2.2.8.2 Process Description ( $\text{NH}_3$ Stripper System)

Ammonium carbonate solution stored in Evaporator Hot Well is delivered to  $\text{NH}_3$  Stripper by Evaporator Condensate Pump. On its way it passes through  $\text{NH}_3$  Stripper, Bottom Exchanger for recovery of waste heat of purge solution from the bottom of  $\text{NH}_3$  stripper.

#### 2.2.9 $\text{NH}_3$ - $\text{CO}_2$ Separation Process

##### 2.2.9.1 Process Description

In this section the mixed gases consist of  $\text{NH}_3$ ,  $\text{CO}_2$  &  $\text{H}_2\text{O}$  vapour are perfectly separated.  $\text{NH}_3$  gas is compressed and

and liquefied and then sent to Urea Synthesis and separator section, while  $\text{CO}_2$  gas is sent to  $\text{CO}_2$  compressor section. Water vapor is condensed and purged to sewer.

The procedure separating  $\text{NH}_3$  from  $\text{CO}_2$  stands on the different solubility of  $\text{NH}_3$  and  $\text{CO}_2$  against MEA solution. This section consists of two complete trains. Each train is divided into the following four systems.

#### 1) Absorber Stripper System

The mixed gases consists of  $\text{NH}_3$ ,  $\text{CO}_2$  &  $\text{H}_2\text{O}$  vapor from Decomposer Separator and  $\text{NH}_3$  Stripper Condenser are fed to absorber zone bottom of Absorber Stripper. Two feed nozzles to Absorber Stripper are installed up and down. Either of two nozzles is used in accordance with separating condition in Absorber Stripper. In Absorber Stripper  $\text{H}_2\text{O}$  is condensed and passes downward and  $\text{CO}_2$  being absorbed by MEA solution passing downward is separated from  $\text{NH}_3$ ,  $\text{NH}_3$  gas is sent through Absorber Gas separator to Secondary Absorber System. Carbonated MEA solution is discharged through Stripper Reboiler to Regenerator System by regenerator feed pump.

#### 2) Regenerator System

Carbonated MEA solution in absorber stripper system is pumped to Regenerator. On the way in MEA Heat Exchanger waste heat is retained from the MEA solution coming back from Regenerator by passing steam through Regenerator Reboiler of  $\text{CO}_2$  is stripped.

$\text{CO}_2$  gas out of Regenerator top is cooled by Regenerator condenser for separating  $\text{H}_2\text{O}$  and is sent to  $\text{CO}_2$ -compressor section after scrubbing with water in  $\text{NH}_3$  scrubber. Water condensated contains small quantities of  $\text{NH}_3$ ,  $\text{CO}_2$  and MEA. The condensate is once stored in Regenerator Condensate Tank. Nearly half portion of this condensate is returned as reflux to Regenerator by regenerator reflux pump.

The remaining condensate is sent to waste in controlled amounts. MEA solution passing downward from Regenerator to Regenerator Reboiler is heated up by  $10 \text{ kg/cm}^2$  steam. Here  $\text{CO}_2$  gas is sufficiently stripped and the vapor is returned to Regenerator for heating up.

Most part of the solution of Regenerator Reboiler is returned to MEA Surge Tank. On the way heat is added to the feed solution to Regenerator and then is cooled up to the normal temperature by cooling, water passing through MEA cooler.

The remaining solution is sent to purifier where it is heated up and evaporated and vaporised MEA is returned to Absorber Stripper. Decomposition products is purged by steaming at times.

### 3) Secondary Absorber System

A small quantity of  $\text{CO}_2$  containing in  $\text{NH}_3$  gas from Absorber Stripper system is scrubbed with  $\text{NH}_3$  aqueous solution in secondary absorber and  $\text{NH}_3$  gas is sent through secondary absorber, secondary absorber gas separator to compressor system.

Heat generated when  $\text{NH}_3$  and  $\text{CO}_2$  are absorbed by water is cooled by brine passing through Secondary Absorber Cooler.

#### 4) Compressor System

$\text{NH}_3$  gas delivered from secondary absorber system is compressed by recovered  $\text{NH}_3$  compressor, liquefied by recovered  $\text{NH}_3$  condenser and sent to synthesis and separation section.

#### 2.2.10 Prilling and Cooling Process

##### 2.2.10.1 Prilling System

Urea solution concentrated in Evaporator is sent to the top of Prilling Tower. This pipe line is branched in two directions at the top and bottom, each branched line is connected to three spray nozzles. The reason why the line is branched in two directions is to equalize delivery pressure at each nozzle. Urea solution sprayed from nozzle plates each having 600 pin holes is cooled to form solid prills, falling down inside the Tower.

##### 2.2.10.2 Cooling System

Urea prills free of oversized prills, which have been screened off by Prills Screen, are lifted up by Prills Elevator and chuted into Prills Cooler. The Prills Cooler serves to lower the temperature of the prills to below  $50^{\circ}\text{C}$ . To improve the cooling effect, air is sent counter-current to the product stream. Air sucked at the air conditioned bagging room passes through the cooler into prills dust collector, where the dust is separated

and is exhausted to the atmosphere by means of Prills Dust Collector Fan. The prills passing from the Prills Cooler through Prills Storage Conveyor and lifted by Cooled Prills Elevator and fall into Prills Storage Screen, where oversized prills above 6 mesh and undersized below 20 mesh are screened off. The correct size prills are lifted by Prills storage elevator for delivery to the Prills Storage and bagging section.

CHAPTER - 3  
UREA FERTILIZER FACTORY, GHORASHAL  
(UFFG)

### 3.1 History of UFFG<sup>4</sup>

The construction of this factory was entrusted to M/s. Toyo Engineering Corporation, Japan. The work started in September 1967 and was completed in May 1970.

The chief raw material for the factory is natural gas which is transported from Titas field to the factory through a 30 mile 14" pipe line. The factory is self-sufficient in utilities, like water, power and steam.

- 1) Capacity of production Urea 3,40,00 MT/yr.
- 2) Daily consumption of raw materials and utilities
  - a) Titas gas - 39 million cu.feet, 23 million cu.feet for generating electricity
  - b) Electricity- 300 MWH.
  - c) Water- 7 million gallone
- 3) Date of commencement of the project -- Sept. 1967
- 4) Date of completion of the project -- May 1970
- 5) Date of commencement of trial production- July 1970
- 6) Date of commencement of commercial production -- Oct. 1970

### 3.2 Description of UFFG<sup>4</sup>

Main ingredients for the manufacture of Urea fertilizer are liquid ammonia and carbon-di-oxide gas. These two components are manufactured from the natural gas in the Ammonia-Plant of the factory.

## AMMONIA UNIT

The raw materials used in UFFG for the synthesis of ammonia are Natural Gas(NG) air and water.

### 3.2.1 Purification of NG

#### 3.2.1.1 Process description

NG contains some dust and oil. It is cleared through Dust Separator, Venturi Scrubber and Filter. NG first enters one of the Dust Separators where water is sprayed counter-currently to the uprising gas and bigger particles of dust are washed away. The gas enters venturi scrubber and further washed with water and then all dust is removed in NG cyclone. Then the gas is passed through one of the NG Filters to get it more cleaned. The filters are switchover type and can be back washed periodically. The gas is later divided into two streams. One stream is used as the raw material for urea fertilizer and known as Process Natural Gas(PNG). The other stream is called Fuel Natural Gas(FNG) and used as fuel in the process.

### 3.2.2 Desulphurisation Process

#### 3.2.2.1 Process Description

The natural gas which is used as process natural gas (PNG) is compressed upto  $35 \text{ kg/cm}^2$  by NG Booster compressor. It is then passed through feed stock steam heater which is of shell and tube type. PNG is heated upto  $400^\circ\text{C}$  by flue gas from Primary Reformer. NG contains sulphur which is poisonous to

catalyst. It is removed by passing through Desulphurisation vessel(DSV) containing ZnO. From Feed Stock Steam Preheater, PNG enters the load vessel of Desulphurisation vessel(DSV) where sulphur is removed by ZnO Catalyst. Then it enters the guard vessel of DSV for final removal of remaining sulphur.

### 3.2.2.2 Process Chemistry

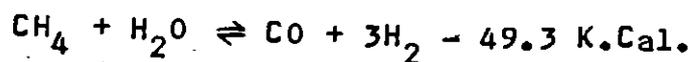
Maximum 1 ppm of sulphur, is allowable to reforming catalyst. But feed stock contain about 10 ppm of sulphur. In presence of ZnO Catalyst  $H_2S$  can be removed at the temperature of  $200^{\circ}C$  (minimum) of organic sulphur at  $370^{\circ}C$  (minimum). Maximum tolerable temperature of ZnO catalyst is about  $420^{\circ}C$  above which catalyst may be pulverised and the vessels may be damaged. So catalyst temperature is maintained at  $400^{\circ}C$  for removal of all kinds of sulphur. The reaction occurs in the following way:



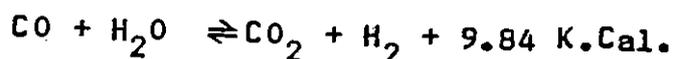
### 3.2.3 Reforming Process

#### 3.2.3.1 Primary Reforming

Minimum temperature methane-steam reforming in presence of nickel catalyst is 538. The reaction occurs in the following way:



Co produced by the above reaction further reacts with steam which is called water gas shift reaction shown as follows:



The reactions as a whole are endothermic. So it is necessary to supply heat from outside.

### 3.2.3.2 Secondary Reforming

In the secondary reformer methane-steam reaction proceeds further to produce Hydrogen and Carbon-monoxide. The necessary heat for this endothermic reaction is obtained by combustion of  $H_2$  with process air which is added at the inlet of secondary reformer. The oxygen of the process air is completely burned and the temperature of the gas mixture rises. This hot gas mixture passes through the nickel catalyst bed and loses temperature due to endothermic reaction. The addition of air is controlled by  $N_2$  and  $H_2$  ratio (1:2:8) required in the ammonia synthesis section.  $CH_4$  content (0.3%) in the secondary reformer outlet gas is controlled by supplying heat in the Primary Reformer.

### 3.2.3.3 Heat Recovery Process

Both the flue gas leaving this furnace and the reformed gas leaving the Secondary Reformer are at the temperature of about  $1000^{\circ}C$ . It is economical to recover this heat.

### 3.2.3.4 Heat Recovery from Flue Gas

The flue gas leaves the reformer furnace at about  $1000^{\circ}C$ . This heat is utilised in generating extra high pressure steam in Flue Waste Heat Boiler, in superheating process steam in First Steam Superheater, in superheating high pressure steam in second steam superheater, in preheating process air in

Process Air Preheater, in heating combustion air in combustion Air Preheater and in heating PNG in Feedstock Preheater.

### 3.2.3.5 Heat Recovery Iron Process Gas

As the temperature of secondary reformer outlet gas is about  $1000^{\circ}\text{C}$  which is too high for the next High Temperature Shift Converter, Reformed Gas Waste Heat Boiler is installed for recovery this extra heat.

### 3.2.3.6 Process Description

PNG from DSV mixes with superheated steam and then enters primary reformer (PR) which consists of refractory lined box-like structure in which 300 tubes are suspended vertically. In these tubes reforming catalyst (Ni on calcium aluminate) is uniformly packed. The gas mixture passes through these tubes from top to bottom. There are 78 burners located in the top of the furnace and 3 auxiliary burners in the gas duct, firing downwards. These burners have been designed to operate on natural gas. Combustion air is supplied from Air-blower through combustion air preheater at about  $300^{\circ}\text{C}$ . The reformed <sup>gas</sup> from Primary Reformer is mixed with process air from Air Compressor (ATB) preheated to about  $538^{\circ}\text{C}$  through process Air Preheater at the Secondary Reformer (2RF) where partial combustion occurs and required, nitrogen for ammonia synthesis is also supplied. The reformed gas comes out from 2RF with the methane content of about 0.3% and temperature of about  $1000^{\circ}\text{C}$  and then enters the Reformed Gas Waste Heat Boiler where it is cooled to less than  $500^{\circ}\text{C}$  with the generation of

100 kg/cm<sup>2</sup> steam and then goes to shift conversion process.

The secondary reformer consists of carbon steel shell lined with refractory inside and water jacket outside. The upper chamber is for combustion and the lower part is packed with catalyst. The combustion chamber is of narrow cylindrical cone to ensure rapid mixing of the gases and subsequent complete reaction with oxygen.

The gas from secondary reformer passes through the Reformed Gas Waste Heat Boiler which are of water tube and natural circulation type with common steam Drum. Then the gas proceeds to Shift Conversion Process.

On the otherhand the flue gas of the Primary Reformer passes through the Flue Gas Waste Heat Boiler, First Steam superheater, Feed Stock Preheater and Combustion Air Preheater and is exhausted to the atmosphere through the chimney stack by Induced Draft Fan.

### 3.2.3.7 Operating Conditions

Steam/Carbon ratio	3.5
Primary Reformer outlet process gas temperature	818°C
Secondary Reformer outlet process gas temperature	1000°C
Steam press. in steam Drum	105 kg/cm <sup>2</sup>
Ist steam superheater outlet steam temp.	475°C
Process Air Preheater outlet air temp.	538°C.

Feed stock preheater outlet natural gas temp.	400°C
Primary Reformer outlet flue gas temp.	1000°C
2nd Steam Superheater outlet steam temp.	490°C

Composition of Primary Reformer outlet gas by volume

- 1) CH<sub>4</sub> 10%(dry)
- 2) Steam/Dry gas mole ratio 0.755.

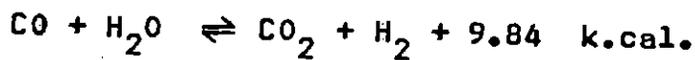
Composition of Secondary Reformer outlet gas by vol.

- 1) CH<sub>4</sub> - 0.2% to 0.6% (dry)
- 2) Steam/dry gas mole ratio -0.569.

### 3.2.4 Shift Conversion Process

#### 3.2.4.1 Shift Reaction

The shift reaction occurs under proper temperature in presence of suitable catalyst in the following manner:



Conversion of carbon monoxide to carbon-di-oxide is done in two steps economically. Primary shift conversion is done in the High Temperature. Shift Converter at the temperature of 350 to 450°C in presence of iron chromium catalyst and Secondary Shift-Conversion is done in the low temperature. Shift Converter at the temperature of 200 to 230°C in presence of zinc-copper catalyst.

### 3.2.4.2 High Temperature Shift Conversion

In this process High Temperature Shift Converter (HTS) and quenching system are provided. HTS is packed with Iron-Chromium catalyst developed by CCI (Chemical Company Industries). Quenching system is provided with venturi and Quenched Gas Separator for controlling the temperature of HTS inlet gas and for increasing steam quantity in this gas required for the reaction. The carbonmonoxide conversion in this vessel is about 75.6% on the basis of carbonmonoxide in the inlet gas.

### 3.2.4.3 Low Temperature Shift Conversion

This unit consists of (1) Low Temperature Shift conversion (LTS) packed with zinc copper catalyst for CO conversion (2) Process Gas Heat Exchanger (PGHEX) for controlling Methanator inlet gas temperature (3) First converted Gas Economyser which is used for heat recovery and for controlling the temperature of LTS inlet gas and (4) Second converted gas Economyser for controlling the temperature to CO<sub>2</sub> removal unit. There is provision for reduction of LTS catalyst. This catalyst is very sensitive to sulphur. Therefore a guard bed of zinc-oxide is provided at the top of LTS to remove sulphur. The carbon monoxide conversion in this vessel is about 87.7% on the basis carbon monoxide in the inlet gas. The total conversion yield is about 97% on the basis of CO content in HTS inlet gas. LTS catalyst being less thermostable than HTS catalyst, the outlet gas temperature should be maintained below 280°C.

#### 3.2.4.4 Process Description

The gas coming out of the reforming unit is cooled by quench water in venturi. Then it enters Quench Gas separator for separation of condensate, if any. Then the gas enters HTS for shift-conversion. The HYS is a vertical cylindrical vessel packed with catalyst in two layers. First converted Gas Economiser is also of horizontal shell and tube type. Boiler feed water flows in the shell side and process gas in the tube side. LTS is a vertical cylindrical vessel containing ZnO catalyst in the top layer and zinc-copper catalyst in the lower in two layers.

There is a condensate cooler in condensate system. Condensate from this cooler is used for washing the gas from CO<sub>2</sub>-Absorber and CO<sub>2</sub>-Regenerator and then used as the make up water for CO<sub>2</sub> removal process. In this unit condensate pump is used for pumping the process condensate from Process Gas Knock Out Drum to Venturi and condensate cooler.

#### 3.2.4.5 Operating Condition

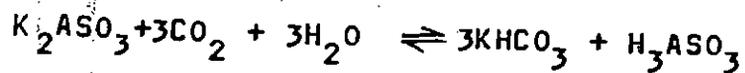
HTS inlet gas temperature about	365°C
inlet steam/gas ratio about	0.67
inlet steam/CO ratio about	5
shift conversion rate	75%
outlet gas temperature	435°C
LTS inlet gas temperature	219°C
inlet steam/gas ratio about	0.52

shift conversion rate about	88%
outlet gas temperature about	235°C
outlet CO concentration	0.35 to 0.5%

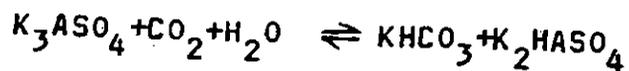
### 3.2.5 CO<sub>2</sub> Removal Process

#### 3.2.5.1 GV Process

CO<sub>2</sub> is removed by Giammarco Ve trocoke (GV) solution in this plant. GV solution is prepared by potassium Carbonate, arsenic trioxide and water. Concentrations of potassium and arsenic are 200 g/l and 140 g/l as K<sub>2</sub>O and As<sup>+3</sup> respectively. As<sup>+2</sup> is used to prevent corrosion, concentration of As<sup>+5</sup> is 10 g/l. Concentration of As<sup>+5</sup> is 10 g/l. K<sub>2</sub>CO<sub>3</sub> solution absorbed CO<sub>2</sub> and anhydrous arsenous acid accelerates absorption and desorption velocities. Absorption of CO<sub>2</sub> by solution proceeds basing the following equilibrium reaction:



Arsenic pentavalent also takes part in absorption of CO<sub>2</sub> in addition to its anticorrosive activity in the following manner:



GV solution can be regenerated by decreasing partial pressure of CO<sub>2</sub> in the gas phase, steam stripping is done. Regeneration by steam is carried out at the boiling point in the CO<sub>2</sub> Regenerator and Reboiler. Due to the presence of AS<sup>+5</sup> corrosion does not occur even at high temperature. If concentration of AS<sup>+5</sup> decreases, air is charged to maintain it.

### 3.2.5.2 Process Description

Process Gas from CO conversion unit enters Process Gas Reboiler for providing the heat required for regeneration of GV solution. Then it enters Process Gas Knock Out Drum for separation of condensate. The gas from Process Gas K.D. Drum passes through Absorber for removing CO<sub>2</sub> contacting GV solution counter-currently. Then it enters Absorbers Effluent cooler for cooling. Then the gas passes through Mist Trap where mist is removed contacting with condensate counter-currently. Then the gas goes to Methanator unit with 0.13% of CO<sub>2</sub>.

GV solution from Absorber bottom goes to solution Heat Exchanger for gaining heat for regeneration after partial filtration in side stream Filter. Then the solution enters regenerator at the top for regeneration of solution. The rich solution is regenerated by the counter-current contact of uprising steam produced in the reboiler. The solution flows catching plate of regenerator goes to the bottom of regenerator through reboiler. Then solution from the bottom of regenerator is pumped by solution pump upto 35 kg/cm<sup>2</sup> and sent to absorber top through flow control valve, solution Heat Exchanger and solution cooler.

CO<sub>2</sub> released from the top tray of regenerator where it is washed by condensate, enters the washing part of CO<sub>2</sub> Direct cooler. Here it is washed again by condensate from Mist Trap.

The washed CO<sub>2</sub> gas is introduced to the lower part of Direct Cooler where CO<sub>2</sub> is cooled to ambient temperature in contact with cooling water counter-currently. CO<sub>2</sub> from Direct cooler is sent to Urea Plant with a purity of 98.5% (dry gas basis).

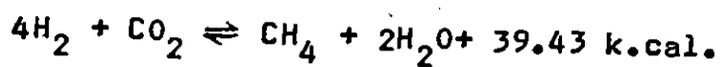
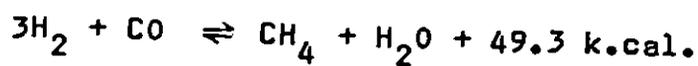
### 3.2.5.3 Operating Conditions

K.O. Drum inlet gas Pressure	26 kg/cm <sup>2</sup>
K.O. Drum inlet gas temperature	120°C
Absolute inlet solution temperature	63°C
Regenerator bottom solution temperature	115°C
CO <sub>2</sub> in purified process gas	0.13%
Regenerator top CO <sub>2</sub> pressure	0.4 kg/cm <sup>2</sup>
Recovered CO <sub>2</sub> temperature	40°C
Recovered CO <sub>2</sub> purify above	98.5%

### 3.2.6 Methanation Process

#### 3.2.6.1 Reaction Mechanism

In order to complete the removal of CO and CO<sub>2</sub> from synthesis gas by Hydrogenation into steam and methane, methanation is used as the final gas purification process. The reactions occur in the following manner:



The reactions are exothermic. So outlet gas temperature rises and carbon oxide are reduced to about 10 ppm. The catalyst is Ni on calcium aluminate.

When the inlet concentration of carbon oxide is 1% the temperature rise is about 65°C. Therefore carbon oxide concentration in the gas entering Methanator should be kept less than 1.5% and normally about 0.6%. If temperature increases to above 400°C, Methanator vessel may be damaged.

### 3.2.6.2 Process Description

The gas from CO<sub>2</sub> removal unit enters the Process Gas Heat Exchanger of the CO-Conversion unit and is heated to 300°C by HTS outlet gas and then enters Methanator. The process gas after removal of carbon oxide in the Methanator heats up the boiler feed water in Methanator Effluent Economyser and is then cooled to ambient temperature in Gas Final Cooler. The condensed water in the gas is removed in Synthesis Gas Knock Out Drum and the gas is sent to Synthesis Gas Compressor.

### 3.2.6.3 Operating Condition

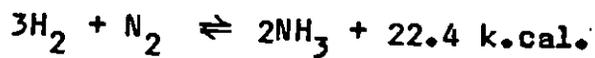
Methanator inlet gas temperature	about 300°C
Methanator outlet gas temperature	about 332°C

Carbon Oxide in Methanator outlet gas	- about 10ppm.
Synthesis Gas Knock Out Drum outlet gas temperature	- 40°C

## 3.2.7 Ammonia Synthesis Process

### 3.2.7.1 Synthesis Reaction

Ammonia synthesis reaction occurs on the iron catalyst promoted with various kinds of metal and is indicated by the following equation:



In addition to nitrogen and hydrogen with ratio of about 1:3, the refined gas has about 1.2% of  $\text{CH}_4$  and  $\text{Ar}$ . But  $\text{CH}_4$  and  $\text{Ar}$  dilute the gas mixture of  $\text{N}_2$  and  $\text{H}_2$  and reduces the partial pressure of effective components and lowers synthesis efficiency. Therefore, the continuous purging of gas is maintained from the synthesis gas loop to control the concentration of inert gases.

### 3.2.7.2 Process Description

The process gas compressed by Synthesis Gas Compressor is mixed with the effluent recycle gas (from the synthesis converter) at the outlet of synthesis Economizer is cooled by the cold gas in the Synthesis Hot Exchanger.

Then it enters Synthesis Boiler Feed Water Heater and Synthesis First Cold Exchanger for Cooling. Entering First Ammonia Cooled Condenser it is cooled by the latent heat of vapourisation liquid ammonia from the refrigeration unit. Then synthesized ammonia present in the gas is condensed. The gas separated from ammonia in Ammonia separator enters Second Cold Exchanger from gaining heat from incoming gas. Then the gas is heated up in synthesis first cold Exchanger and enters recirculator where it is pressurised. After being heated up in synthesis Hot Exchanger, it enters Synthesis Converter. The temperature of the gas to synthesis converter is adjusted by the by pass valve for synthesis hot exchanger. After synthesis

reaction is Ammonia converter, effluent gas enters Synthesis Economiser for exchanging heat with boiler feed water. The gas from Synthesis Economiser, after being mixed with the synthesis make up gas, enters Synthesis Hot Exchanger and circulate in the synthesis system described above. About 7.2% of the total gas converted into  $\text{NH}_3$  in one pass through the converter. From  $\text{NH}_3$ - Separator, liquid ammonia passes through Let Down Tank and then stored into Ammonia Storage Tank from where  $\text{NH}_3$  is supplied for Urea Plant.

#### UREA UNIT

The urea plant is divided into three section, namely-

- a) Synthesis Section
- b) Decomposition and Purification Section
- c) Recovery Section.

#### 3.2.8 Synthesis Process

##### 3.2.8.1 Process Description

The compressed  $\text{CO}_2$  gas, pre-heated Ammonia and recycle solution are fed into the Urea Reactor (Autoclave) where urea synthesis is carried out under high pressure and temperature  $\text{CO}_2$  gas is compressed upto  $230 \text{ kg/cm}^2$  by a two staged compressor and is directly fed with gaseous state into the Reactor. Liquid  $\text{NH}_3$  from Recover ammonia reservoir is fed into Reactor at  $230 \text{ kg/cm}^2$  by plunger-type ammonia

feed pump. The recycle carbonate solution from high pressure absorber cooler is also pumped upto  $230 \text{ kg/cm}^2$  by a plunger type Recycle solution feed pump and is fed into Reactor.

### 3.2.8.2 Operating Condition

- 1)  $\text{CO}_2$  inlet press of the Reactor =  $230 \text{ kg/cm}^2$   
Temperature =  $120-130^\circ\text{C}$
- 2)  $\text{NH}_3$  inlet press of the Reactor =  $230 \text{ kg/cm}^2$   
Temperature =  $70-90^\circ\text{C}$
- 3) Reactor press =  $230 \text{ kg/cm}^2$   
Temperature =  $190-195^\circ\text{C}$
- 4) Mol. ratio :  $\text{NH}_3 : \text{CO}_2 = 4:1$

### 3.2.8.3 Reaction Mechanism

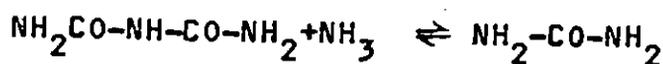
The first reaction in the Reactor is formation of ammonia carbonate from ammonia and carbon-di-oxide



This reaction release to large heat. Next a part of ammonium carbonate dihydrates, to be water and urea.



The reaction is also equilibrium reaction of urea and biuret in Reactor.



### 3.2.9 Decomposition and Purification

#### 3.2.9.1 Process Description

The reaction product consisting of urea, free Ammonia,

ammonium carbamate and water is reduced its pressure from  $230 \text{ kg/cm}^2$  to  $18 \text{ kg/cm}^2$  and enter into the High Pressure Decomposes.

This decomposer is special type, the upper part of which is a separator to separate flush gas from effluent mixture, the lower part is a distillator with steam heater at the bottom. After separation of flush gas, the Urea solution overflows down at into the lower part and is heated further to decompose carbamate, the decomposed carbamate is distilled to reduce moisture content in the gas during its ascending up through sieve trays.

Here, about 90% of free ammonia and carbamate will vaporize and decompose to ammonia and  $\text{CO}_2$  gas. Small quantity of Urea hydrolyzes to ammonia and  $\text{CO}_2$  and the liquid enters into low pressure Decomposer. The gas consisting of  $\text{NH}_3$ ,  $\text{CO}_2$  and water from High Press. Decomposer into High Press absorber cooler.

The liquid from H.P. Decomposer enters into low pressure decomposer where more gas is removed.

The composition product which is mixture gas of  $\text{NH}_3$ ,  $\text{CO}_2$ ,  $\text{H}_2\text{O}$  flows in low pressure absorber from low pressure decomposer. The liquid of low press. Decomposer enters into Gas separator. In Gas Separator, it is heated by steam. In this step,  $\text{NH}_3$ ,  $\text{CO}_2$  and little water is vaporized and sent to gas condenser. The operating condition of gas separator is  $0.3 \text{ kg/cm}^2$   $112^\circ\text{C}$ .

The urea solution of gas separator enters into concentrator through the circulation pump for crystallizer. Almost all water in the supplied urea solution are vaporised at this condition. From concentrator urea slurry goes to crystallizer where crystal of urea deposited.

An agitator is installed in crystallizer to prevent agglomeration of urea crystal.

Urea slurry having nearly 35% of crystal is sent to centrifuge by slurry feed pump.

In order to reduce the biuret content, water raising of cake is necessary in centrifuge. Urea solution enters into centrifuge and slurry is divided into crystal and mother liquid by centrifugal force. Mother liquid is gathered in the Mother Liquor Tank and 90% is returned back to concentrator and remaining 10% is recycled to recovery section as absorbent.

Crystal urea from urea from centrifuge contains 1.0-1.5% of  $H_2O$  and 0.1-0.15% of biuret.

#### 3.2.9.2 Drying

Crystal urea sent to Dryer is dried by hot air of about  $110^{\circ}C$  and dried upto 0.1-0.2% of moisture and then caught in the cyclone.

#### 3.2.9.3 Prilling

The crystal urea from the cyclone are fed to melter where the crystal is melted. From melter, molten urea flows

down to head tank. The liquid in head tank falls down from distributor as rain drops. During falling down in the tower the molten urea is prilled and cooled by the ascending air which comes from blower for fluidizing cooler and from the bottom of the prilling tower as natural draught.

### 3.2.10 Recovery Process

#### 3.2.10.1 Process Description

Mother liquor from mother liquor tank and pure water are fed to Gas condenser continuously as absorbent. The gas from Gas Separator is absorbed in Gas condenser and the non-condensable inert gas is purged to atmosphere from the top. The liquid in gas condenser is pumped up to the low pressure absorber as absorbent.

The liquid in L.P. Absorber is pumped up and is fed to mixer in which the liquid is mixed with dig.  $\text{NH}_3$  to adjust its composition, then cool and finally fed to packed section of H.P. Absorber. High Pressure Absorber consists two part, the lower part is packed section to wash away. Most part of  $\text{CO}_2$  in ascending gaseous ammonia, the upper part is tray tower to purify ammonia in scrubbing trace of  $\text{CO}_2$  in decomposed gas from H.P. Decomposer is absorbed in cooler and packed section and the effluent gas from the top of the tower, almost pure ammonia is condensed to liquid state in Ammonia condenser after mist separator. The composition of concentrated carbamate solution in H.P. Absorber cooler is as follows:

Urea	- 8.3%	}	Water	- 25.4%
NH <sub>3</sub>	- 36.0%		Biuret	- 0.3%
CO <sub>2</sub>	- 30.0%			

## CHAPTER 4

## COMPARISON OF PROCESSES USED IN UFFG &amp; NGFF

#### 4.1 Background of Process Comparison

Various processes exist in urea production in the world. But the basic theory of production of urea remain same. In different processes, the influencing factors such as pressure, temperature catalyst etc. are also different. In advance of science and technology, the influencing factors is changing for better benefits. In UFFG and NGFF, different process are used in urea production and their influencing factors are also different.

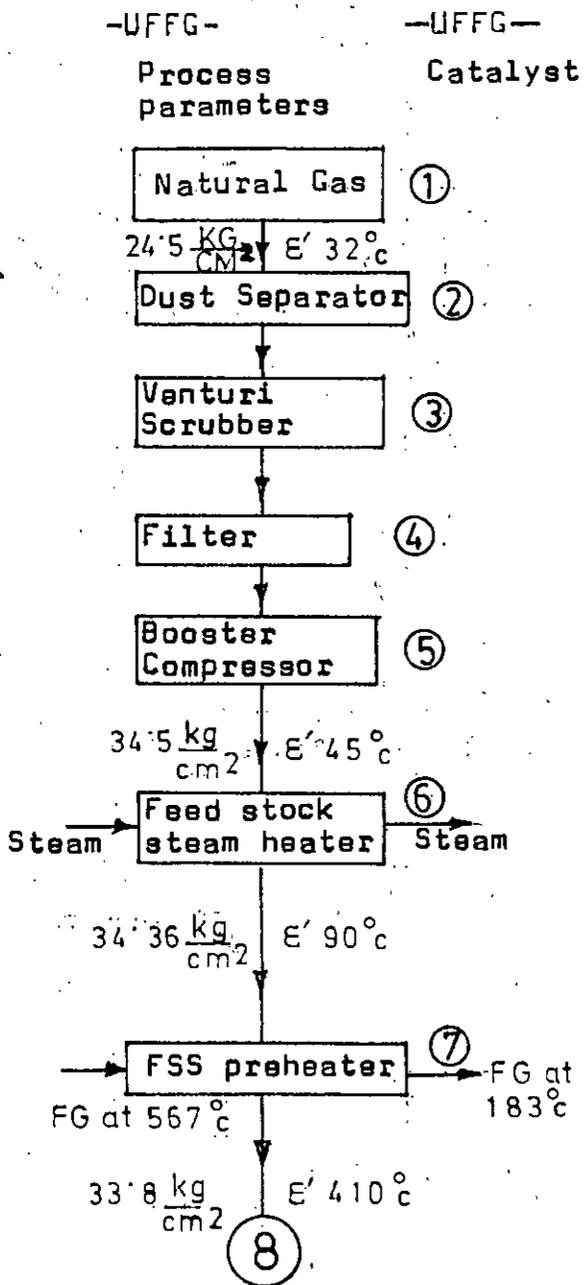
To find out the process restrictions and influencing factors, comparison of above processes is necessary. This comparison will help in analysing the down-time as well as in finding the causes of failure of different machineries of the above factories.

#### 4.2 Comparison of processes

From the detail description of the processes in the previous chapter, it is seen that,, in one process, the reformation of NG takes place at low pressure, synthesis of  $N_2$  and  $H_2$  takes place at higher pressure than that of another process. Different types of catalyst are used for same reaction. In a process the whole  $CO_2$  is taken from the PNG but in another process it is taken from PNG and FG. For  $CO_2$  removal liquid used in NGFF differs from than that of UFFG. Conversion rate in  $NH_3$  synthesis, addition moles of  $NH_3$  with the mols of  $CO_2$  for urea reaction are also differ

in both the process. But some stages of the production of urea remain same in each process.

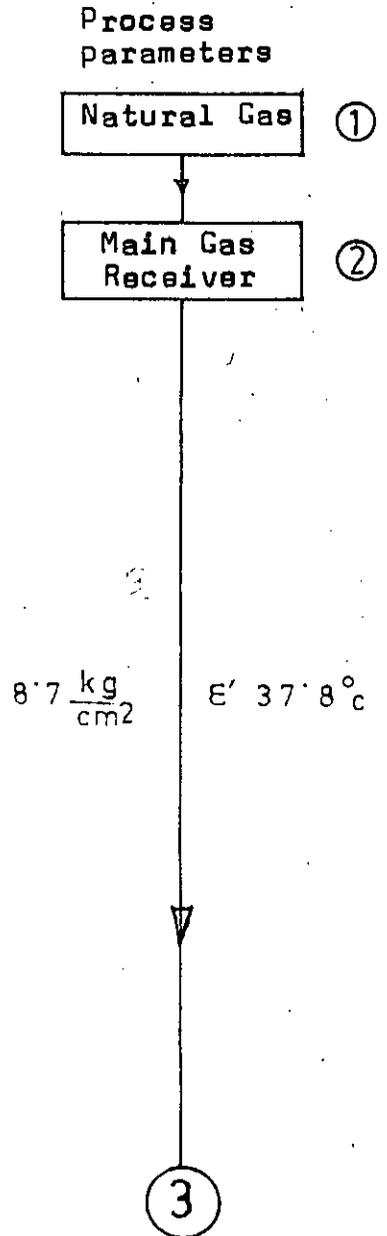
Finally, to find out the above differences as well as similarity in the processes, the stages of urea production are listed below stagewise



**-UFFG-**

Catalyst

**-NGFF-**



Catalyst

**-Diff. Between-**

UFFG                      NGFF

— UFFG —

— NGFF —

— Diff. between —

Process parameters

Catalyst

Process parameters

Catalyst

UFFG

NGFF

\* 1

Desulphurization vessel

⑧ Zinc oxide

Organic sulphur removal vessel

③ Activated Charcoal

- |   |  |
|---|--|
| a) Separated by Chemical means.                       | a) Separated by mechanical means.                          |
| b) At high temp.                                      | b) At low temp.  |
| c) At outlet sulphur content is allowable upto 1 ppm. | c) At outlet sulphur content is allowable upto 1 to 2 ppm. |

32.8 kg  
E' 400°C

Steam

8.6 kg/cm<sup>2</sup>  
E' 37.8°C

Mixed gas preheater

④ Steam

To ⑨

8.2 kg  
E' 399°C

\* 2  
FG to ④

1RF

⑨ Heating by Fuel

Ni on calcium aluminate

\* 3  
FG to ②

1RF

⑤ Heating by Fuel

22-25% Ni deposited on the refractory of MgO, Al<sub>2</sub>O<sub>3</sub>, CaAl<sub>2</sub>O<sub>3</sub> & CaSiO<sub>3</sub>.

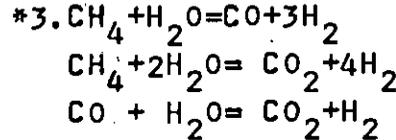
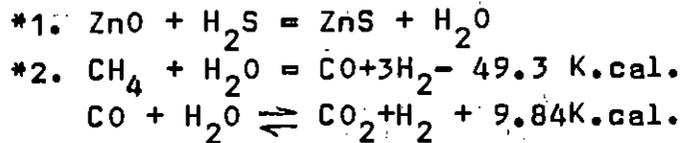
- |   |  |
|---|--|
| a) Steam/Carbon ratio is 3.5:1.   | a) Steam/Carbon ratio is 1:1.  |
| b) At outlet process gas temp. is 818°C and the flue gas temp. is 1000°C. | b) At outlet process gas temp. is 620°C and flue gas temp. is 800°C. |
| c) CH <sub>4</sub> content at outlet is 10%.                              | c) CH <sub>4</sub> content at outlet is 7.7%.                        |

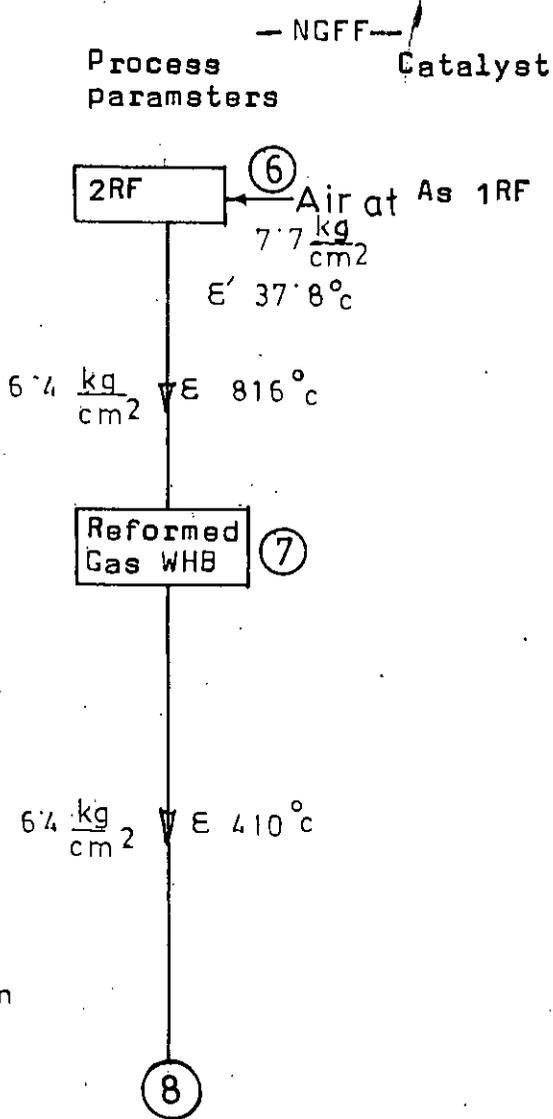
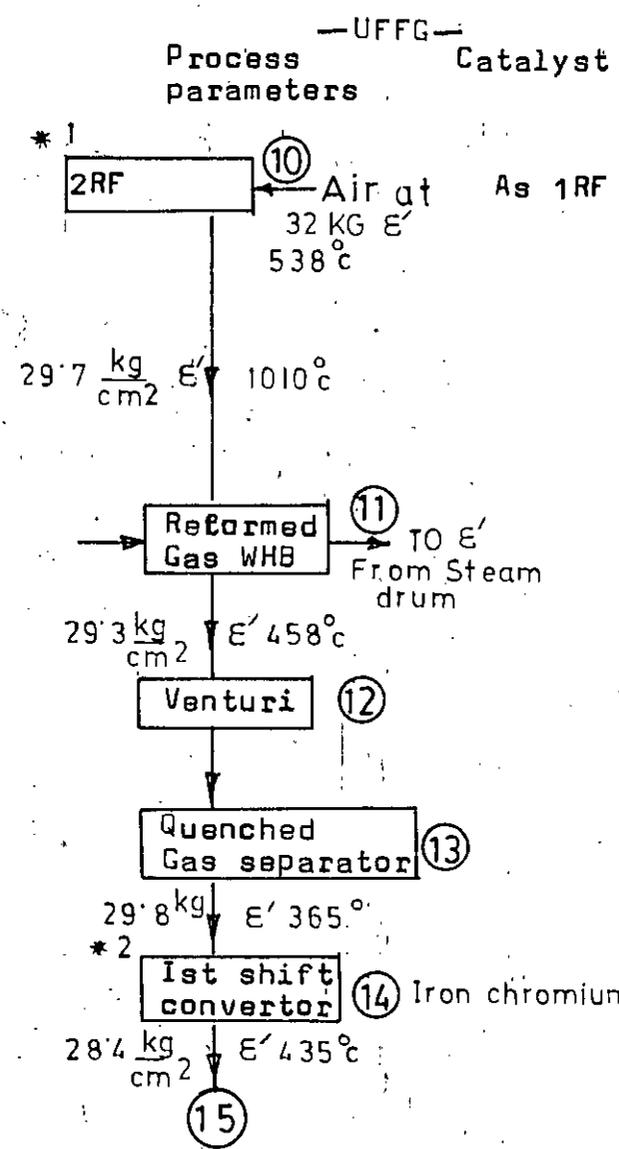
30.3 kg/cm<sup>2</sup>  
E' 817.8°C

⑩

6.5 kg/cm<sup>2</sup>  
E' 620°C

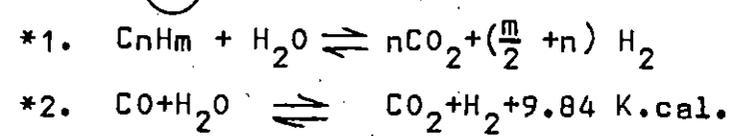
⑥





- Diff. between—
- UFGG                      NGFF
- a) At outlet process Gas temp. is 100°C.
  - a) at outlet process gas temp. is 816°C.
  - b) CH<sub>4</sub> content at outlet is 0.2% to 0.6%.
  - b) At outlet CH<sub>4</sub> content is 0.3%.

In Ist shift converter the conversion rate of CO into CO<sub>2</sub> is about 75.6%.



—UFFG—

—NGFF—

—Diff. between—

Process parameter

Catalyst

Process parameters

Catalyst

UFFG

NGFF

Process Gas Heat Exchanger (15)

317°C

Ist converted Gas Economizer (16)

27.8 kg/cm<sup>2</sup> 219°C

\* 1 2nd Shift convertor (17)

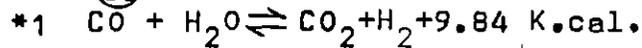
27kg/cm<sup>2</sup> 235°C

2nd converted Economizer (18)

198°C

Process Gas Reboiler (19)

26.3 kg/cm<sup>2</sup> & 120°C (20)



Zinc-Copper

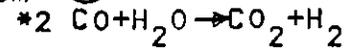
\* 2

CO convertor (8) Fe<sub>3</sub>O<sub>4</sub>

6.3 kg/cm<sup>2</sup> & 243°C

Gas from (3) Mixed Gas preheater (4) Gas to (5)

6.3 kg/cm<sup>2</sup> & 243°C (9)

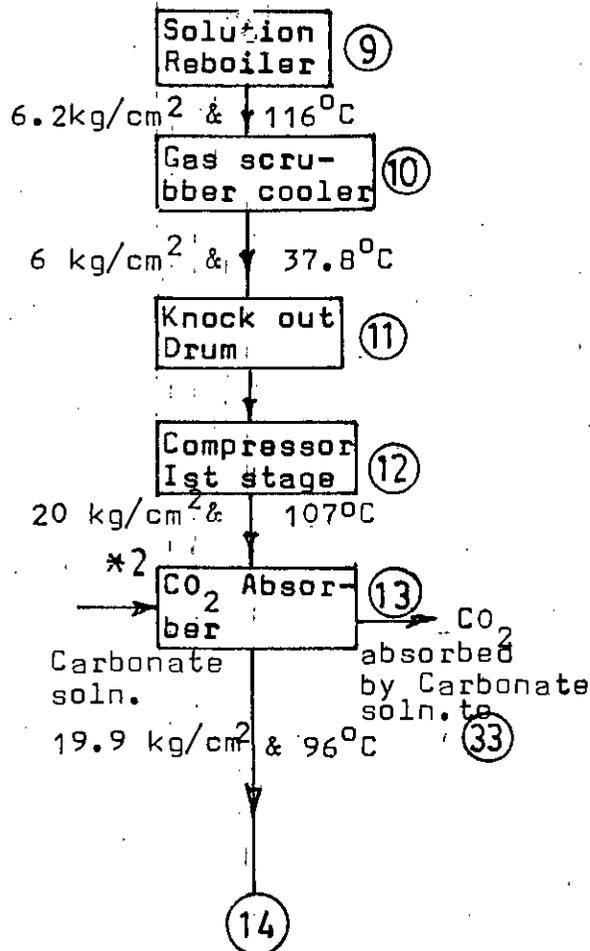
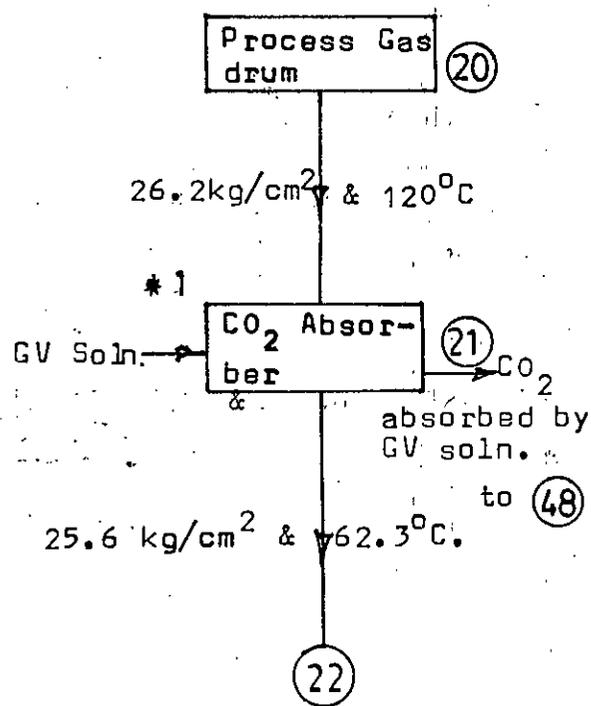


- a) In 2nd shift convertor, the conversion rate of CO into CO<sub>2</sub> is 87.7%.
  - b) The total conversion yield is 97%.
  - c) The total content of CO in the outlet is 0.35 to 0.5%.
- a) The total conversion yield is 60%.
  - b) The total content of CO in the outlet is 1.2%.

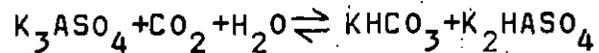
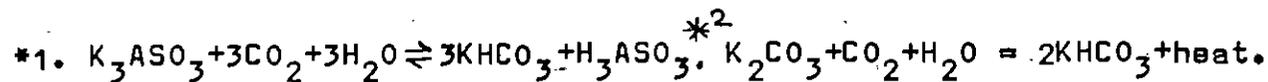
—UFGF—  
 Process parameter Catalyst

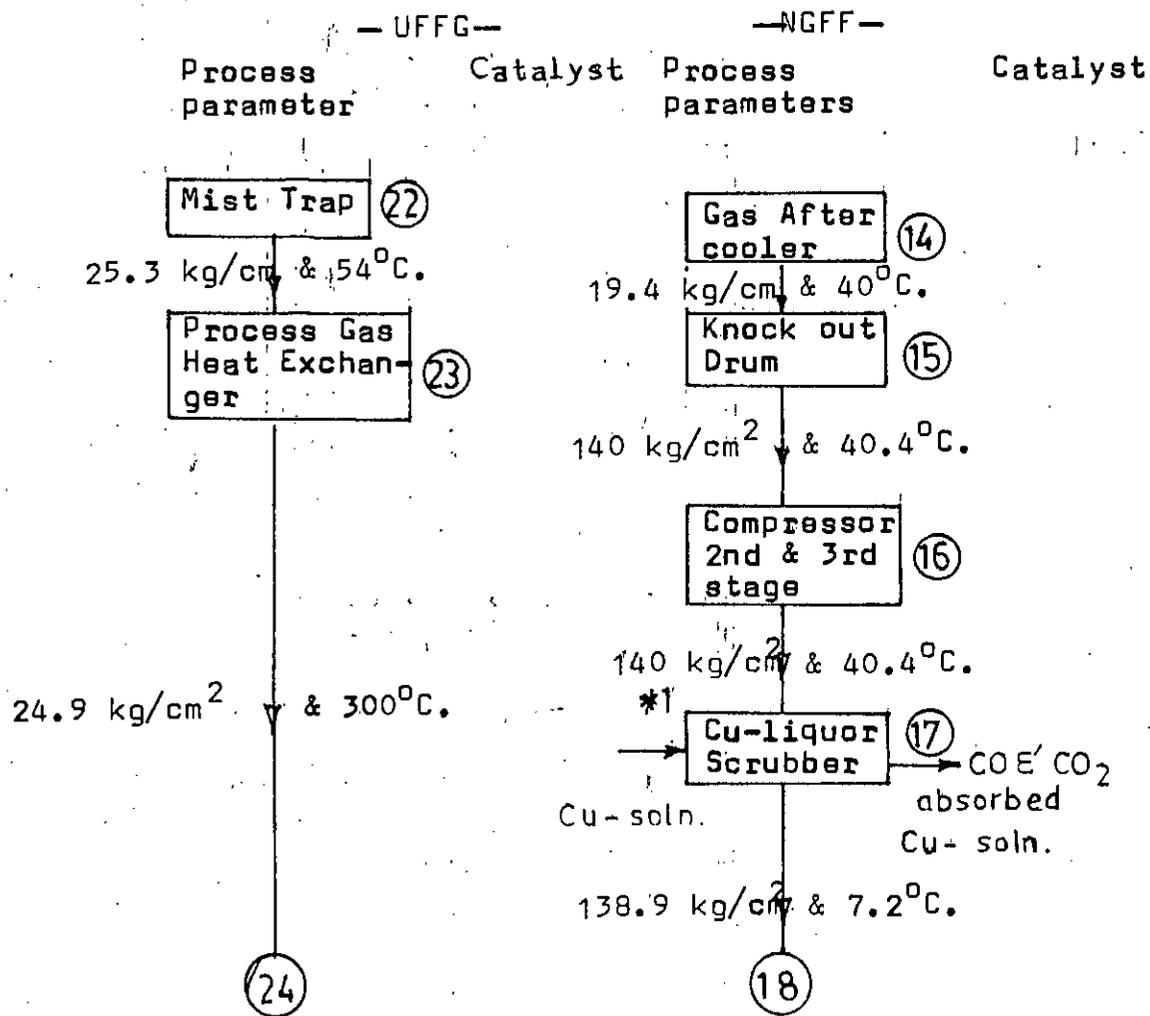
—NGFF—  
 Process parameters Catalyst

— Diff. between —  
 UFGF NGFF

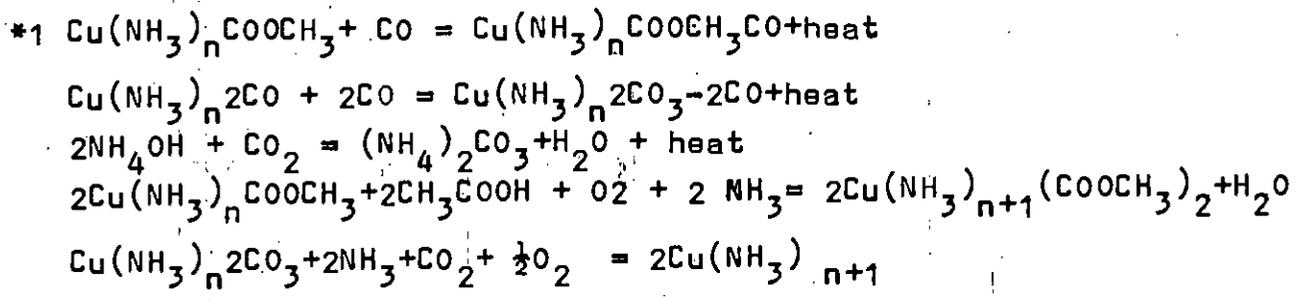


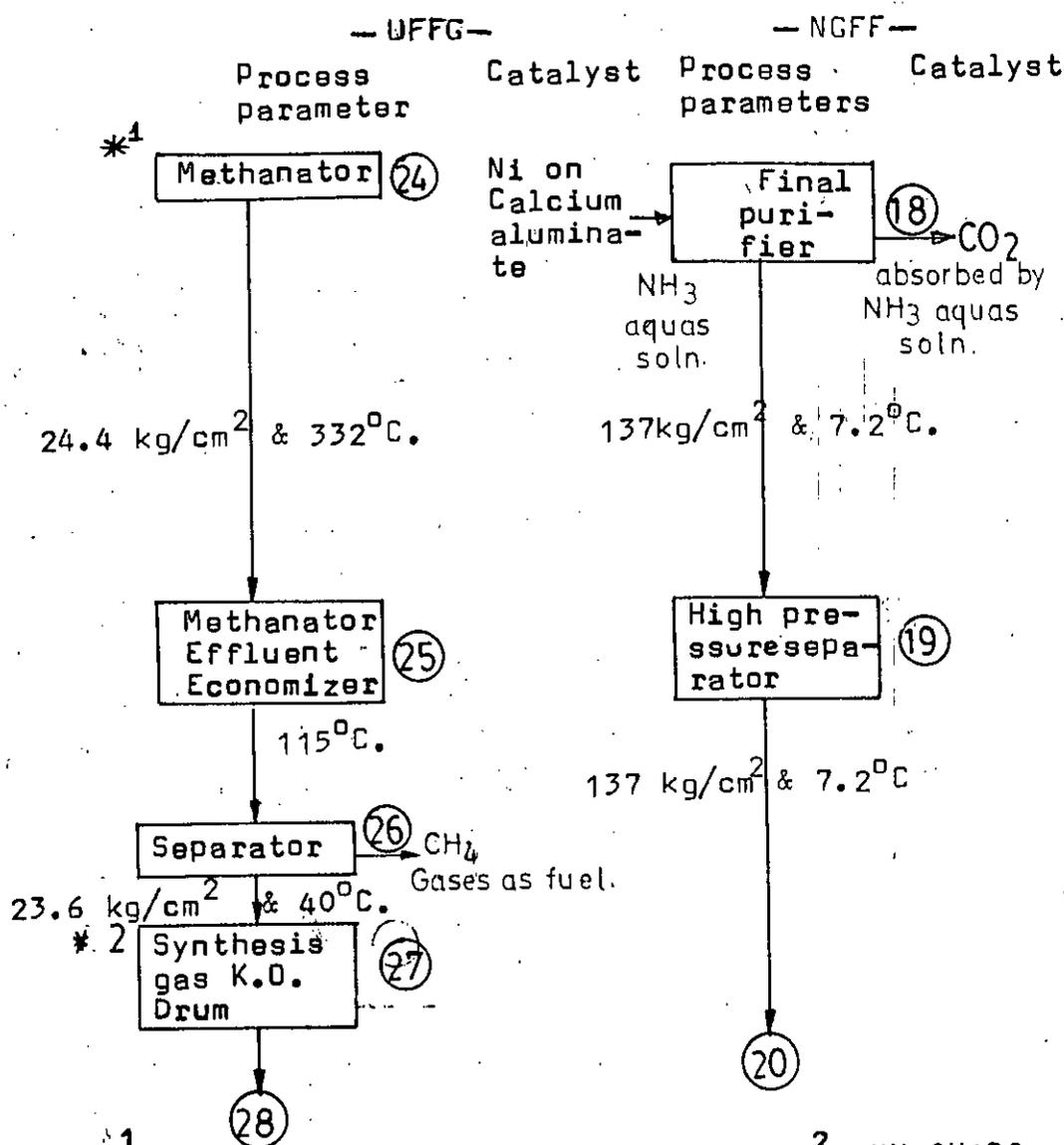
- a) Absorbing solution is GV solution which is prepared by  $K_2CO_3$ .
- a) Carbonate sol<sup>n</sup> is used.
- b) Purified gas contains 0.8%  $ASO_3$  and water concentrations of potassium & arsenic are 200 g/l & 140 g/l as  $K_2O$  and  $AS^{+3}$  respectively. Anhydrous arsenous acid accelerates absorption & desorption velocities.
- b) Purified gas contains 0.8%  $CO_2$  and 1.4%  $CO$ .
- c) 0.13%  $CO_2$  is present in purified gas.



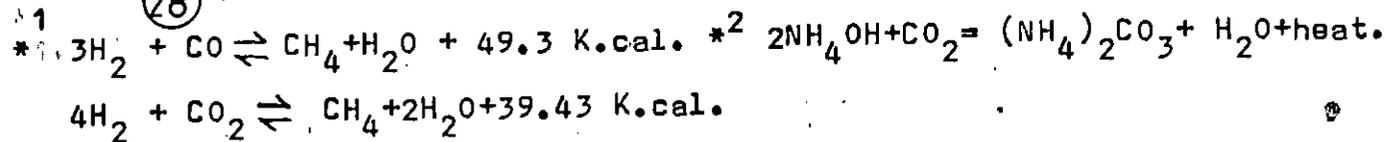


- a) To remove the 0.8% CO<sub>2</sub> and 1.4% CO, Cu-liquor scrubber is used which is packed with iron rasching rings and reduces CO to 5 to 10 ppm and CO<sub>2</sub> to 50 to 60 ppm.
- b) Here for Scrubbing Cu-Sol<sup>n</sup>.liquor is used.





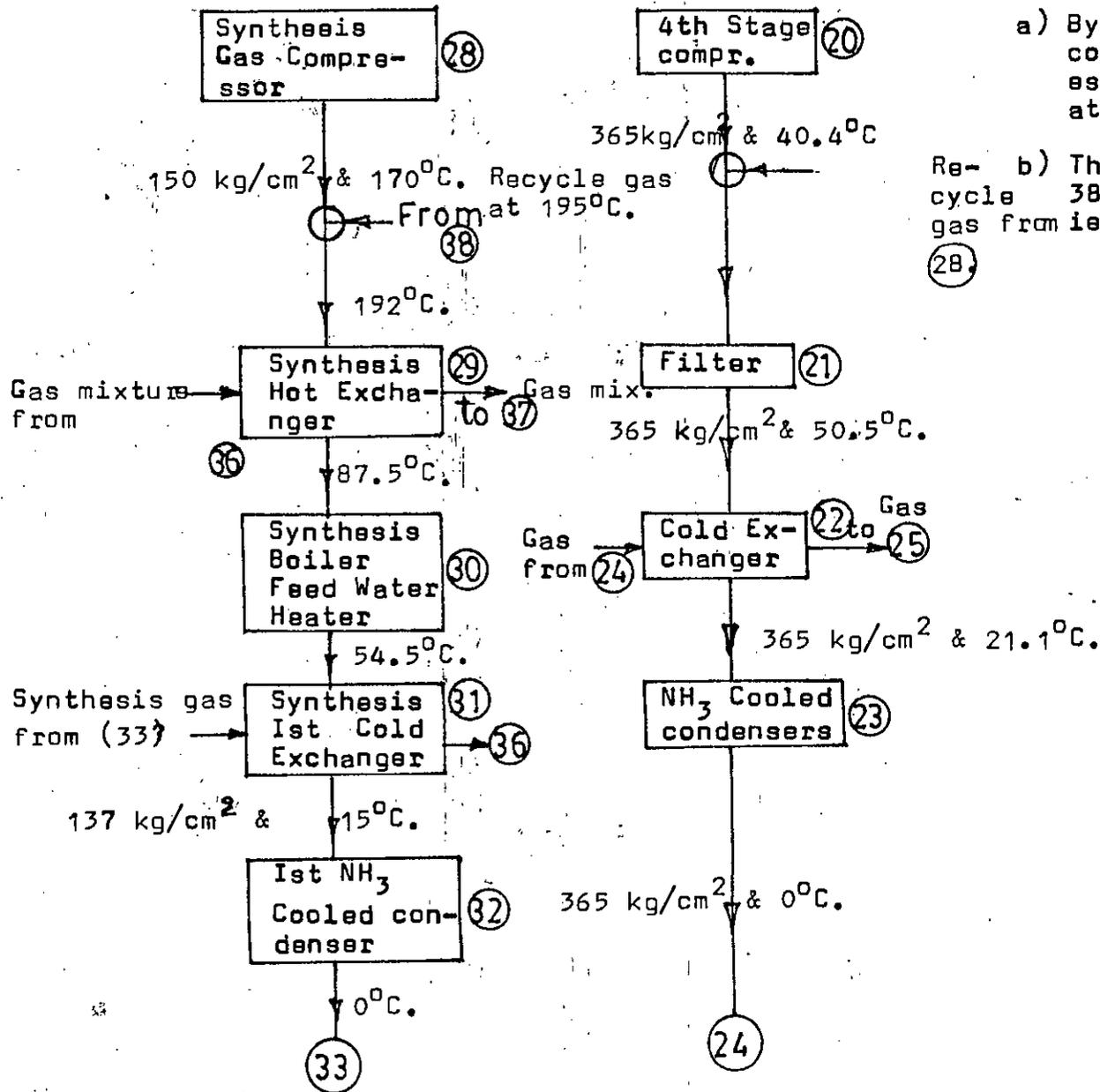
- b) For Final purification the gas enters into the final purifier where CO<sub>2</sub> content is reduced to 0 to 5 ppm by NH<sub>3</sub> aq. solution.



— UFFG —  
Process  
parameter

— NGFF —  
Catalyst Process  
Catalyst parameters

UFFG Diff. between  
NGFF



a) By synthesis gas compressor gas pressure is maintained at 139 kg/cm<sup>2</sup> & 170°C.

a) By 4th stage compressor gas pressure is maintained at 365 kg/cm<sup>2</sup> and 40.4°C.

Re-cycle gas is 38% and make up gases from is 68%. (28)

b) The recycle gas is 22% and make up gases is 78%.

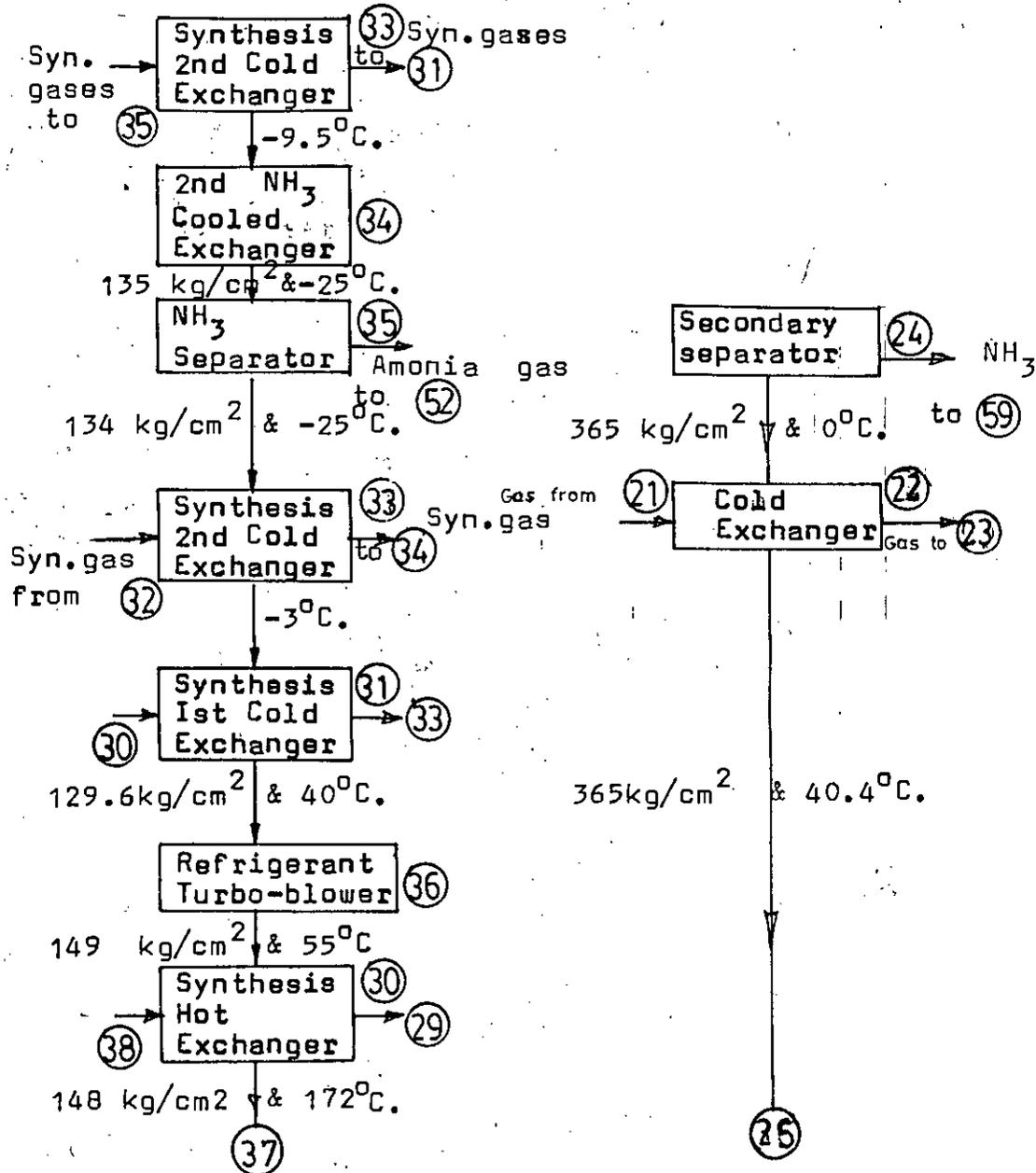
— UFFG —

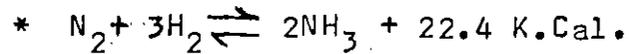
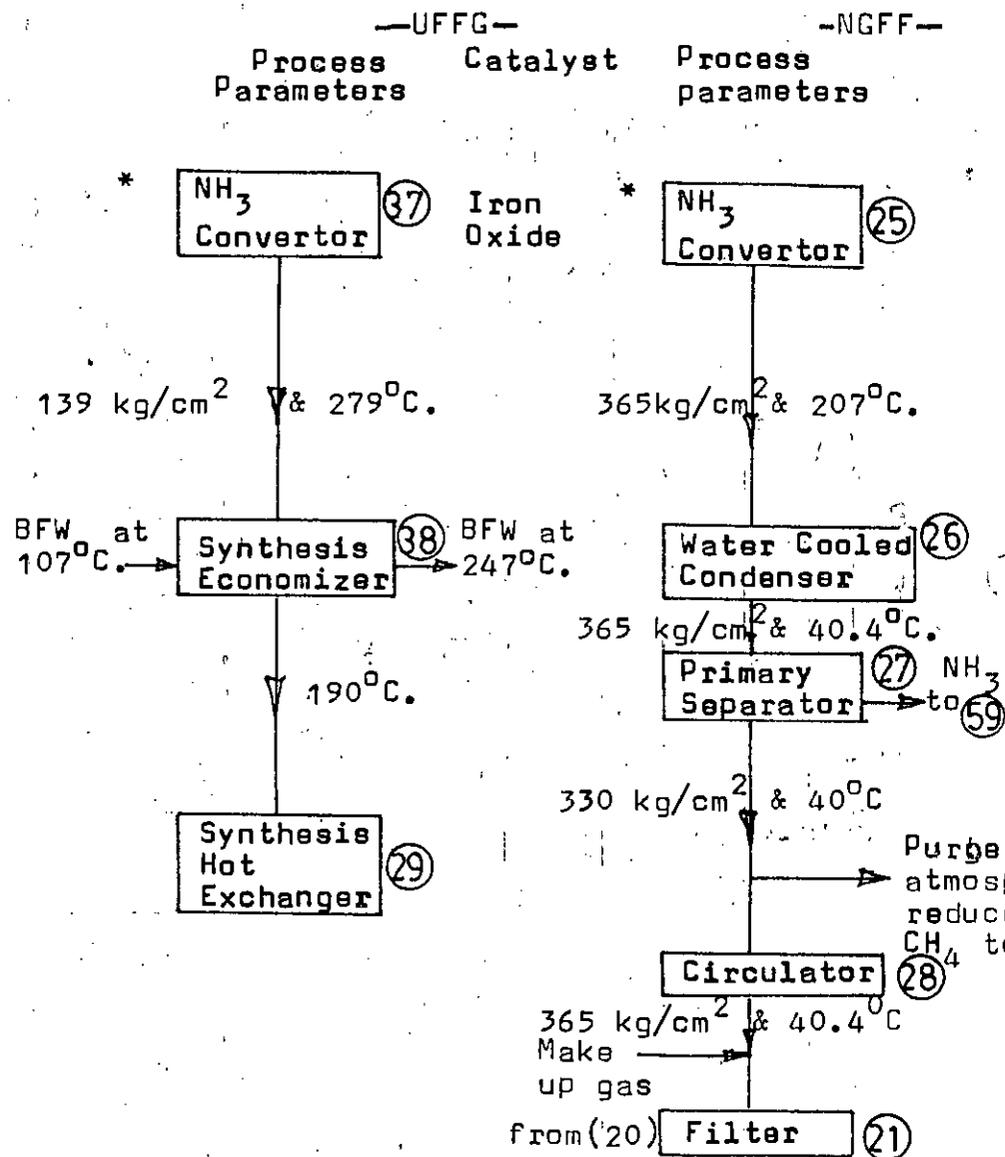
Process Parameters Catalyst

— NGFF —

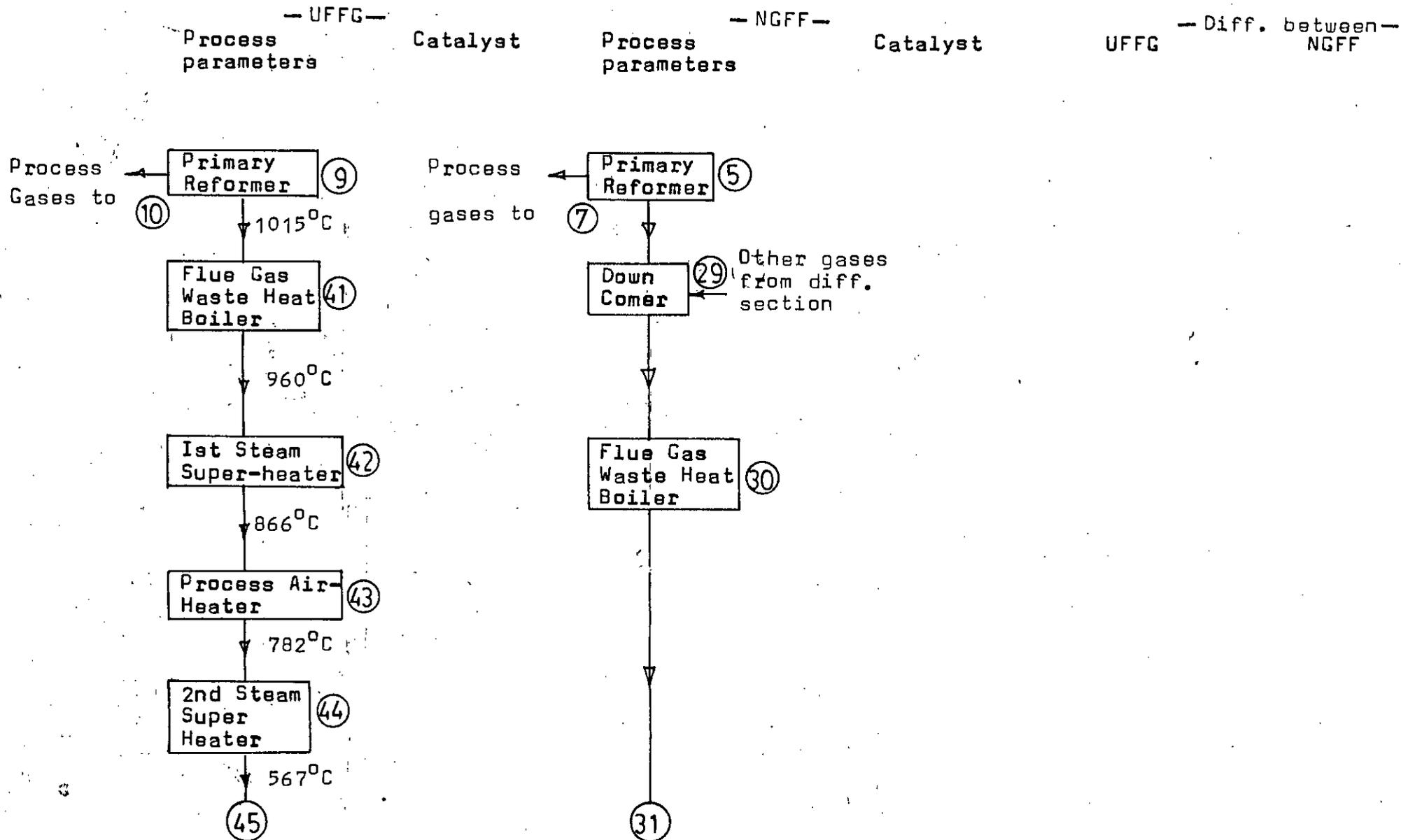
Process parameters Catalyst

— Diff. between —  
UFFG NGFF

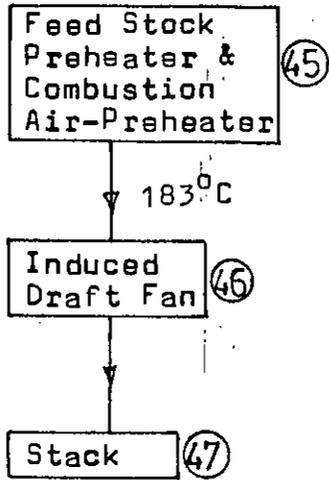




### FLUE GAS FROM PRIMARY REFORMER

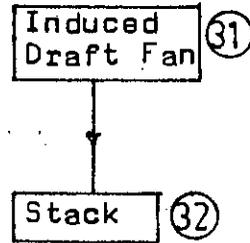


—UFFG—  
Process  
parameters



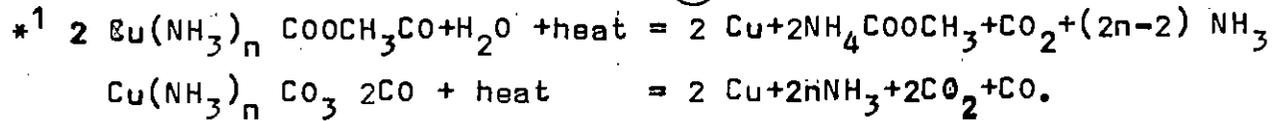
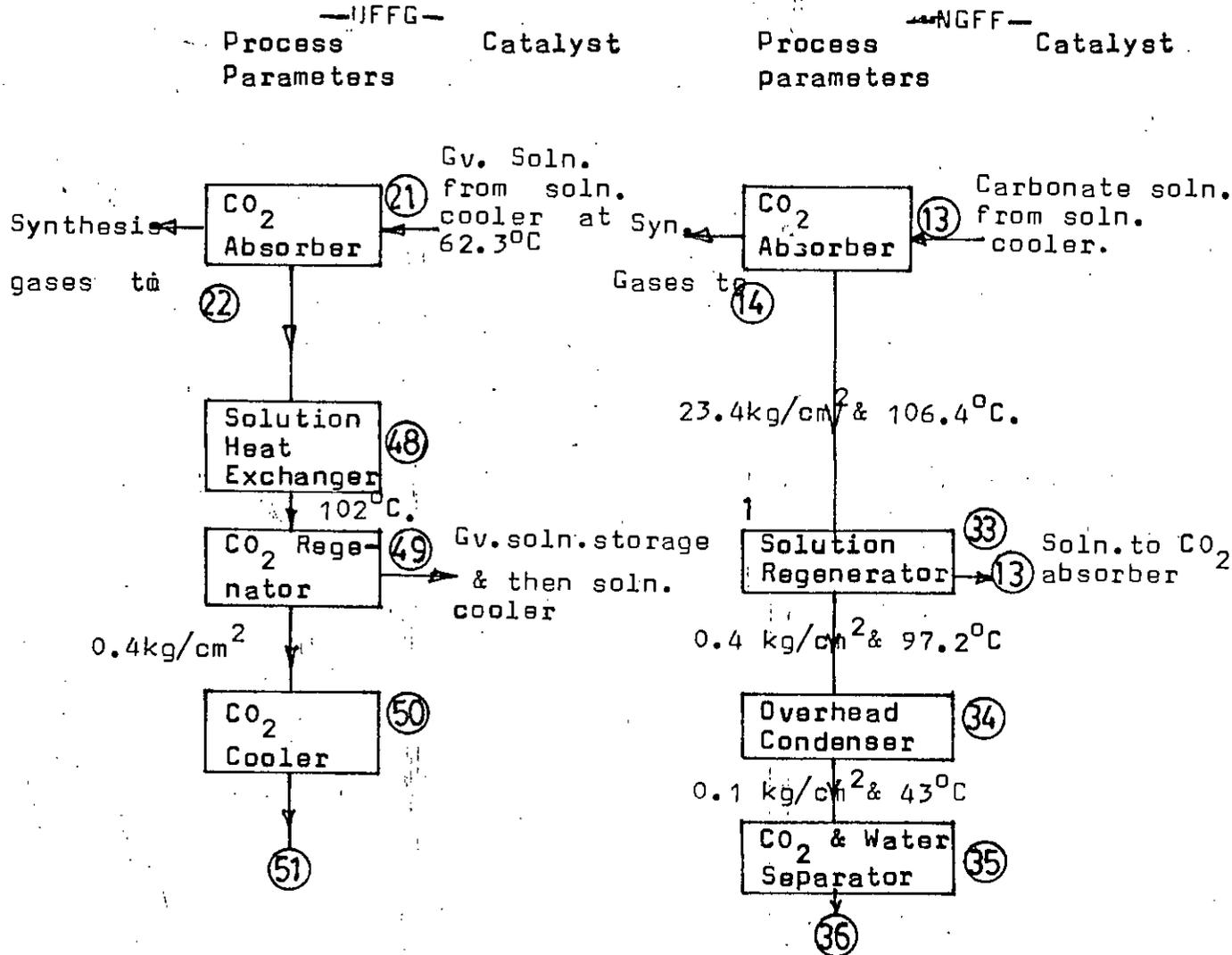
Catalyst

—NGFF—  
Process  
parameters



—Diff. between—  
UFFG NGFF

# PRODUCTION OF CO<sub>2</sub> GAS



— UFFG—  
 Process parameters Catalyst

CO<sub>2</sub> Gas (51)

— NGFF—  
 Process parameters Catalyst

— Diff. between—  
 UFFG NGFF

CO<sub>2</sub> Gas (36)

CO<sub>2</sub> Condensate separator (37)



From NH<sub>3</sub> scrubber

Ist Stage CO<sub>2</sub> Compressor & Ist Stage Discharge Cooler (38)

Separator (39)

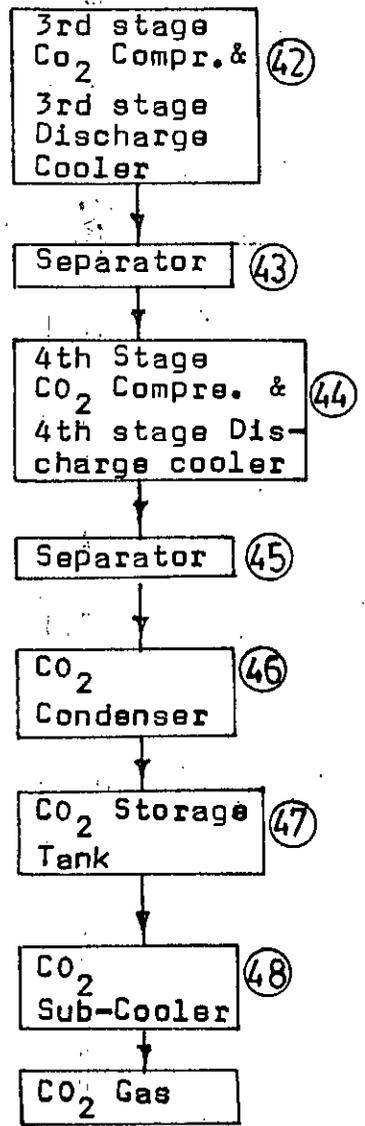
2nd Stage CO<sub>2</sub> Compr. and 2nd Stage Discharge Cooler (40)

Separator (41)

(42)

a) 100% of CO<sub>2</sub> is obtained from this section. a) Only 88.88% of CO<sub>2</sub> is obtained from this section.

- UFFG -		- NGFF -		- Diff. between -	
Process parameters	Catalyst	Process parameters	Catalyst	UFFG	NGFF



—UFFG—  
 Process parameters

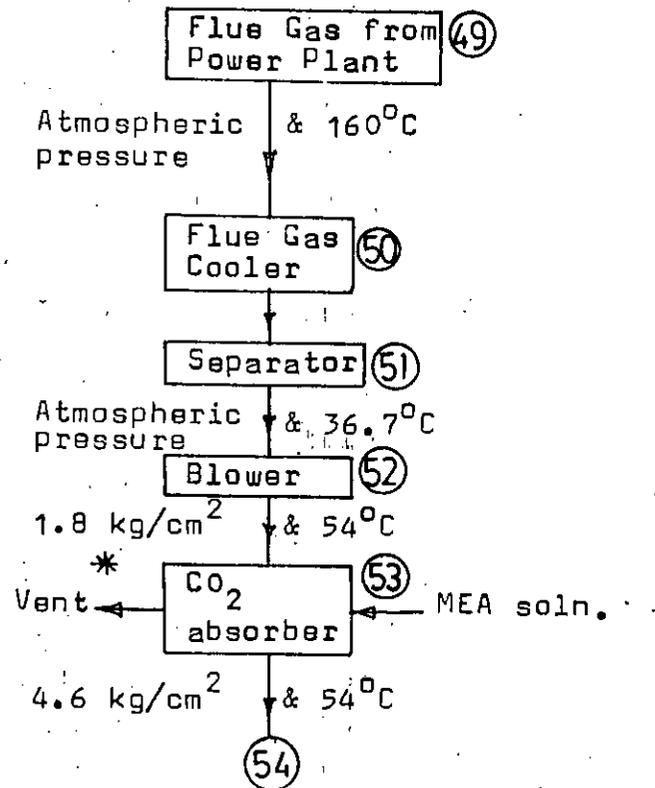
Catalyst

Process parameters

—NGFF—

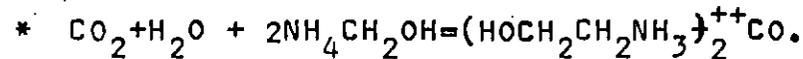
Catalyst

— Diff. between—  
 UFFG NGFF



a) CO<sub>2</sub> (10.6%) in the Flue Gas of the Power plant is recovered.

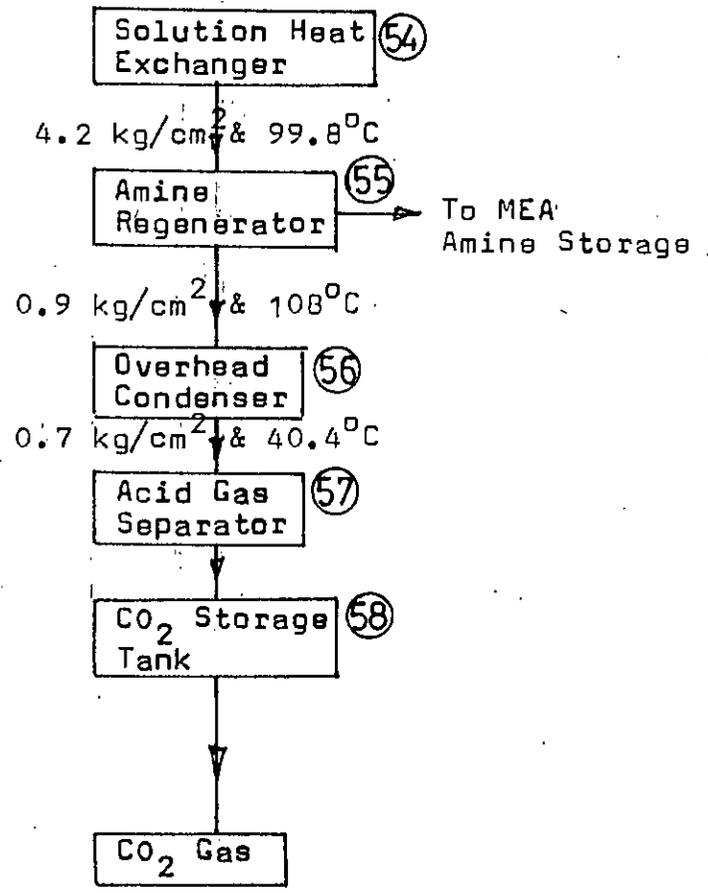
- a) Mono-ethanol-Amine is used to absorb CO<sub>2</sub> from the flue gases.
- b) O<sub>2</sub> content should be below 2.5% in the flue gas.
- c) Soda ash is used as Mea detergent and 10% CuSO<sub>4</sub> aqua is added as inhibitor of corrosion.



— UFFG —  
Process parameters Catalyst

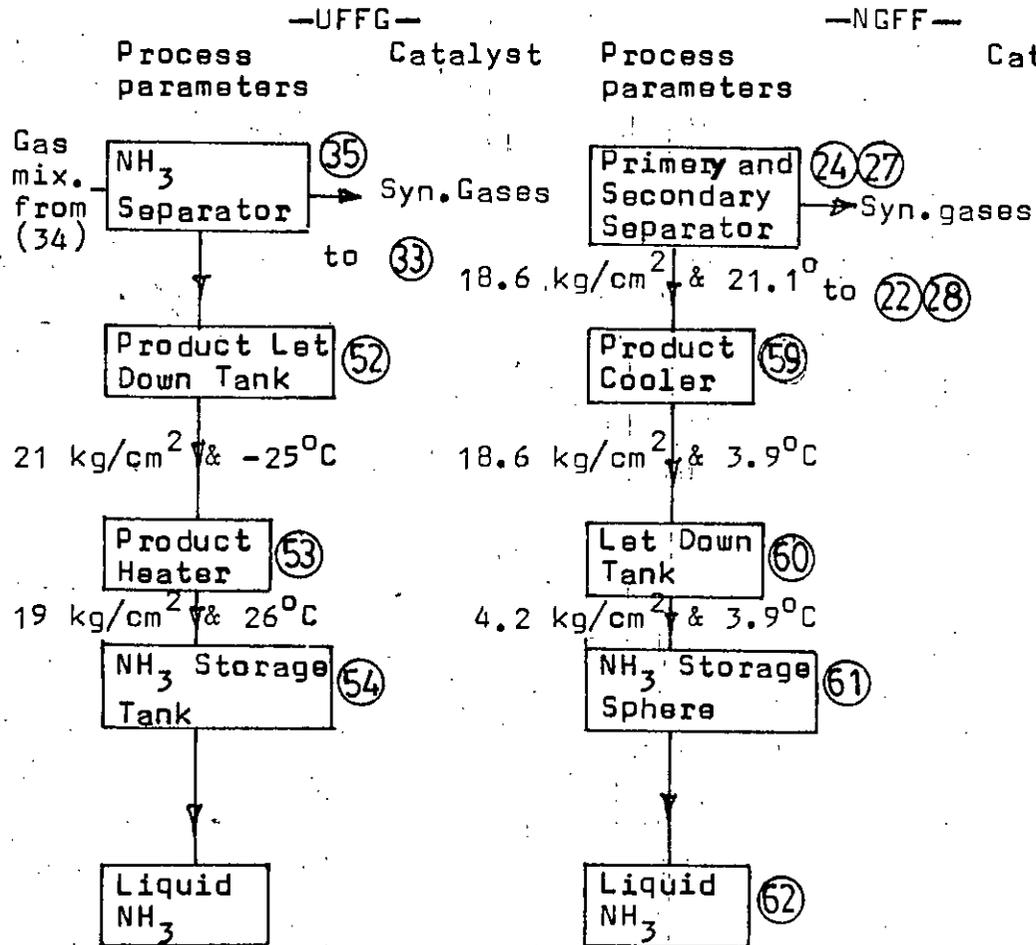
— NGFF —  
Process parameters Catalyst

— Diff. between —  
UFFG NGFF



a) 11.11% of CO<sub>2</sub> is obtained from this section for urea production.

# PRODUCTION OF LIQUID AMMONIA



- Diff. between -  
NGFF

a) Most of NH<sub>3</sub> is separated in NH<sub>3</sub> separator.

a) In Primary Separator 50-60% of NH<sub>3</sub> is separated and in Secondary Separator, 47-37% of rest NH<sub>3</sub> is separated.

a) 660 Tons of NH<sub>3</sub> is produced per day.

b) 204 Tons of NH<sub>3</sub> is produced per day.

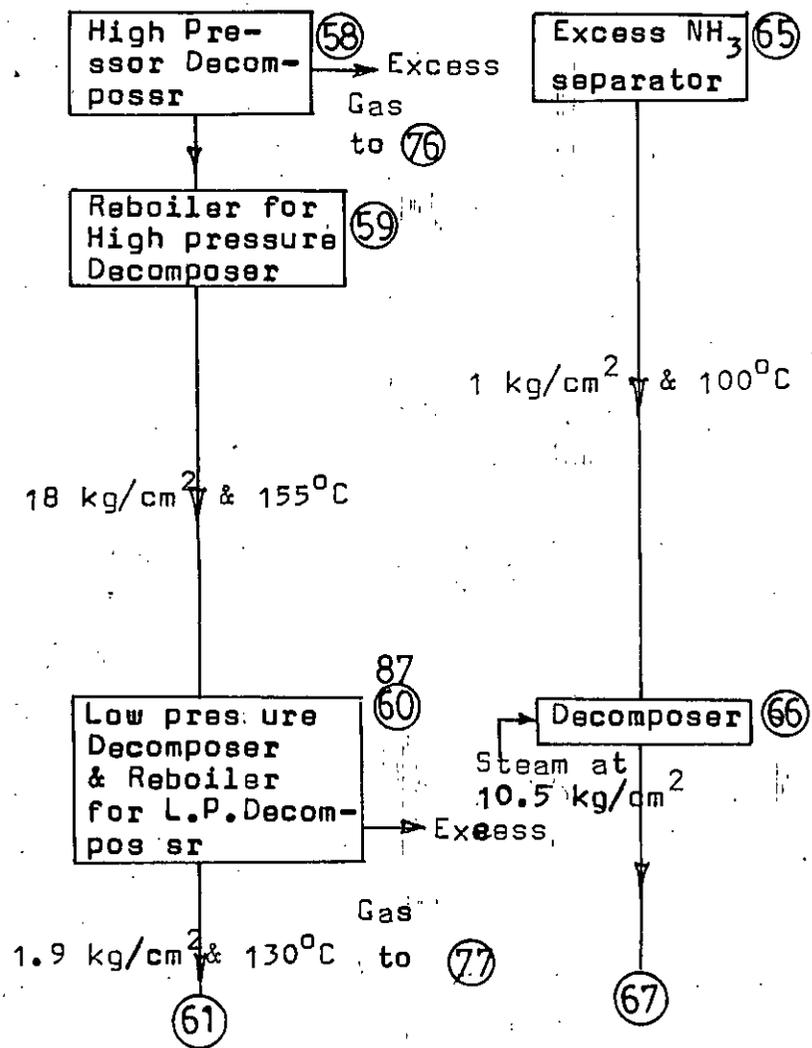


—UFGF—  
Process Catalyst  
parameters

—NGFF—  
Process Catalyst  
parameters

— Diff. between —  
UFGF NGFF

- c) 60% of urea reaction is completed.
- d) Mol. ratio:  
 $\text{NH}_3:\text{CO}_2 = 5.6:1.$



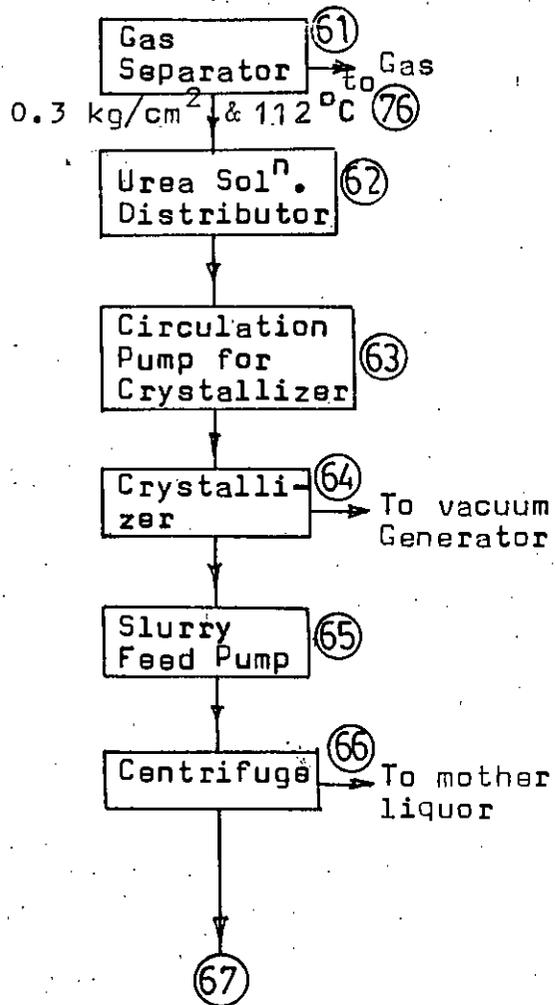
- a) Press. = 18 kg/cm<sup>2</sup>  
Liq. = 155°C  
Gas = 120°C
- b) Gas Gas Liq.  
Urea - 56.2%  
NH<sub>3</sub> - 68.4% 8.79%  
CO<sub>2</sub> - 25.2% 3.49%  
H<sub>2</sub>O - 6.4% 31.5%  
Biuret - 0.02%

a) Press. - 1.9 kg/cm <sup>2</sup>	a)	kg mol/-% by wt.
liq. - 130°C		hr.
Gas - 120°C	NH <sub>3</sub> - 69.4	- 12.1
b)	Gas	Liq.
Urea -	66.8%	Carbamate - 26.7
NH <sub>3</sub> 48.5%	1.1%	Urea - 80.1
CO <sub>2</sub> 19.3%	0.5%	H <sub>2</sub> O - 98.7
H <sub>2</sub> O 32.2%	31.2%	- 18.0
Biuret -	0.3%	

—UFFG—

Process parameters

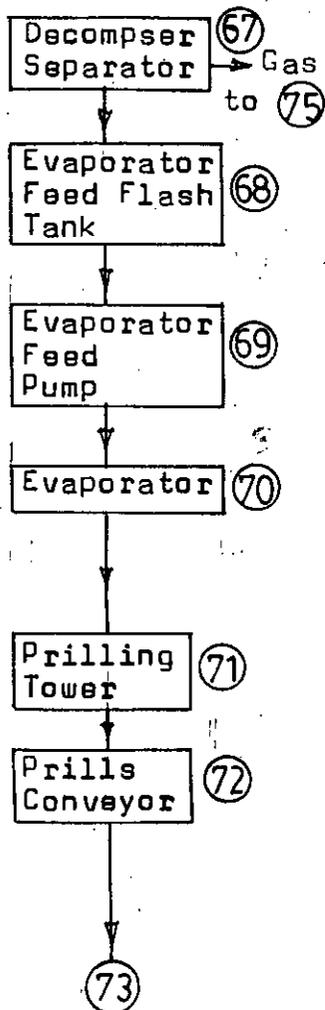
Catalyst



—NGFF—

Process parameters

Catalyst



— Diff. between —

UFFG

NGFF

- a) Urea Slurry having nearly 35% of crystal is sent to centrifuge.
- b) Crystal Urea from Centrifuge contains 1.0-1.5% of  $\text{H}_2\text{O}$  and 0.10 - 0.15% of biuret.

-UFFG-

Process parameters

Catalyst

Screws & Belt  
Conveyors (67)

Dryer (68)

Cyclone (69)

Screw  
Conveyor (70)

Melter (71)

Strainer, Head  
Tank Distributor (72)

Fluidizing  
Cooler (73)

Tommel (74)

Product  
Urea (75)

-NGFF-

Process parameters

Catalyst

Prills  
Cooler (73)

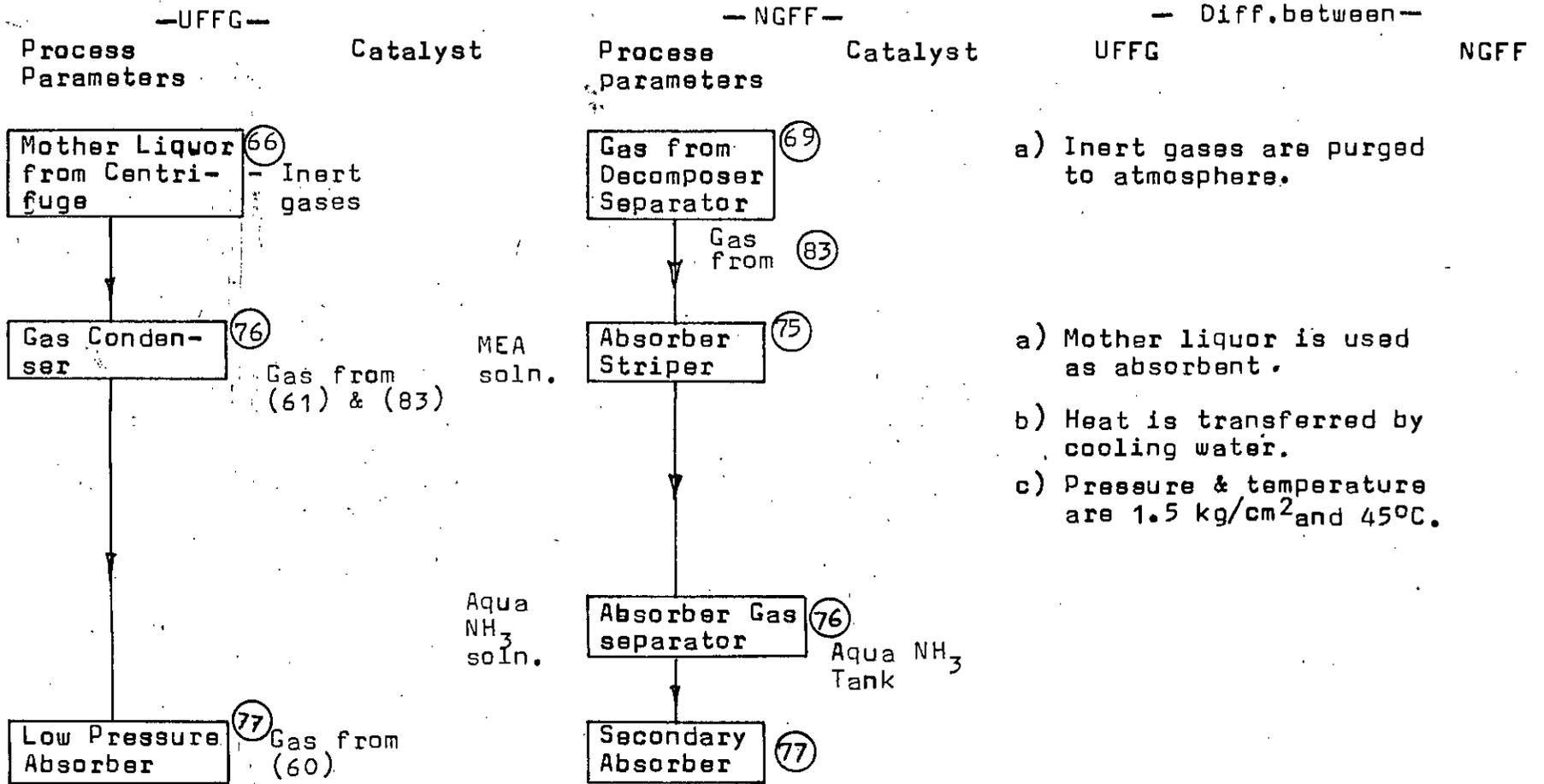
Prills of  
Urea (74)

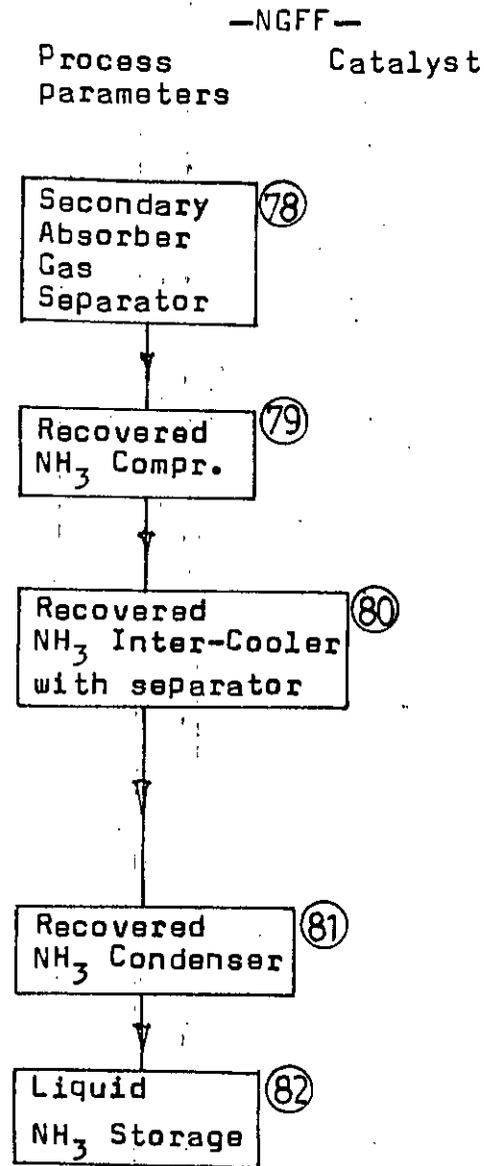
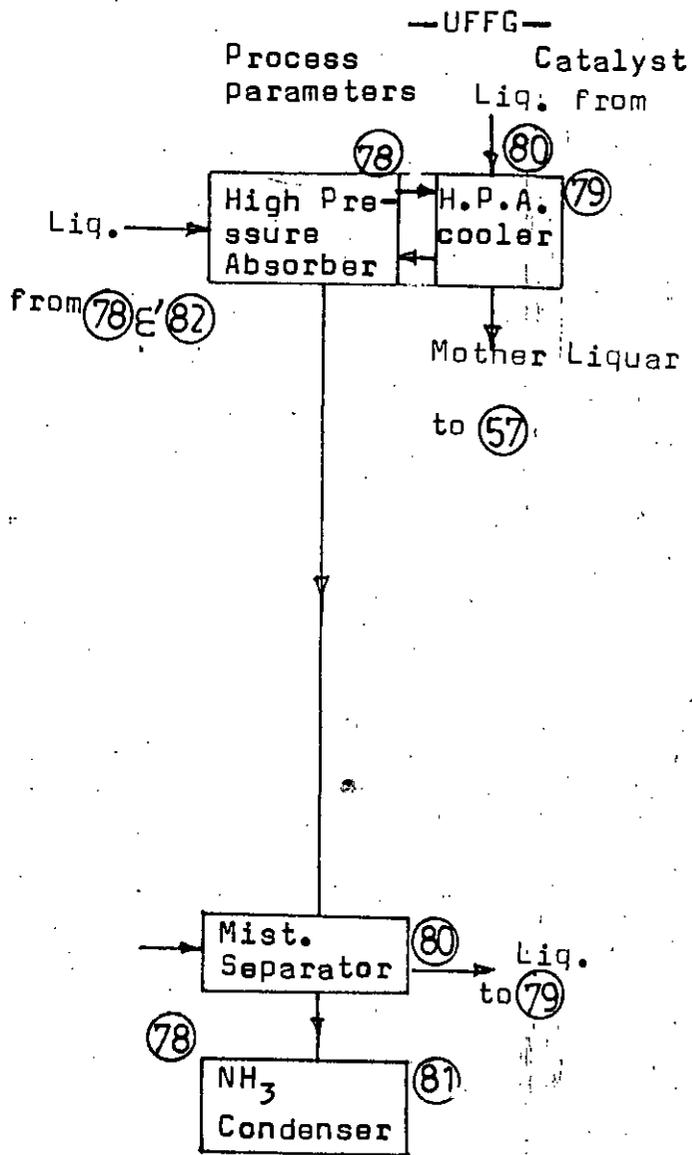
-Diff.between-

UFFG

NGFF

# RECOVERY SECTION





-- Diff. between --

NFFG      NGFF

- a) Heat is transferred by cooling water.
- b) Pressure and temp. of liquid are 17.5 kg/cm<sup>2</sup> & 50°C.
- c) The CO<sub>2</sub> concentration of mother liquor in H.P. Absorber is 30%.
- d) Pressure & temperature of liquid is 17.5 kg/cm<sup>2</sup> & 100°C.
- e) Concentrated carbamate Soln. contain
 

Urea	-	8.3%
NH <sub>3</sub>	-	36.0%
CO <sub>2</sub>	-	30%
H <sub>2</sub> O	-	25.4%
Biuret	-	0.3%

-UFFG-

Process  
parameters

Catalyst

-NGFF-

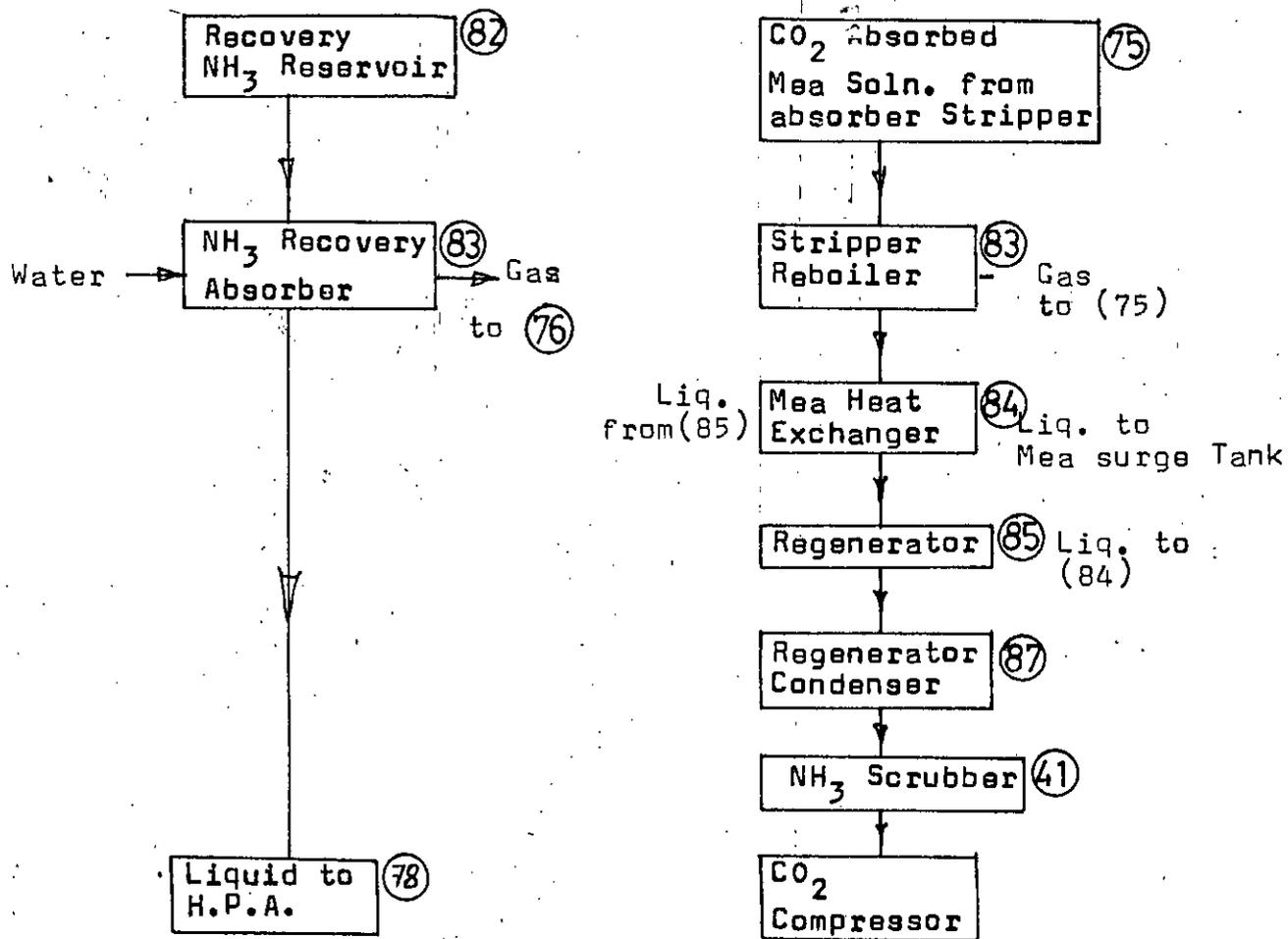
Process  
parameters

Catalyst

- Diff.between-

UFFG

NGFF



#### 4.3 Findings and Comments

1. For removal of sulphur from NG, Chemical process is used in UFFG and mechanical process is used in NGFF. In Chemical process, ZnO is used as catalyst in DSV whereas activated charcoal is used in mechanical process in DSV.
2. In UFFG, reformation takes place at higher pressure and temperature than that of NGFF. The values are respectively  $30.3 \text{ kg/cm}^2$ ,  $817.8^\circ\text{C}$  and  $6.5 \text{ kg/cm}^2$ ,  $620^\circ\text{C}$ .
3. For reformation, steam/carbon ratio is maintained at 3.5:1 in UFFG. It is 1:1 in NGFF. The higher steam/carbon ratio will give minimum soot formation resulting in high  $\text{CO}_2/\text{CO}$  conversion.
4. Ni on calcium alluminate used as reforming catalyst in UFFG and Ni on refractory of  $\text{MgO}$ ,  $\text{Al}_2\text{O}_3$ ,  $\text{Ca}(\text{AlO}_2)_2$  and  $\text{Ca}(\text{SiO}_3)$  used as catalyst in the other.
5.  $\text{CH}_4$  content at the end of reformation is 0.2% to 0.6% in UFFG and 0.3% in NGFF.
6. After reformation, for conversion of rest of CO into  $\text{CO}_2$ , Zinc-copper used as catalyst in UFFG and conversion rate is about 97%. Thus total content of CO at the outlet of shift convertor of UFFG varies 0.35 to 0.50%.  
In NGFF, where  $\text{Fe}_3\text{O}_4$  is used as a catalyst conversion rate is 60% and the total content of  $\text{CO}_2$  in the outlet of CO-convertor is 2%.

7. To remove  $\text{CO}_2$  in the UFFG plant, BV solution is used as absorbing solution. The purified gas contains about 0.13%  $\text{CO}_2$ . But in NGFF carbonate solution is used as the absorbing solution. The purified gas contains 0.8% and 1.4%  $\text{CO}_2$  and CO respectively. GV solution is prepared by  $\text{K}_2\text{CO}_3$ ,  $\text{AsO}_3$  and water.

8. For final purification of gases in UFFG, the Methanator is used where  $\text{CO}_2$  and traces of CO are converted into  $\text{CH}_4$  and used as fuel in another section.

Cu-Liquor Scrubber and Final Purifier is used in NGFF for purification of gases. After purification, the gas contains  $\text{CO}_2$  and CO from 0 to 5ppm and from 5 to 10 ppm respectively.

9. In UFFG, total  $\text{CO}_2$  is obtained from reforming gases in  $\text{CO}_2$  Regenerator but in NGFF, 88.88% of total  $\text{CO}_2$  is obtained from the same process and the rest are obtained from flue gases of utility section. For  $\text{CO}_2$  removal from FG Mono-ethanol-Amine is used as absorbing solution.

10. In UFFG, synthesis of  $\text{N}_2$  &  $\text{H}_2$  gases takes place at  $150 \text{ kg/cm}^2$  and converted into  $\text{NH}_3$  and rate of conversion is about 7.2%. In NGFF, synthesis takes place at  $300\text{--}365 \text{ kg/cm}^2$  and the conversion rate is higher about 12.8%.

11. In UFFG, liquid  $\text{NH}_3$  and  $\text{CO}_2$  gas at a ratio 4:1, added with a carbamate solution in Urea Synthesis Reactor where 70% of urea reaction is completed.

Liquid  $NH_3$  is added with liquid  $CO_2$  at a ratio 5.6:1 in Synthesis Autoclave of NGFF where 60% of urea reaction is completed. But in this process a small amount of air is supplied for protecting corrosion.

## CHAPTER 3 5

DOWN-TIME ANALYSIS OF AMMONIA PLANT IN UGFF & UFFG

### 5.1 Background of Down-Time Analysis

The production of a urea fertilizer factory depends on the available production time and the average capacity utilisation during the operation. Non productive time consists of two factors. The one management(technology) independent and the other depends primarily on the management(technology). The former factors are, for example, failure of external power and NG supply, political unrest, periodic overhaul. Thus these factors cannot be considered while comparing the technology assimilation in different factories and will be called in this thesis, non-productive time. The second factor, which can be controlled through better assimilation of technology, should be reduced to a minimum. The example of the second factor are interruption in normal production due to failure of the units). The loss in production time due to this factor will be termed here as down-time. Hence the analysis of the down-time identifies the weak links of the fertilizer factory, so that in the future selection, design and operation of such a factory necessary precautions can be taken.

However, it should be kept in mind that one hour of down-time of NGFF and UFFG is not the same. One hour of down-time of NGFF costs 5.6 tons of urea which that of UFFG 45.3 tons of urea( Annexure-11).

Down-time of various machineries & equipments is also dependent on their age. Hence to equalize the age factor, down-time of both factories were analysed at the same age level, namely 6 years after the starting of Commercial Production.

Down-time of NGFF and UFFG will now be analysed in sections 5.2 and 5.3 respectively.

#### 5.2 Down-Time analysis of Ammonia-Plant in NGFF

It was described earlier that the ammonia plant is the major component in the urea production and that the production ammonia is based on Haber-Bosch process of ammonia-synthesis.<sup>5</sup> The NGFF has its own powerplant to secure uninterrupted power supply. The factory has two separate lines of production (A & B) and each flow line is equipped with separate machineries and equipments has equal capacity.

The analysis of down-time of the ammonia plant includes that of the components of ammonia plant as well as that of other plants, machineries or equipments which will stop the operation of Ammonia-plant. Production(hour) losses in Ammonia plant also includes the down-time of this plant as well as the schedule shutdown of the same. The total production (hour) losses of Ammonia-Plant due to overhauling, power-plant and other cause is listed in Table 5.1.

Table 5.1 : Production loss of Ammonia-Plant of NGFF during July, 1968-June, 1971

Month	Production loss due to																		
	Failure of m/cs and equipments of NH <sub>3</sub> and other plants (hour) (A)				Power Generating unit (Local) (hour) (B)				Over-hauling (hour) (C)		Strike (hour) (D)		Shortage of lubri-cating oil (hour) (E)		Total production loss (hour) (F = A+B+C+D+E)				
	1968	1969	1970	1971	1968	1969	1970	1971	1968	1969	1970	1971	1970	1971	1968	1969	1970	1971	
July	14.7	165.8	63.7	304.0	-	68.1	50.2	487	-	-	-	-	-	-	-	-	-	-	-
Aug.	180.5	1.9	13.8	384.2	8.3	15.9	-	-	-	-	-	-	-	-	14.7	233.9	114.0	791.0	
Sep.	128.2	114.5	173.5	231.4	1.2	-	-	138.4	-	14.9	-	-	-	-	188.9	17.8	13.8	384.2	
Oct.	46.2	-	17.0	233.7	31.8	-	-	-	696	1440	960	-	-	-	129.9	126.5	173.6	370.3	
Nov.	130.2	335.4	-	168.9	-	-	-	-	1200	547.1	1440	-	-	443.5	774	1440	977	677.2	
Dec.	149.1	-	-	4.3	-	-	-	-	82.8	-	1664	1265.6	-	-	87	1330.2	882.5	1440	255.9
Jan.	203.4	262.9	-	24.5	6	23.3	-	34.5	-	-	-	-	354	-	231.8	-	1464	1269.8	
Feb.	19.1	53.1	7.6	380.9	-	179.2	-	-	-	-	-	-	-	-	209.4	286.2	-	413.0	
Mar.	109.1	-	49.3	490.3	279.3	31.8	59.3	13.9	-	-	-	-	-	-	19.1	232.2	7.6	380.9	
Apr.	-	87.3	12.6	-	-	15.3	17.1	-	-	-	-	-	-	-	388.3	31.8	108.7	504.24	
May	45.6	52.7	17.4	-	1.1	-	-	-	-	-	-	-	-	-	-	102.6	29.7	-	
June	34.3	9.4	120.8	-	-	-	11.1	-	-	-	-	-	-	-	46.7	52.7	17.4	-	
															34.3	9.4	131.8	-	
Total	1060.4	1083	475.7	2222.3	327.7	333.6	137.8	1978.8	1999	3888	1612.6	-	530.5	3366.8	4501.6	5046.6			
								674.4							3415.7				

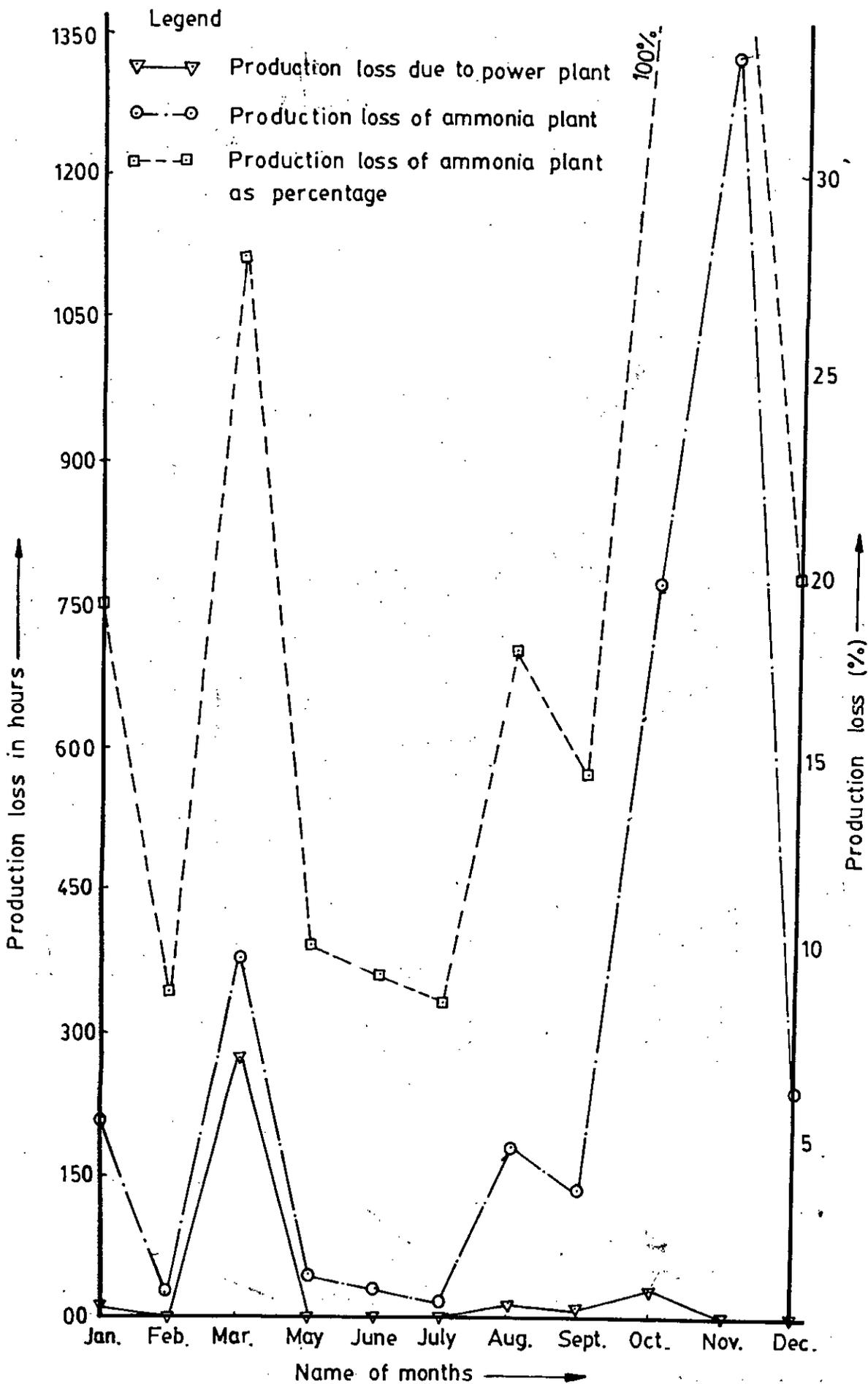


Fig 5.1 Production loss of ammonia plant and power plant in NGFF in 1988.

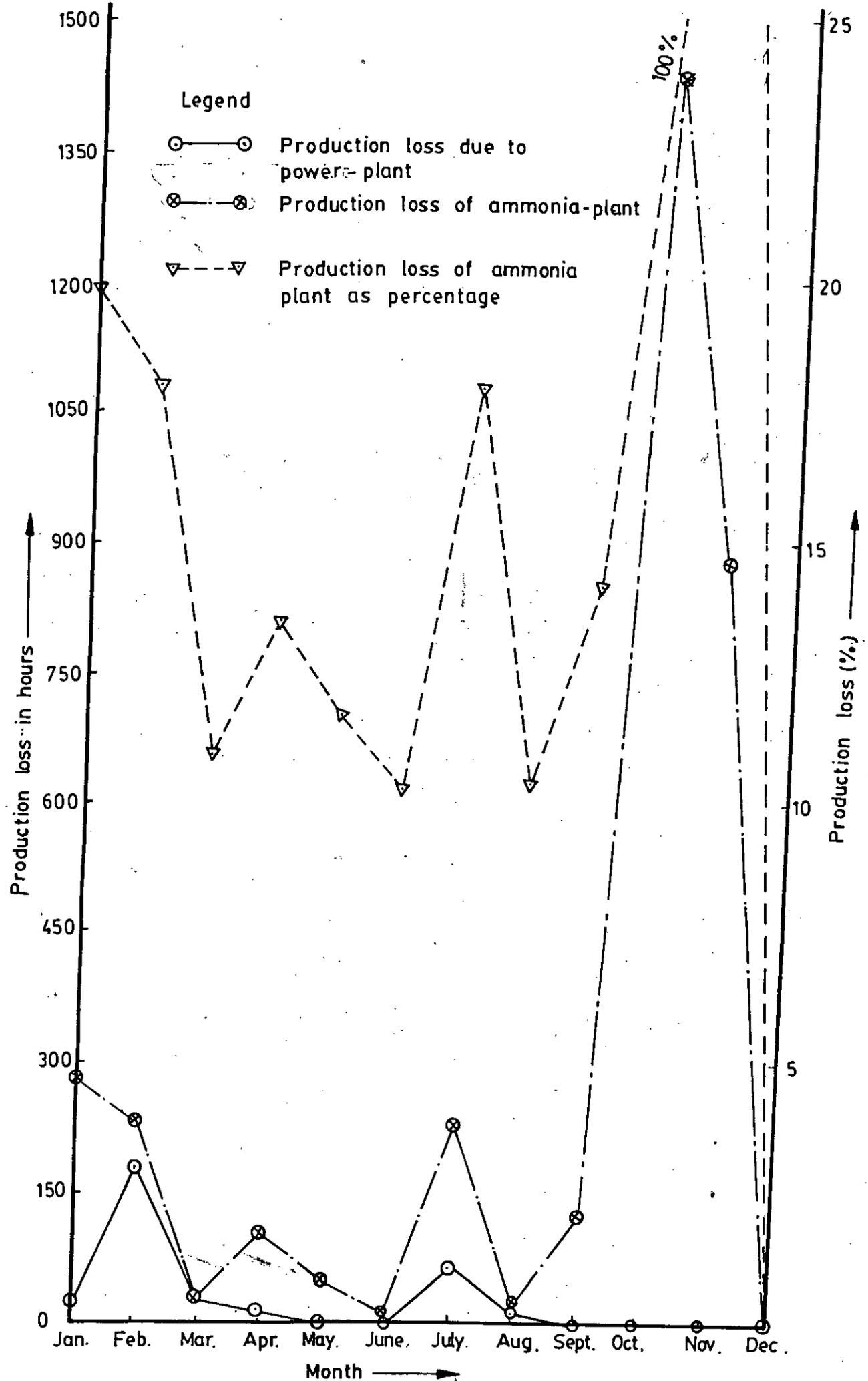


Fig.5.2 Production loss of Ammonia plant & power plant of NGFF in 1969.

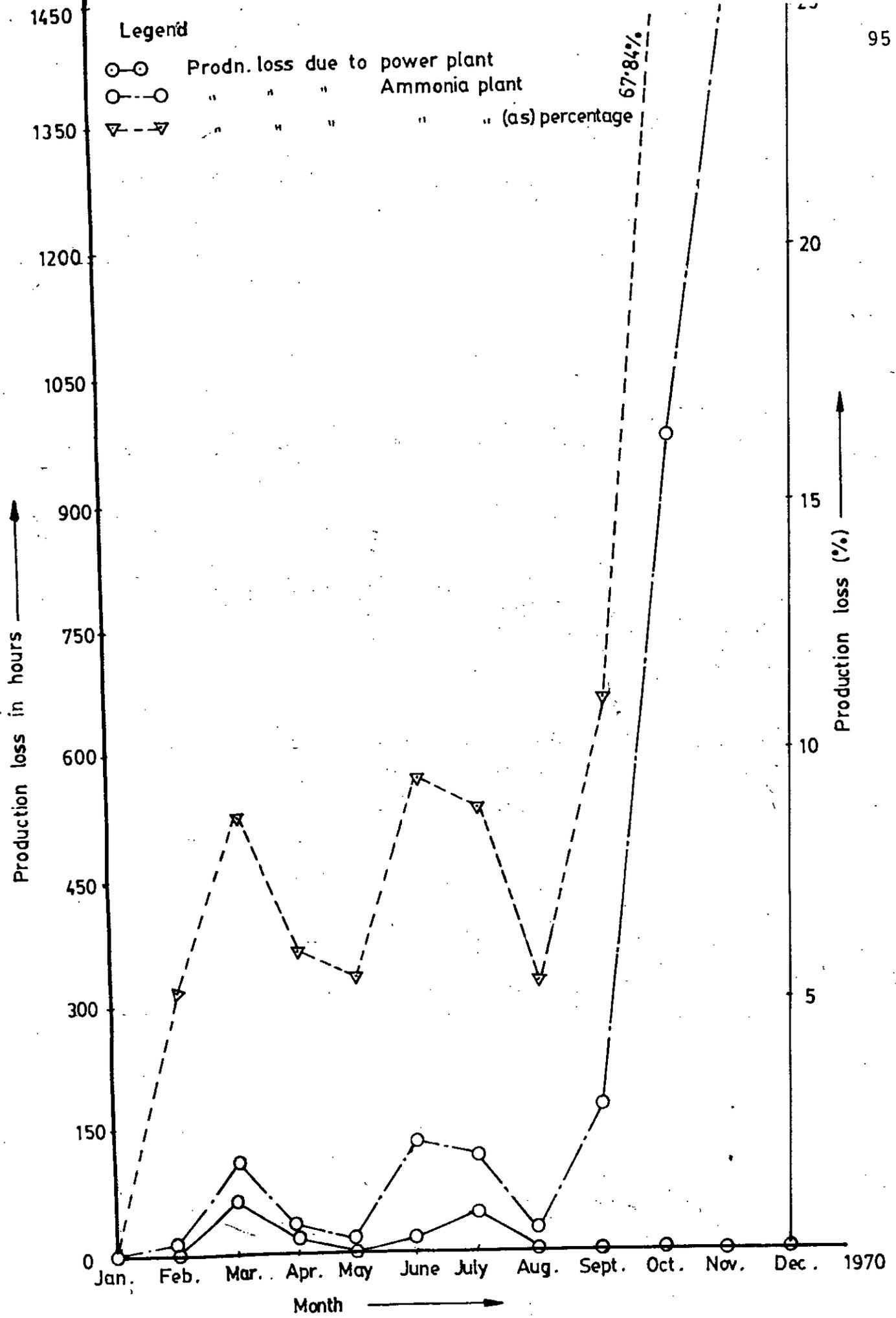


Fig. 5.3. Production loss of ammonia plant & power plant in NGFF in 1970

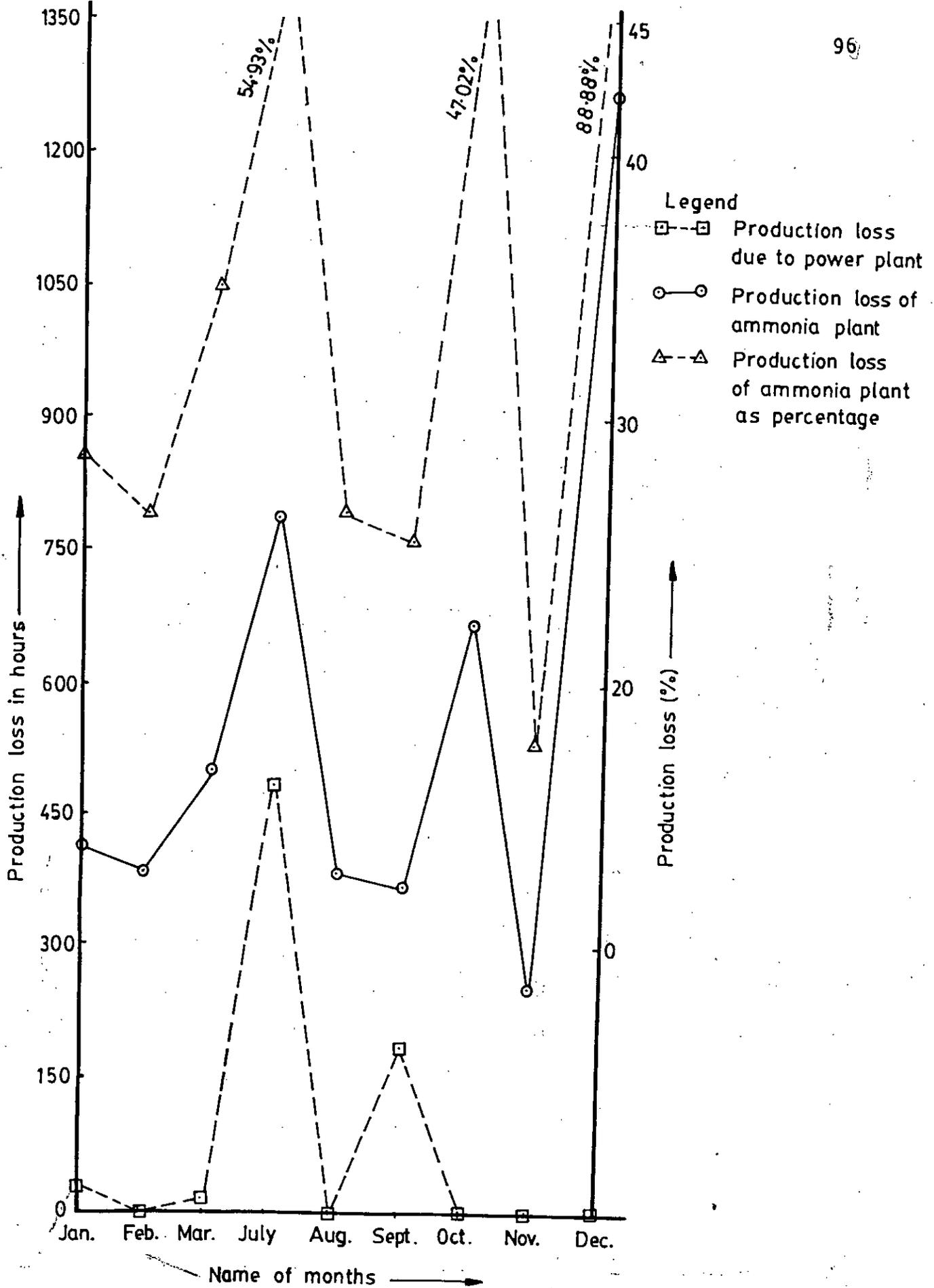


Fig5.4 Production loss of ammonia plant and power plant of in NGFF in 1971.

### 5.2.1 Down-Time Analysis(Unit-Wise)

Down-time of ammonia-plant due to failure of its machineries and equipments is listed in table 5.2. The causes of failure of different items are also different. After careful analysis, it can be found that the failure of different items depend on common factors such as:

- 1) High temperature
- 2) Leakage
- 3) Excessive vibration
- 4) Wear
- 5) Blockage
- 6) Damage
- 7) Breakage

Table 5.2 : The influence of different causes for the failure of machineries and equipments in Ammonia-plant of NGFF during 1968-71.

Name of Machineries	Causes of failure							Total (hour)
	Leakage (hour)	High Temp. (hour)	Excessive Vibration (hour)	Wear (hour)	Breakage (hour)	Damage (hour)	Blockage (hour)	
1. Primary Reformer(IRF)	499.3	221.5	-	-	-	-	-	720.6
2. Secondary Reformer(2RF)	-	413.0	-	-	-	-	-	413.0
3. Secondary Air compressor	-	126.7	31.3	-	-	35.1	6.7	199.8
4. Heater-tube of cu-liquor section	-	-	-	-	-	-	273.8	273.8
5. Main Gas Compressor	34.8	279.9	225.7	37.8	378.8	177.0	13.2	1151.1
Ist stage	7	7	-	-	16.9	4.8	4.8	33.5
2nd stage	-	40.7	13.6	4.0	72.4	15.7	-	146.4
3rd stage	-	21.2	9.3	5.3	141.4	80.6	8.4	266.1
4th stage	34.8	211.0	202.8	32.5	148.1	75.9	-	705.1
6. Circulator	51.48	21.5	65.3	8.75	0.8	-	-	147.8
7. Primary Separator	147.00	-	-	-	-	-	-	147.0

Contd---)

(Continuation from prepage )

Name of Machinerie	Causes of failure							
	Leakage (hour)	High Temp. (hour)	Excessive Vibration (hour)	Wear (hour)	Breakage (hour)	Damage (Hour)	Blockage (hour)	Total (hour)
8. Bearing box of CO <sub>2</sub> removal section	-	2.3	-	-	-	-	-	2.3
9. Safety valve of Recovery section	2.1	-	-	-	-	-	-	-
10. Heater element of convertor	-	-	-	-	-	95.5	-	95.5
11. Level control valve of Primary separator	-	-	-	-	-	60.5	-	60.5
12. Level Gauge of Primary Separator	14.3	-	-	-	-	-	-	14.3
13. Level Gauge of Secondary Separatár	14.3	-	-	-	-	-	-	14.3
14. Final Purifier of Cu-Liquor	-	-	-	-	-	-	1.8	1.8
Total:	763.1	1064.9	322.3	50.5	379.6	368.0	295.3	3243.9

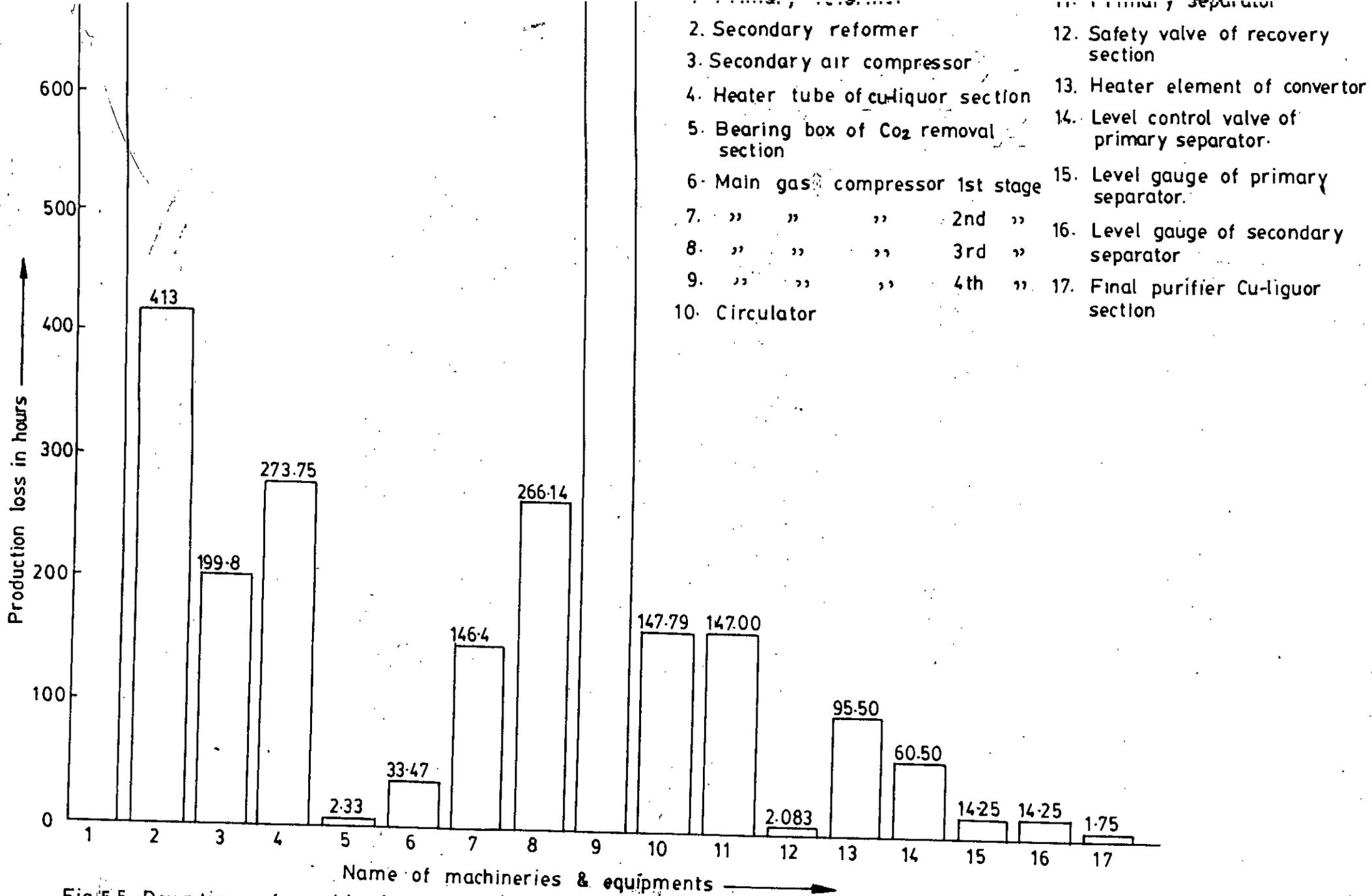


Fig 5.5. Downtime of machinevies and equipments of ammonia plant in NGFF.  
(Jan. 1968 to Dec 1971)

### 5.2.2 Down-Time Analysis(Factor-wise)

More detail analysis of the down-time is necessary to find out the different elements of the items, which cause the failure.

Tables 5.3, 5.4, 5.5, 5.6, 5.7, 5.9 show the down-time of various machineries and equipments and the factors which are responsible for the same. The data are represented in matrices.

Table 5.3 : The failure of different parts of different machineries and equipments of NGFF due to LEAKAGE during period of analysis.

Name of Parts	Name of Machineries and equipments (hours)										Total	
	Ist.Re- former	Main Gas Circulator		Primary Separator		Recovery Section		Primary Separator		Secondary Separator		
		Compr. (4th Sta- ge)	A	B	A	B	A	B	A	B		
1. Tubes	227.3	-	-	-	-	-	-	-	-	-	-	-
2. Flange of tubes	272.0	-	-	-	-	-	-	-	-	-	-	272.3
3. Gas from 4th stage joint	-	7.0	5.9	-	-	-	+	-	-	-	-	272.0
4. Gas from drain valve	-	7.1	-	-	-	-	-	-	-	-	-	12.9
5. Delivery Pressure gauge	-	-	4.1	-	-	-	-	-	-	-	-	7.1
6. Gas from bottom metallic point	-	-	10.8	16.8	-	-	-	-	-	-	-	4.1
7. Gas from Top cover holes	-	-	-	-	34.7	-	-	-	-	-	-	27.6
8. Level glass	-	-	-	-	-	3.0	144.0	-	-	-	-	34.7
9. Safety valve	-	-	-	-	-	-	-	2.1	-	-	-	147.0
10. Level Gauge	-	-	-	-	-	-	-	-	14.3	14.3	-	2.1
Total (Individual flow line)	499.3	14.1	20.8	16.8	34.7	3.0	144.0	2.1	14.3	14.3	-	28.6
Total:	499.3		34.9		51.5		147.0	2.1	14.3	14.3		763.4
												763.4

\*A, B indicate two separate flow lines of ammonia production.

Leakage in tubes and flanges of that tubes in 1RF causes the production loss at high percentage. Another important leakage occurs in Primary Separator Level glass of synthesis section. Leakage in level gauge of Primary and Secondary Separator are the causes which hampers in production.

Table 5.4 The failure of different parts of different machineries and equipments due to WEAR during the period of analysis.

Name of parts	Name of machineries and equipments					Total
	Main Gas Compressor			Circulator		
	2nd stage	3rd Stage	4th Stage			
	(A)	(B)	(A)	(B)		
Neck bush of cylinder	4.0	5.3	7.6	24.9	8.8	50.6
Total :	4.0	5.3	-	32.5	8.8	50.6

Maximum wear occurs in the neck bush of cylinder of 4th stage of Main Gas compressor and circulator.

Table 5.5 : The failure of different parts of different machineries and equipments due to BLOCKAGE

Name of Parts	Name of machineries and equipments				Total
	Secondary Air compressor	Ammonia cooled cu-solution cooler	Main Gas compressor 1st Stage	2nd Stage	
	B	B	A	B	
1. Water jacket & cooler	6.6	-	-	-	6.6
2. Heater tube	-	273.7	-	-	273.7
3. Strainer	-	-	4.7	4.7	9.5
4. Oil Filter	-	-	-	3.91	3.91
Total	6.6	273.7	4.7	8.6	293.7

Blockage of tubes of Ammonia-cooled copper solution Cooler causes the higher percentage of production loss, Blockage of strainer, oil filter of Main Gas compressor also hamper in production at less percentage.

Table 5.6 The failure of different parts of individual machineries equipments due to HIGH TEMPERATURE

Name of Parts	Name of machineries and equipments										Total
	1RF	2RF	Secondary Air-Compr.	Main Gas Compressor					Circulator	CO <sub>2</sub> Removal Section	
				1st Stage	2nd Stage	3rd Stage	4th Stage				
(A)	(A)	(A)	(B)	(A)	(B)	(A)	(A)	(B)	(A)		
1. Tubes	195.2	-	-	-	-	-	-	-	-	-	195.2
2. Catalyst	26.2	-	-	-	-	-	-	-	-	-	26.2
3. Brick Lining	-	413.0	-	-	-	-	-	-	-	-	413.0
4. Piston	-	-	8.5	-	-	-	-	-	-	-	8.5
5. Metallic	-	-	-	-	-	-	-	-	-	-	-
6a. Packing	-	-	24.7	-	14.6	20.2	8-0	77.9	93.6	21-5	260.6
6. Bearing	-	-	17.6	75.8	7.0	5.9	7.3	-	-	-	113.6
7. Cylinder	-	-	-	-	-	-	5-9	-	-	-	5.9
8. Piston ring-	-	-	-	-	-	-	-	39.4	-	-	39.4
9. Bearing box-	-	-	-	-	-	-	-	-	-	-	2.3
Total (individual) flow line	221.5	413.0	49.8	75.8	7.0	26.1	26.1	21.1	43.6	21.5	1078.8
								117.4			
Total	221.5	413.0	125.6	7.0	40.7	21.1	211.0	21.5	2.3	1078.8	

High temperature of tubes of 1st Reformer and of brick-lining of 2nd Reformer hampers in production at high percentage. Repairing or replacement of tubes and brick-lining requires long shut-down of the plant. High temperature of metallic packing of Secondary Air-Compressor, Main Gas Compressor and Circulator causes the production loss at high percentage. High temperature of bearing and piston ring are also the causes of production loss at less percentage.

Table 5.7 The failure of different parts of individual machineries & equipments due to EXCESSIVE VIBRATION

Name of Parts	Name of machineries and equipments										Total
	Secondary Air-Compressor			Main Gas Compressor Circulator							
			2nd stage	3rd stage		4th Stage					
	(A)	(B)		(A)	(B)	(A)	(B)	(A)	(B)		
Cylinder	11.8	-	13.5	5.4	3.9	120	36.6	52.8	12.5	256.6	
Bearing	-	19.5	-	-	-	-	-	-	-	19.5	
Liner	-	-	-	-	-	-	46.1	-	-	46.1	
Total	-	31.3	13.5	9.3	-	202.7	-	65.3	322.2		

Excess vibration of cylinder of Main Gas Compressor, Circulator and Secondary Air Compressor Causes production loss at higher percentage. Liner and bearing of the above compressors are also the parts for which production fails at less percentage.

Table 5.8 The failure of different parts of individual machineries+equipments due to BREAKAGE

Name of Parts	Name of machineries and equipments								Total
	1st stage		2nd stage		3rd stage		4th stage		
	(A)	(B)	(A)	(B)	(A)	(B)	(A)	(B)	
1. Strainer	5.6	4.9	-	-	-	-	-	-	10.5
2. Valve Plate of suction & delivery line		6.2	-	42.7	-	114.0	-	88.8	251.7
3. Spring of valve	-	-	4.7	-	14.7	12.6	2.0	-	34.0
4. Piston ring-		-	-	25.0	-	-	-	4.8	29.8
5. Liner of Cylinder	-	-	-	-	-	-	41.0	-	41.0
6. Separator drain valve	-	-	-	-	-	-	-	2.5	2.5
7. Filter drain valve	-	-	-	-	-	-	-	9.0	9.0
<b>Total</b>		<b>16</b>		<b>72.4</b>		<b>141.4</b>		<b>148.1</b>	<b>378.7</b>

Breakage of valve-plate of suction and delivery line of Main Gas Compressor causes production loss. Breakage of spring of valve, piston ring and liner of cylinder of main gas compressor are the causes of failure of production at less percentage.

Table 5.9 The failure of different parts of different machineries and equipments due to DAMAGE.

Name of parts	Name of machineries and equipments									
	Secondary Air Compr.		Main Gas Compressor					Conver- tor	Primary Separa- tor	Total
	(A)	(B)	(A)	(A)	(B)	(A)	(A)			
1. Instrument	0.8	35.1								
2. Valve Plate			4.8	10.6	-	80.6	75.9	-	-	35.8
3. Oil check valve					5.0					171.9
4. Heater element								83.0		5.0
5. Heater element transformer								83.0		83.0
6. Level Control valve								12.5		12.5
									60.5	60.5
<b>Total</b>	<b>35.9</b>		<b>4.8</b>	<b>15.6</b>		<b>80.6</b>	<b>75.9</b>	<b>95.5</b>	<b>60.5</b>	<b>368.8</b>

Damage of valve plate of Main Gas Compressor causes production loss at high percentage. Damage of Heater element of convertor and Level Control valve of synthesis section are also hamper in production at less percentage.

### 5.3 Down-time analysis of Ammonia-Plant in UFFG

It was described earlier that the ammonia-plant is the major component unit of urea production because  $\text{CO}_2$  and  $\text{NH}_3$  gas is produced in this plant. In UFFG, if the ammonia plant stops once a time, it requires minimum 48 hours to start production again. The UFFG has its own power-plant for continuous power-supply. The ammonia plant has an auxiliary boiler which supplies steam required in process and also in turbines of ammonia and other plants. Two types of instrumental process trip system is used ammonia-plant. This trip system is used as safety devices for process as well as machineries. One is A-Class trip which stops all machineries in this plant except the auxiliary boiler and another is AA-class trip which stops all with auxiliary boiler.

The analysis of down-time of ammonia-plant includes that of the components of ammonia plant as well as that of other plants, machineries or equipment which will stop the production of ammonia-plant. Production(hour) losses in ammonia-plant includes the down-time of this plant, overhauling, shortage of raw materials misoperation etc. The total production(hour)loss due to above causes is listed in table 5.10.

Table 5.10 Production loss of Ammonia plant of UFFG

Year	Down-time due to		Production loss due to			Total
	Failure of Machinery & equipments of Ammonia and other plants (hour)	Power plant and PDB (Hour)	Mis-operation (hour)	Over-hauling (hour)	Non-supply of NG (hour)	
1877	1542.48	-	128.4	1438	-	3469.60
1978	667.68	136.56	-	934.85	-	1739.09
1979	292.48	297.94	-	-	865.39	1455.81
1980	872.86	382.32	146.25	1595.06	-	2996.49
1981	-	-	-	-	81.48	81.48
Total	3375.50	816.82	274.65	3967.91	946.87	7859.54

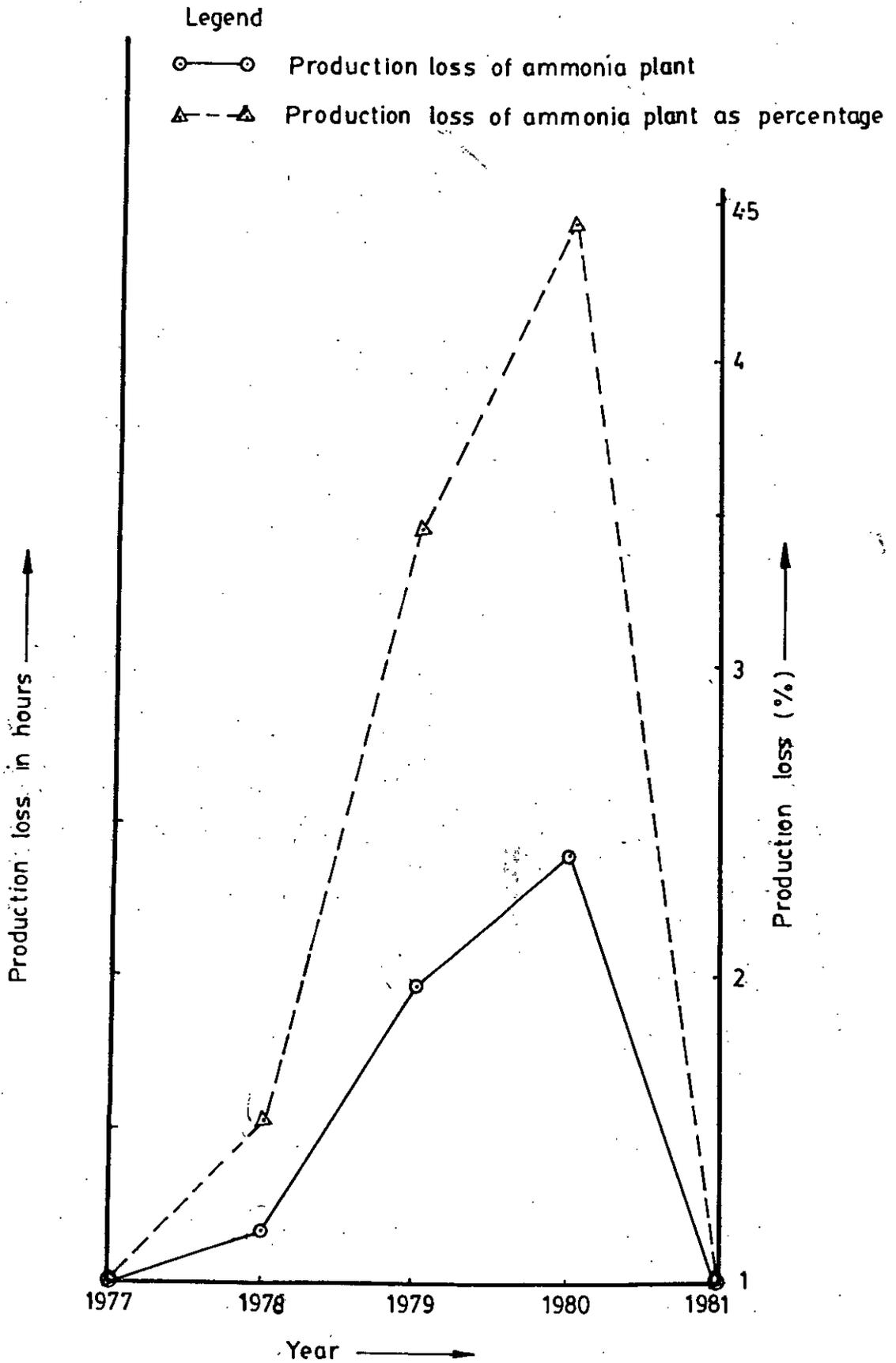


Fig5.6 Down-time of ammonia-plant due to PDB and local power-plant in UFFG. (June 1977 to June 1981)

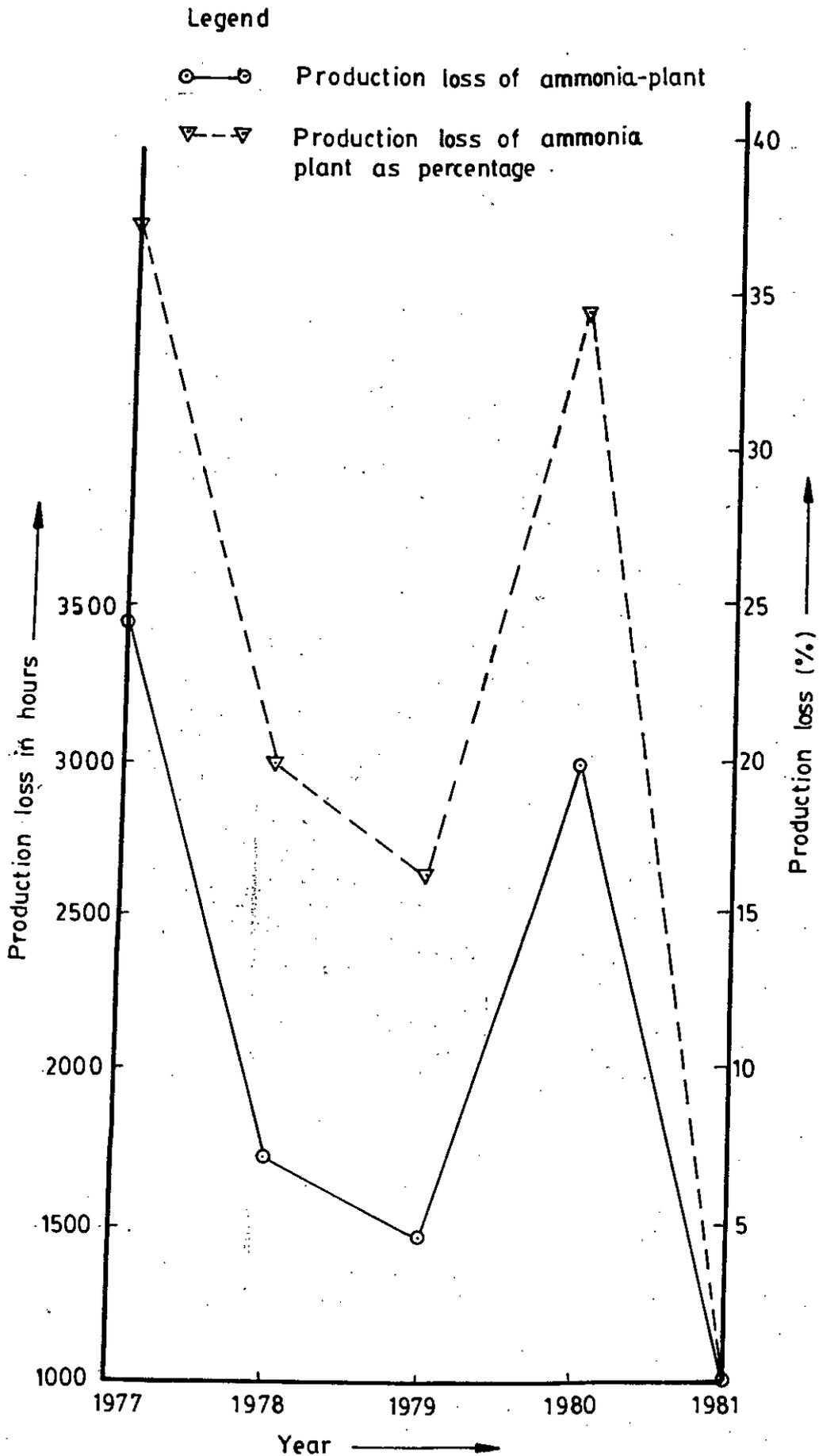


Fig. 5.7. Production loss of ammonia plant of UFFG (June 1977 to June 1981)

5.3.1 Down-time analysis (unit-wise)

The down-time of ammonia-plant include the down-time of its machinery and equipment and that of other plants which stops the ammonia-plant. The failure of different items depends on different causes. After careful analysis it can be found that the failure of different items depends on common factors such as:

- 1) High temperature
- 2) Trouble
- 3) Blockage
- 4) Damage
- 5) Leakage

Table 5.11 The influence of different causes for the failure of machineries and equipments in ammonia-plant of UFFG during 1977-1981.

Sl. No.	Name of Machineries equipments	Leakage (hour)	Damage (hour)	Bloc-kage (hour)	Trouble (hour)	High Temp. (hour)	Total (hour)
1.	Methanator effluent economizer	498.6		27.6			526.2
2.	Synthesis Compresor	232.32				246.48	478.8
3.	Vapor line of process Gas Reboiler	145.92					145.92
4.	Methanator Catalyst		284.40				284.40
5.	Air filter		233.28				233.28
6.	Askania valve of Hyper Compressor		170.40			144.00	170.40
7.	G.V. solution pump		248.88			144.00	392.88
8.	Refrigerant-Turbo-blower		56.40	82.32	46.32		185.04
9.	Demi water line	176.18					176.18
10.	Coupling of Gas-Air-Heater Oil pump		89.76				89.76
11.	Heat Exchanger			96.00			96.00
12.	PT line of PA-117		102.96				102.96
13.	Turbine of Synthesis Compressor					76.80	76.80
14.	Knock-out Drum of synthesis gas			27.6			27.60
15.	NG Emergency valve				24.00		24.00
Total		1053.02	1083.12	233.52	70.32	467.28	3010.22

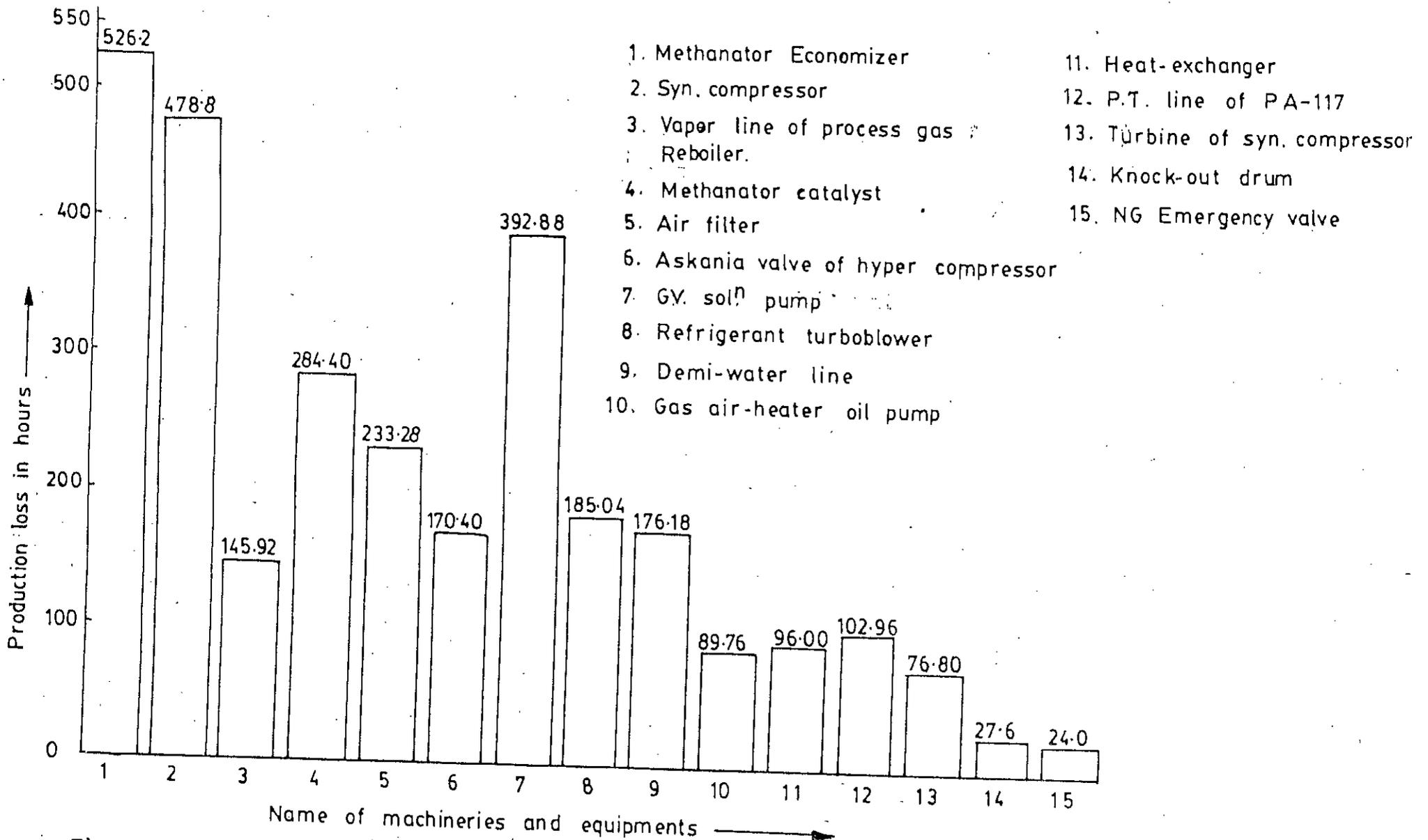


Fig 5.8 Down-time of machineries and equipments of ammonia plant in UFFG (June 1977 to June 1981).

### 5.3.2 Down-time Analysis (Factor-wise)

More detail analysis of the down-time is necessary to find out the different elements of the items which cause the failure.

Table 5.18, 5.13, 5.14, 5.15, 5.16 show the down-time of various machineries and equipments and the factors which are responsible for the same. The data are represented in matrixes.

Table 5.12 The failure of different parts of machineries and equipments of UFFG due to LEAKAGE (Down-time expressed in hour)

Name of Parts	Name of machineries and equipments					Total
	Methanator Effluent Economizer	Synthesis Compressor	Process Gas Reboiler of GV-section	Boiler of Feed Water Pump	Waste Heat Boiler	
1. Tubes	498.60					498.60
2. Vapor line			145.92			145.92
3. Inter cooler tubes		232.32				232.32
4. Down Stream line				111.00		111.00
5. Joint of FCV-524 & Feed Water line					65.18	65.18
Total:	498.60	232.32	145.92	111.00	65.18	1053.02

It is found that the tube leakage of Methanator Effluent Economizer gives high percentage of production loss. Inter-cooler tubes of synthesis compressor and vapor-line of Process Gas Reboiler of GV-section are also important for which repairing require long shutdown of the plant. Down stream line of Boiler Feed Water Pump and Waste Heat Boiler feed water line are less important in hampering the production because this type of leakage requires less time in repairing.

Table 5.13 The failure of different parts of machineris and equipments of UFFG due to DAMAGE (Down-time expressed in hour)

Name of Parts	Name of equipments and machineris								Total
	Methana- tor	EMV- 101	Air- Turbo- blower	Hyper Compre- ssor	Refri- gerant Turbo- blower	PA- 117	GAH Oil Pump	CV solu- tion pump	
1. Catalyst	284.4								284.4
2. Air Filter		156.00	77.98						233.28
3. Askania valve				170.4					170.40
4. Fuse breaker of instru- ment Panel					56.40				56.40
5. PT Line						102.96			102.96
6. Coupling							89.76	248.88	338.64
Total	284.40	156.00	77.98	170.4	56.40	102.96	89.76	248.88	1083.12

Methanator and GV solution pump coupling are the main item, which hamper the production at high percentage due to their damage. They require long shut-down for repairing and or replacement. Air filter, Askania valve of Hyper-compressor, PT line of PA-117 and coupling of Gas Air Heater (GAH) oil pump are the items which cause the production loss at considerable percentage. Fuse breaker of instrument panel of RTB is also a item of which damage hampers production at less percentage.

Table 5.14 The failure of different parts of machineries and equipments of UFFG due to BLOCKAGE (Down-time expressed in hour)

Name of Parts	Name of machineries and equipments				Total
	Heat Exchange of GV solution	Methanator Effluent Economizer	RTB	Synthesis Gas Knock-out-drum	
1. Tubes	96.00				96.00
2. Drain Valves		27.60			27.60
3. Luboil line			82.32		82.32
4. Level Glass				27.60	27.60
Total	96.00	27.60	82.32	27.60	233.52

Blockage of tubes of Heat Exchanger by GV solution causes the highest production loss. For cleaning of tubes of any Heat Exchange is a time consumable work. Blockage of lubricating oil line of RTB also causes production losses at high percentage. Drain valves of Methanator Effluent Economizer and Level Glass of Synthesis Gas Knock-Out-Drum are also the items for hampering Production.

Table 5.15 The failure of different parts of machineries and equipments of UFFG due to TROUBLE (down-time expressed in hour)

Name of Parts	Name of machineries and equipments		Total
	Steam regulating valve of Refrigerant-Turbo-Blower	NG Emergency valve	
1. Lift	46.32		46.32
2. Seat		24.00	24.00
Total	46.32	24.00	70.32

The lift of the steam regulating valve of RTB and NG emergency valve give some trouble i.e the valve lift and seat does not move up to the specified mark during operation.

Table 5.16 The failure of different parts of machineries and equipments of UFFG due to HIGH TEMPERATURE (down-time expressed and in hour)

Name of Parts	Name of machineries and equipments			Total
	Synthesis Compressor	Turbine of Hyper Compressor	GV solution Pump	
1. Radial bearing	246.48			246.48
2. Thrust bearing		76.80		76.80
3. Coupling			72.00	72.00
4. Bearing			72.00	72.00
Total	246.48	76.80	144.00	467.22

Radial bearing of synthesis Compressor, thrust bearing of Hyper compressor and bearing of GV solution pump fails due to high temperature. Failure of any bearing of Synthesis Compressor causes production loss at high percentage because its repairing and or replacement requires a long shutdown. In the above table it is found that bearing is the main part for which equipment fails due to high temperature.

CHAPTER → 6  
DISCUSSION AND CONCLUSION

It was discussed in Chapter 5 that ammonia plant is one of the plants, which is mainly responsible for the down-time of both the factories. The major units of the ammonia-plants which caused these failures are:

Table 6.1 Down-time of different units of ammonia-plants in NGFF and UFFG during the period of analysis

Name of units	Percentage of Total productive time		Percentage of down-time of ammonia plant	
	NGFF	UFFG	NGFF	UFFG
1. Main Gas Compressor /Synthesis Compressor	2.15	1.61	36.50	15.87
2. 1RF	1.34	-	22.05	-
3. 2RF	0.77	-	12.07	-
4. Ammonia-cooled Cu-solution cooler	0.51	-	8.65	-
5. Secondary Air-Compressor/ATB	0.374	-	5.34	-
6. Circulator/RTB	0.277	0.46	4.70	4.53
7. Methanator Effluent Economizer	-	1.77	-	17.45
8. Methanator	-	0.95	-	9.36
9. GV-solution pump	-	0.786	-	7.75
10. Air Filter	-	1.320	-	13.01
11. Turbine for synthesis Compressor	-	0.25	-	2.46
12. Turbine for RTB	-	0.15	-	1.47
13. Other equipments	0.63	2.80	10.69	27.61
Total	5.89	10.14	100.00	100.00

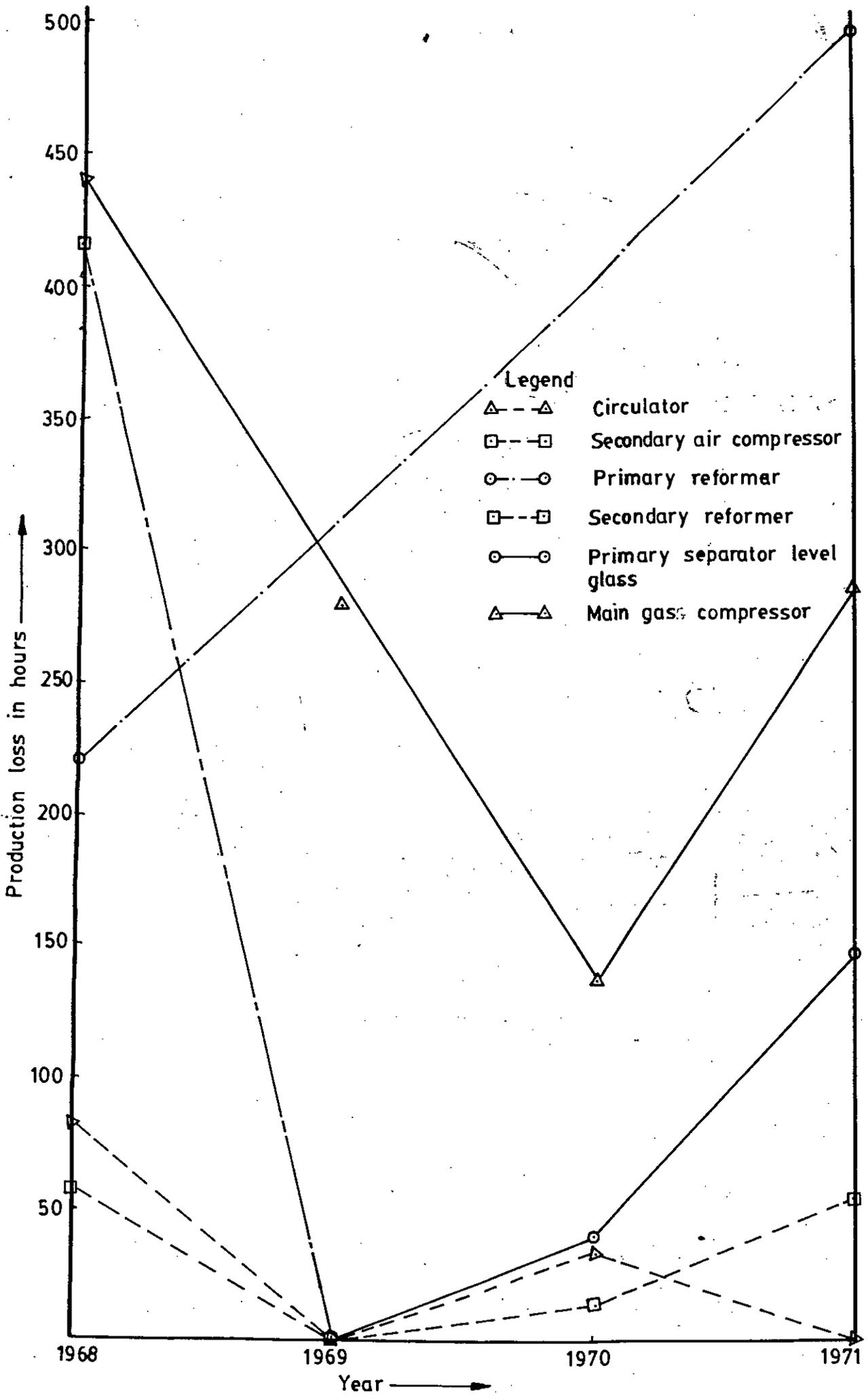


Fig 6.1. Down-time of different equipments of ammonia plant in NGFF

The major factors which caused the failure of ammonia-plants during the period of analysis are:

Table 6.2 Down-time of ammonia-plants due to different factors in NGFF & UFFG during the period of analysis

Name of factors	Percentage of total productive time		Percentage of down-time of ammonia-plant	
	NGFF	UFFG	NGFF	UFFG
1. High temperature	1.94	1.57	33.78	15.88
2. Leakage	1.43	3.55	24.27	35.00
3. Excessive vibration	0.60	-	10.18	-
4. Wear	0.09	-	1.52	-
5. Breakage	0.71	-	12.05	-
6. Damage	0.68	3.65	11.54	35.99
7. Blockage	0.55	0.78	9.33	7.69
8. Trouble	-	0.23	-	2.26
Total	5.89	10.14	100.00	100.00

In the following the various units which caused the down-time will be discussed to identify the operational suitability in both the factories.

#### Main Gas Compressor(MGC) /Synthesis Compressor:

In NGFF Main Gas Compressor is used to attain the synthesis pressure in four stage(reciprocating compressor).

Synthesis Compressor is multi-stage centrifugal compressor used for same purpose in UFFG.

Fig. 7.2 gives a relationship between down-time of MGC and the output pressure of the various stages. The equation is,

$$X - 0.947Y - 9.478 = 0 \quad (1)$$

where,  $X$  = Down-time of the compressor in hour  
 $Y$  = Output pressure of a compressor ( $\text{kg}/\text{cm}^2$ )

The range of synthesis compressor ( $23.6-150 \text{ kg}/\text{cm}^2$ ) is equivalent to the range of 2nd and 3rd stage ( $19.4-140 \text{ kg}/\text{cm}^2$ ) of MGC. Surprisingly the point due to synthesis compressor (shown as  $\otimes$  in Fig. 7.2) fits quite well to the equation 1. Thus from breakdown-time analysis it can be said that for the same working range the centrifugal compressor does not have an edge over the reciprocating compressor.

#### Primary Reformer (1RF)

The 1RF of NGFF caused a production loss of 4024.8 MT of urea due to its down-time whereas that of UFFG gives uninterrupted production during the period of analysis. In NGFF, primary reformation occurs at  $8.2-6.5 \text{ kg}/\text{cm}^2$  and  $399-620^\circ\text{C}$  whereas in UFFG it occurs at  $34.36-30.3 \text{ kg}/\text{cm}^2$  and  $410^\circ\text{C}-817^\circ\text{C}$ . The high pressure and temperature in the reformer tubes in UFFG has thus been found suitable. The design, material and manufacturing problems have not been studied in this case. But it is apparent that these problems will be less significant in UFFG.

### Secondary Reformer(2RF)

2RF causes production loss in NGFF whereas it does not hamper in production in UFFG. Secondary reformation occurs at 6.5-6.4 kg/cm<sup>2</sup>, and 620-816°C in NGFF whereas it occurs in UFFG at 30.3-29.7 kg/cm<sup>2</sup> and 817.8-1010°C. The high temperature and pressure in the brick-lining of 2RF of UFFG is suitable. The design and manufacturing problems of this lining are not studied. It is seems that these problems will be less significant in UFFG.

### Secondary Air Compressor(SAC)/Air-Turbo-Blower(ATB)

Secondary Air Compressor(reciprocating Compressor) of NGFF elevates the pressure of air to be used in 2RF. ATB (centrifugal compressor) is used in UFFG for same purpose.

From down-time analysis it is found that SAC causes production loss in NGFF whereas ATB gives uninterrupted production in UFFG. SAC works at 1-7.7 kg/cm<sup>2</sup> in NGFF whereas ATB works at 1-34 kg/cm<sup>2</sup> in UFFG.

Thus for this working range the centrifugal compressor has an edge over the reciprocating compressor.

### Circulator/Refrigerant-Turbo-Blower(RTB)

In NGFF, circulator is used to a elevate the pressure of the loop gases to the synthesis pressure. RTB is used for same purposes in UFFG.

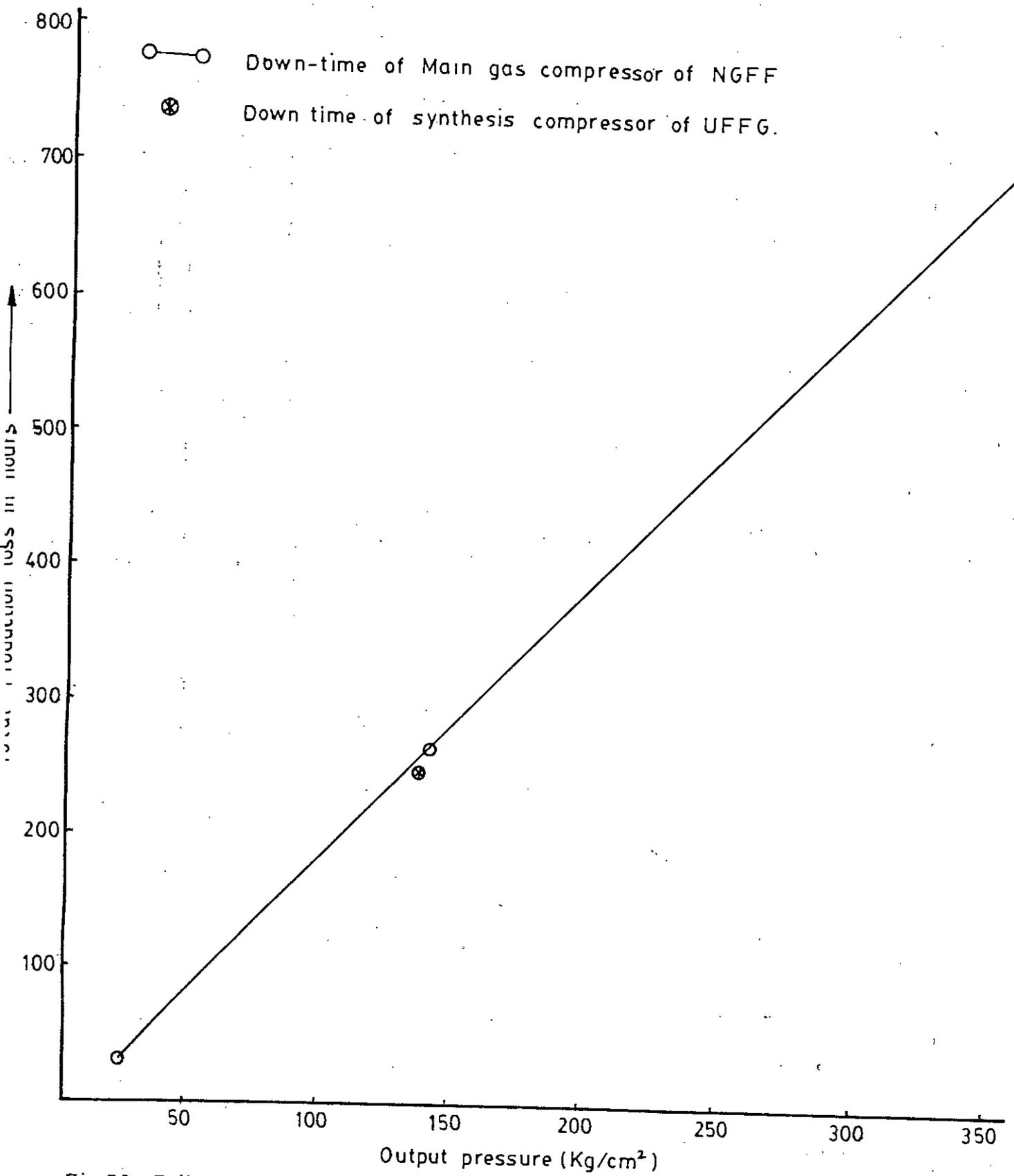


Fig.7.2 Failure of main gass compressor of NGFF and synthesis compr: of UFFG with output pressure (Ref. tables 6.2 and '6.11)

Circulator is a reciprocating compressor working at  $40.4^{\circ}\text{C}$  and  $330\text{-}365\text{ kg/cm}^2$  whereas RTB is centrifugal compressor working at  $129.6\text{-}150\text{ kg/cm}^2$  and  $40\text{-}50^{\circ}\text{C}$ . The observed down-time of the circulator is 147 hours, while that of RTB is 138 hours. Thus for the given working range, none is significantly better than the other.

It may be recalled that, the fourth stage of MGC in NGFF, which operates at  $137\text{-}365\text{ kg/cm}^2$  caused a down-time of 701 hours. This difference in down-time between the circulator and the fourth stage of MGC is mainly due to the working range. Similarly the difference in the down-time of centrifugal compressor in RTB as well as in synthesis compressor is due to the operating condition.

#### GV Solution Pump

GV solution pump of UFFG is used to circulate the GV solution ( $\text{CO}_2$ -absorber solution) from  $\text{CO}_2$ -regenerator to  $\text{CO}_2$ -Absorber. GV solution pump caused a production loss of 17821.9 MT of urea due to its down-time during the period of analysis. Out of 392.9 hours (13.01% down-time of ammonia plants) down-time of GV solution pump, 248.9 hours (63.26%) down-time is due to damage of coupling and 144 hours (36.65%) is caused by the high temperature at coupling and bearing.

Ammonia-Cooled Copper-Solution Cooler/Methanator, Methanator Effluent Economizer:

The process involved in urea production in NGFF requires

a Ammonia-cooled Copper-Solution Cooler for final purification of synthesis gases from carbon-oxide gases. In UFFG, for final purification of synthesis gases from carbon-oxide gases, methanation of carbon-oxide gases takes place in presence of catalyst Ni(Ni on calcium aluminate) in Methanator. After methanation, the synthesis gases is cooled in Methanator Effluent Economizer.

Ammonia-Cooled Copper-Solution Cooler causes 8.65% down-time of ammonia-plant due to the blockage of its heater tube. Methanator causes 9.36% down-time of ammonia plant due to the damage of its catalyst. Methanator Effluent Economizer caused a down-time of 520.2 hours(17.45% down-time of ammonia plant) while leakage of tubes 498.6 hours(94%) and blockage of drain valves 27.6 hours(6%) . The process of final purification of synthesis gases in UFFG caused a total down-time of 810 hours (25.91% of down-time of ammonia-plant).

Therefore, from above discussion and from table 6.1 it can be said, process of final purification, used in NGFF will give less down-time than that used in UFFG.

Table 6.2 shows the major factors which caused the failure of the ammonia-plants during the period of analysis. Both high temperature and leakage causes 58% of the down-time of ammonia plant of NGFF. But in UFFG, leakage and damage are responsible for 71% of the down-time. Though UFFG operates at

much lower pressure, but its failure due to leakage is more predominant than the case in NGFF (Table 5.2, 5.11, 6.2).

Damage of items such as instrument, valve plate, heater element, transformer, catalyst, coupling caused much higher down-time in UFFG (1083 hours, 35% of total down-time of ammonia plant) than in NGFF (368 hours, 11.5% of total down-time of ammonia plant). Coupling is the major cause in UFFG while the valve plates in NGFF.

In NGFF, high temperature at items such as 2RF, 1RF, MGC caused much higher down-time (1064 hours, 33.78% of total down-time of ammonia plant) than in UFFG (467 hours, 15.48% of total down-time of ammonia plant). The temperature in 2RF in NGFF is  $816^{\circ}\text{C}$  and the down-time due to high temperature in 2RF was 413 hours (12.07% of total down-time of ammonia plant). The 2RF in UFFG also works at  $816^{\circ}\text{C}$  and it does not have any problem. The cause of breakdown in NGFF may be due to product design.

High temperature at the bearing of Synthesis Compressor causes 246 hours (8.17% of total down time of ammonia plant) down-time. There is no such problem in NGFF.

## CHAPTER - 8

## SCOPE OF FUTURE WORKS

Scope of future study may extend over the following areas:-

- 1) Study of the design and manufacturing aspects of the troubled equipments keeping in view the technology transfer.
- 2) Detail analysis of the causes of failures and identification of the contribution of the plant inherent, operational and maintenance problems.
- 3) In-depth down-time analysis in urea plants.
- 4) Determination of scale of economy of urea fertilizer factories in Bangladesh with a view to achieve technological self-reliance in this field.

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## A-1 Gross National Products of Bangladesh at current prices

Sectors	(Million Tk)				
	1976-77	1977-78	1978-79	1979-80	1980-81 (Provisional)
1. Agriculture	53691	72248	78945	93299	104976
2. Mining & quarrying	2	2	2	2	2
3. Industry	8660	9403	10315	12511	15050
i) Large scale	4787	5298	5812	7051	6512
ii) Small scale	3873	4105	4503	5460	6538
4. Construction	5807	6155	7232	9289	11930
5. Power, Gas Water & Sanitary services	281	245	260	369	470
6. Transport, storage and communication	7334	8819	9547	11521	11902
7. Trade Services	10384	12832	15634	19048	21575
8. Housing services	8024	8249	9338	11464	12953
9. Public Administration & defence	3350	3457	3567	4231	5019
10. Banking & Insurance	1002	1034	1454	1833	2192
11. Professional and Miscellaneous service	6896	7846	8680	8883	9893
12. GDP at market prices	105361	130290	144774	172450	195962
13. Indirect Tax, net of subsidies(-) :	5720	6929	8142	11083	13316
14. GDP at current factor cost:	99641	123361	136632	161367	182646
15. Net factor in come from the rest of the world(+):	425	1229	1842	3470	4269
16. GNP at current factor cost	100066	124590	138474	164837	186915
17. Net National Pro- duct(income)	93878	117136	130224	154982	175833
Population(in million)	81.8	83.7	85.6	87.7	89.9
Per capita GDP at factor cost(Taka)	1218	1474	1596	1840	2031
Per capita GNP at factor cost(Taka)	1223	1488	1618	1885	2079
Per capita NNP (National Income)	1148	1399	1521	1767	1956

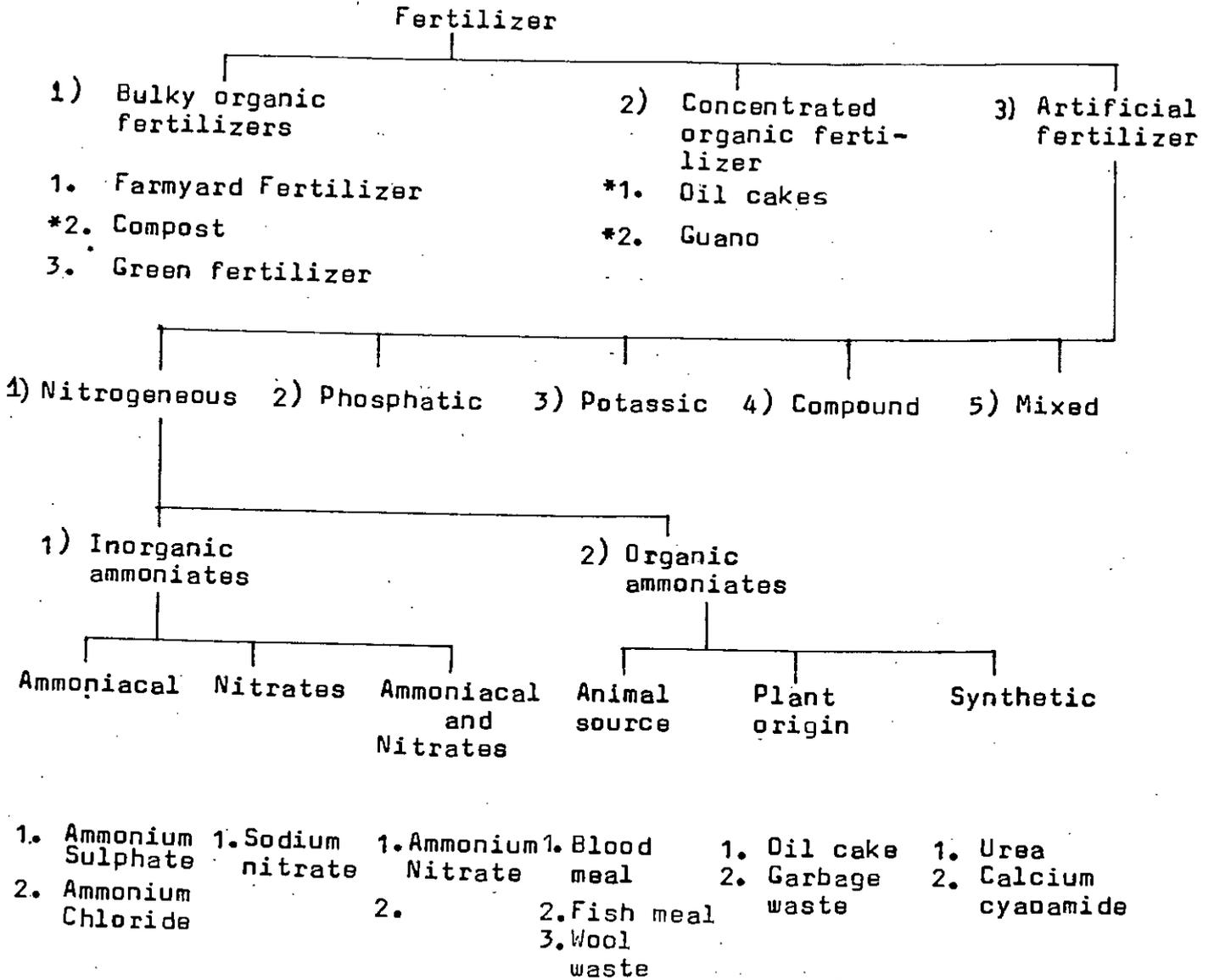
Source: Statistical pocket book of Bangladesh.

## A-2 Sectoral shares of GDP at current price

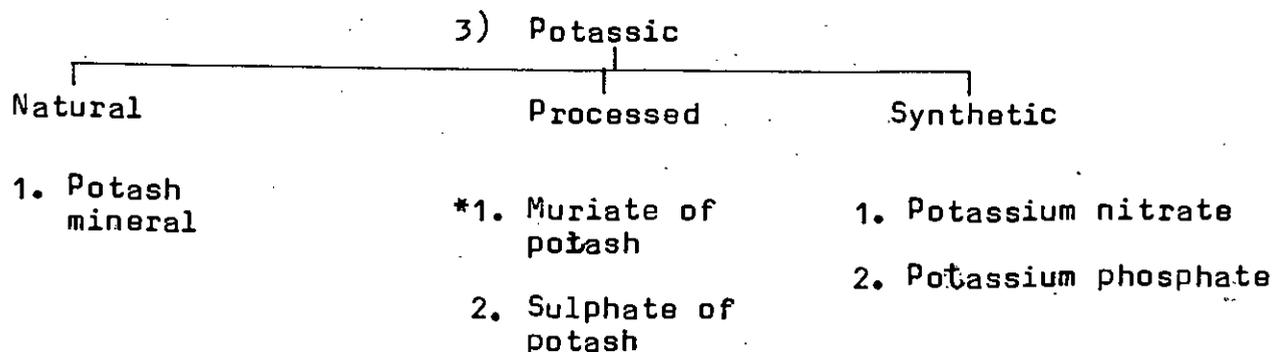
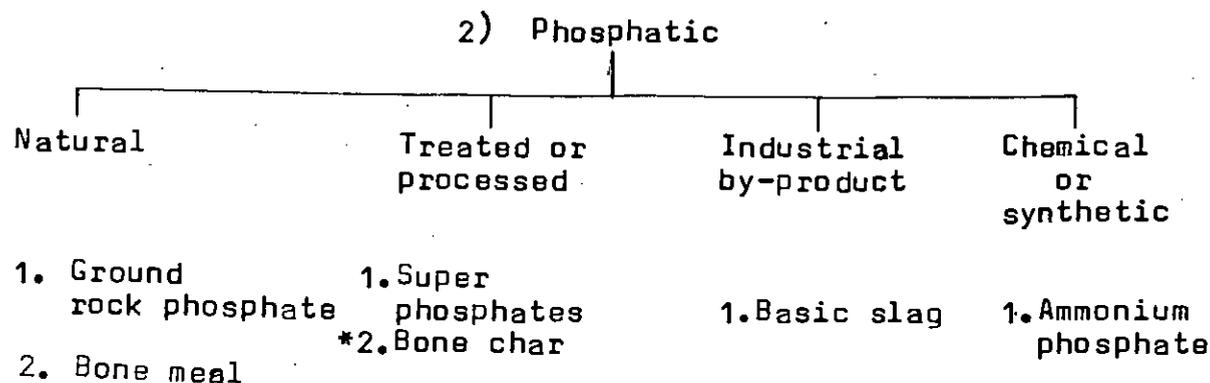
Sectors	Years				1980-81 (provisional)
	1976-77	1977-78	1978-79	1979-80	
1. Agriculture	50.9	55.5	54.4	64.0	53.6
2. Mining & quarrying	.002	.002	.001	.001	.001
3. Industry	8.2	7.2	7.1	7.3	7.7
i) large scale	4.5	4.1	4.3	4.1	4.4
ii) Small scale	3.7	7.1	3.1	3.2	3.3
4. Construction	5.5	4.8	5.0	5.4	6.1
5. Power, Gas, Water & Sanitary services	0.2	0.2	0.2	0.2	0.2
6. Transport, Storage and communication	7.0	6.8	6.6	6.7	6.1
7. Trade Services	9.9	9.8	10.8	11.0	11.0
8. Housing Services	7.6	6.3	6.4	6.6	6.6
9. Public Administration and Defence	3.2	2.7	2.5	2.5	2.6
10. Banking and Insurance	1.0	0.8	1.0	1.0	1.1
11. Professional & Misc. services	6.5	6.8	6.0	5.2	5.2
12. GDP at Market price	100.00	100.0	100.0	100.0	100.0

Source: Statistical pocket book of Bangladesh.

A-3 Scientific classification of fertilizers



## A-3 Scientific classification of Fertilizers(Contd)



\* Guana is a naturally occurring concentrate organic fertilizer, consisting of excrements and dead bodies of sea-birds and sea-animals. The fertilizer contains varying of Nitrogen 5 to 16%,  $P_2O_5$  8 to 25% and  $K_2O$  2 to 3%.

Bonechar is a product of bones obtained by heating the bones in the absence of oxygen and contains 30 to 35%  $P_2O_5$ .

Compost is the bulky organic fertilizer obtained by the decomposition of waste organic residues through the action of micro-organism.

Blood meal is a concentrated organic ammoniates containing 8 to 14% nitrogen, 0.3 to 1.5%  $P_2O_5$  and 0.5 to 0.8%  $K_2O$ .

Oil cakes are concentrated organic fertilizers containing varying amounts of nitrogen, phosphoric acid and potash.

Muriate of Potash is the popularly used potassic fertilizers made by refining crude mineral sylvinite or canallite and analyzes for 48 to 62%  $K_2O$ .

## A-4 Production of fertilizers by types in selected countries

(Thousand metric tons)

Country	Nitrogeneous		Phosphate		Potash	
	1976-77	1977-78	1976-77	1977-78	1976-77	1977-78
World	46223	49529	28211	30073	25259	25755
Bangladesh	126.1	105.7	20.2	17.6	-	-
Burma	555	* 555	-	-	-	-
Canada	1204.0	*1342.0	631.8	*650.6	*5655.4	*6206.5
China**	3842.0	4600.0	1348.5	1386.9	320.0	320.00
France	*1462.0	*1470.0	*1490.4	*1560.0	*1567.1	*1669.0
Germany Fed. Republic	1789.8	1304.7	733.1	722.5	*2217.2	2445.4
India	1862.4	199.7	478.3	670.00	-	-
Indoneisa	*184.2	*457.6	-	-	-	-
Iran	136.2	177.9	109.0	110.4	-	-
Italy	985.3	1028.7	*282.6	493.7	141.1	134.2
Japan	1171.0	1446.0	625.0	696.0	-	-
Korea, Rep.	534.5	668.8	214.8	309.3	-	-
Pakistan	309.3	306.7	11.9	13.5	-	-
Romania	1331.0	1881.0	493.0	548.0	-	-
Turkey	189.0	189.3	344.2	275.8	-	-
U.K.	1071.0	1199.0	438.0	437.0	149.00	1204.0
USSR	8531.0	9025.0	5226.0	5586.0	8310.0	8347.0
USA						

\* Estimated figure

\* FAO Estimate

Source: 1. UN Statistical year book 1978.

2. Bangladesh data Statistical Pocket book of BD 1979.

A-5 Consumption of fertilizers per hectare of agricultural area  
in selected countries

Country	( 1000 grams N, P <sub>2</sub> O <sub>5</sub> , K <sub>2</sub> O)			
	Years			
	1961-65	1966	1971	1976
World	8.5	11.4	15.9	20.8
Australia	1.7	2.4	1.9	-
Bangladesh	4.2	7.9	11.8	24.2
Burma	0.7	1.1	4.1	4.9
Canada	8.3	12.6	12.8	20.7
China	4.5(F)	8.5(F)	13.1(F)	18.4(F)
France	82.3	99.8	152.0	157.2
Germany F.R.	185.6	196.5	244.3	256.7
India	3.4	7.0	14.8	18.8
Indonesia	4.3	5.0	8.0	16.7
Italy	95.3	54.5	82.1	210.1
Japan	298.9	347.7	333.6	291.1
Korea DPR	76.3	73.0	161.0	195.8
Korea Rep.	156.5	183.2	261.3	285.1
Netherlands	232.2	257.1	282.5	302.2
New Zealand	27.7	29.9	35.0	37.8
Norway	154.4	169.7	224.3	248.7
Pakistan	2.7	4.5	15.7	25.7
Philippines	11.9	13.6	26.2	31.8
Srilanka	36.4	38.3	34.9	35.7
Thailand	2.1	5.1	9.0	13.2
U.S.A.	21.3	29.1	35.9	46.7
USSR	6.2	10.3	17.1	27.9
Vietnam	13.2	11.7	28.6	36.8

Source: F.A.O. Annual Fertilizer Review 1977.

1 Hectore = 2.4711 acres

F = F.A.O. Estimate.

## A - 6 Production fertilizers by types in Bangladesh.

(In tonnage)

Year	Urea	TSP	Ammonium Sulphate
69-70	94,500	-	-
70-71	51,200	-	-
71-72	47,000	-	-
72-73	207,200	-	5,006
73-74	274,590	-	10,129
74-75	68,011	32,337	4,850
75-76	275,541	48,190	5,710
76-77	287,343	47,222	9,114
77-78	208,715	37,667	9,304
78-79	290,067	57,120	5,244
79-80	355,803	66,443	9,404

Source : Bangladesh Chemical Industries Corporation.

A-7 Distribution of fertilizers by type and by nutrient content in Bangladesh

Year	Urea	TSP	MP	HP	Total	(000 Tons)				Total
						N 46% of Urea TSP	P <sub>2</sub> O <sub>5</sub> 60% of TSP	K <sub>2</sub> O 60% of MP	NH <sub>3</sub> CO <sub>3</sub> 28% of H.P.	
1970-71	212.34	74.81	17.08	-	304.30	97.67	34.44	10.24	-	142.35
1971-72	169.71	60.13	13.92	-	249.82	78.06	27.65	8.35	-	114.06
1972-73	275.64	88.91	18.47	-	384.02	126.89	40.89	11.08	-	178.76
1973-74	267.67	93.82	18.39	-	379.83	120.36	43.15	11.03	-	174.54
1974-75	176.06	76.08	17.62	11.42	281.20	80.20	34.99	10.57	3.19	129.23
1975-76	311.96	110.06	21.74	4.36	448.12	143.50	50.62	13.04	1.22	208.38
1976-77	349.26	125.59	22.38	4.03	505.27	160.66	55.77	13.43	1.13	230.99
1977-78	477.45	191.33	41.00	3.27	713.05	219.63	88.01	42.60	0.92	333.16
1978-79	470.56	177.65	47.50	3.33	704.17	216.46	81.72	28.50	0.93	327.61
1979-80	535.98	206.19	88.24	3.11	841.97	246.55	94.89	32.94	0.89	395.20

Source : Bangladesh Agricultural Development Corporation.

A-8 Gas analysis(Mol-%) of Natural Gas of Haripur  
Gas Fields(NG of NGFF)

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Methane	96.26
Ethane	1.99
Butane & Higher	0.32
Propane	0.14
Nitrogen	0.95
Carbon-di-oxide	0.54
Total Sulphur(Grain/100 SCF)	0.298

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Specific Gravity(Air-1)	0.575
Calorific Value (BTU/SCF)	1052
Gas Pressure(before Compression)	450 PSIG.

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Source: Petro Bangla

## A-9 Specification of NG of Titas Gas Field(NG of UFFG)

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1.	Specific gravity(air=1)	= 0.574
2.	Carbon/Hydrogen(Wt.ratio)	= 3.036
3.	Molecular weight	= 16.616
4.	Net calorific value(at 16.6°C)	= 8761 K.cal./NM <sup>3</sup>
5.)	Composition	Volume %
	Carbon-di-oxide	0.21
	Nitrogen	0.28
	Methane	97.11
	Ethane	1.77
	Propane	0.38
	Iso-butane	0.09
	N-butane	0.06
	Iso-pentane	0.03
	N-pentane	0.02
	Hexanes	0.03
	Heptanes	0.02
	Sulphur	10 ppm.
6.	Gas pressure(before compression)	24.5 to 28.0 kg/cm <sup>2</sup> G.
7.	Gas temperature	32°C

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Source : Petro Bangla

A-10 Actual Production of urea and its capacity utilisation during the period of analysis

NEFF				UFFG			
Year	Capacity (MT)	Actual Production (MT)	Capacity Utilisation (%)	Year	Capacity (MT)	Actual Production (MT)	Capacity utilisation (%)
1967-68	106,000	73,969	69.78	1977-78	340,000	151,037	44.42
1968-69	106,000	54,315	51.24	1978-79	340,000	236,123	69.44
1969-70	106,000	60,852	57.40	1979-80	340,000	240,614	70.76
1970-71	106,000	37,637	35.50	1980-81	340,000	245,482	72.20
1971-72	106,000	35,156	33.16	1981-82	340,000	252,863	74.37
Average production		52,385	49.4			228,423	67.2

A-11 Actual production of urea in each flow-line of UFFG and NGFF during the period of analysis.

Factory	Total operating hours	Capacity			Average production		
		Annual (MT)	Total (MT/Hr)	In each flow line (MT/hr)	Annual (MT)	Total (MT/hr)	In each flow line (MT/hr)
NGFF (Jul. 1967 to Jun. 1972)	46,834	106,000	14.1	7.0	52,385	11.2	5.6
UFFG (Jul. 1977 to Jun. 1982)	24,824	340,000	47.2	47.2	228,423	45.3	45.3

A-12 Working pressure and temperature of compressors of NGFF & UFFG

Name of equipments	NGFF						UFFG							
	Stage	Input		from	Output		Name of equipment	Input		From	Output			
		Press kg/cm <sup>2</sup>	Temp. (oC)		Press kg/cm <sup>2</sup>	Temp (oC)		Press kg/cm <sup>2</sup>	Temp. (oC)		Press kg/cm <sup>2</sup>	Temp. (oC)		
1. Main Gas Compressor	Ist	6	378	Knock out Drum	20	107	CO <sub>2</sub> absorber	28	32.5	1.NG compr.	Filter	35	45	Feed Stock Steam Heater
	2nd & 3rd	19.4	40	Gas after cooler	140	40.4	Cu-liq. scrubber	23.6	40.0	2.Synthesis compr.	Syn. knock out Drum	150	170	Synthesis Hot Exchanger
	4th	137	7.2	High Pressure separator	365	40.4	Filter & then cold exchanger	129.6	40	3.Recirculator	Syn. Ist Cold Exchanger	150	55	Syn. Hot Exchanger
2. Circulator	-	330	40.4	Primary separator	365	40.4	Filter & then Cold Exchanger	Atmospheric press and temp.	4.Air Compr.	Atmospheric	Atmosphere	34	140	2nd steam superheater
3. Secondary Air-Compr.	-	Atmospheric pressure & temp.		Atmosphere	7.7	37.8	Air-Receiver							

A-13 Productive and Non-productive hour of NGFF and UFFG during the period of analysis

Factory	Period of analysis	Available hour	Non productive hour due to					Non Prod. hour	Productive hour						
			over hauling	Power failure PDB	fail- from	Risk Created by political background	Cre- Non Supply of Natural gas			Lub.oil					
			To- -tal	%	To- -tal	%	Total %	To- -tal	%	To- -tal	%	Total %			
NGFF	Jan./68 to Dec./71	63,360 *	3977.74	6.27	-	-	5507.6	8.69 **	6	5305	.837	10021.8	15.01	53338	84.18
UFFG	Jul/77 to Jun/81	35040	3967.9	11.32	526.56	1.5	884.4	2.52		5378.9	15.35	29661.1	84.64		

\* From 1.01.68 to 01.02.71 include 42 months, but 4 months deducted because their down-time analysis is not obtained. For double flow line of production, total available hour = 44 months x 30x24 = 63,360 hour.

\*\* During 1971 war and liberation movement in 1969.

A- 14 Down-time of different machineries and equipments of Ammonia-Plant in NGFF during January 1968 to December, 1971

Total production hours	Main Gas Compr.	Ist Reformer	2nd Reformer	Heater Tube of Cu-liq. section	Secondary Air Compr.	Circulator	Other equipments
Total Perct. (hr)	Total Perct. (hr)	Total Perct. (hr)	Total Perct. (hr)	Total Perct. (hr)	Total Perct. (hr)	Total Perct. (hr)	Total Perct. (hr)
53,338.2	1151.1	720.75	413	273.75	199.8	147.8	337.66
	2.15	1.34	0.77	0.51	0.374	0.277	0.63

A-15 Down-time of different machineries and equipments of Ammonia-Plant in UFFG during July 1977-Jan, 1981

Total Productive hours	Syn. Compr.	Methanator Effluent Economizer	GV. Soln. pump	Methanator Catalyst	Air Filter	RTB	Turbine for RTB	Turbine for Syn. Compr.	Other Equipments	Loss of Productive hours
Total (hr)	Per-ct.	Total (hr)	Per-ct.	Total (hr)	Per-ct.	Total (hr)	Per-ct.	Total (hr)	Per-ct.	Total (hr)
29661	478.8	526.2	392.9	284.4	233.3	138.7	46.3	76.8	832.8	3010.2
	1.61	1.77	1.32	0.95	0.786	0.56	0.10	0.25	2.7	10.14

