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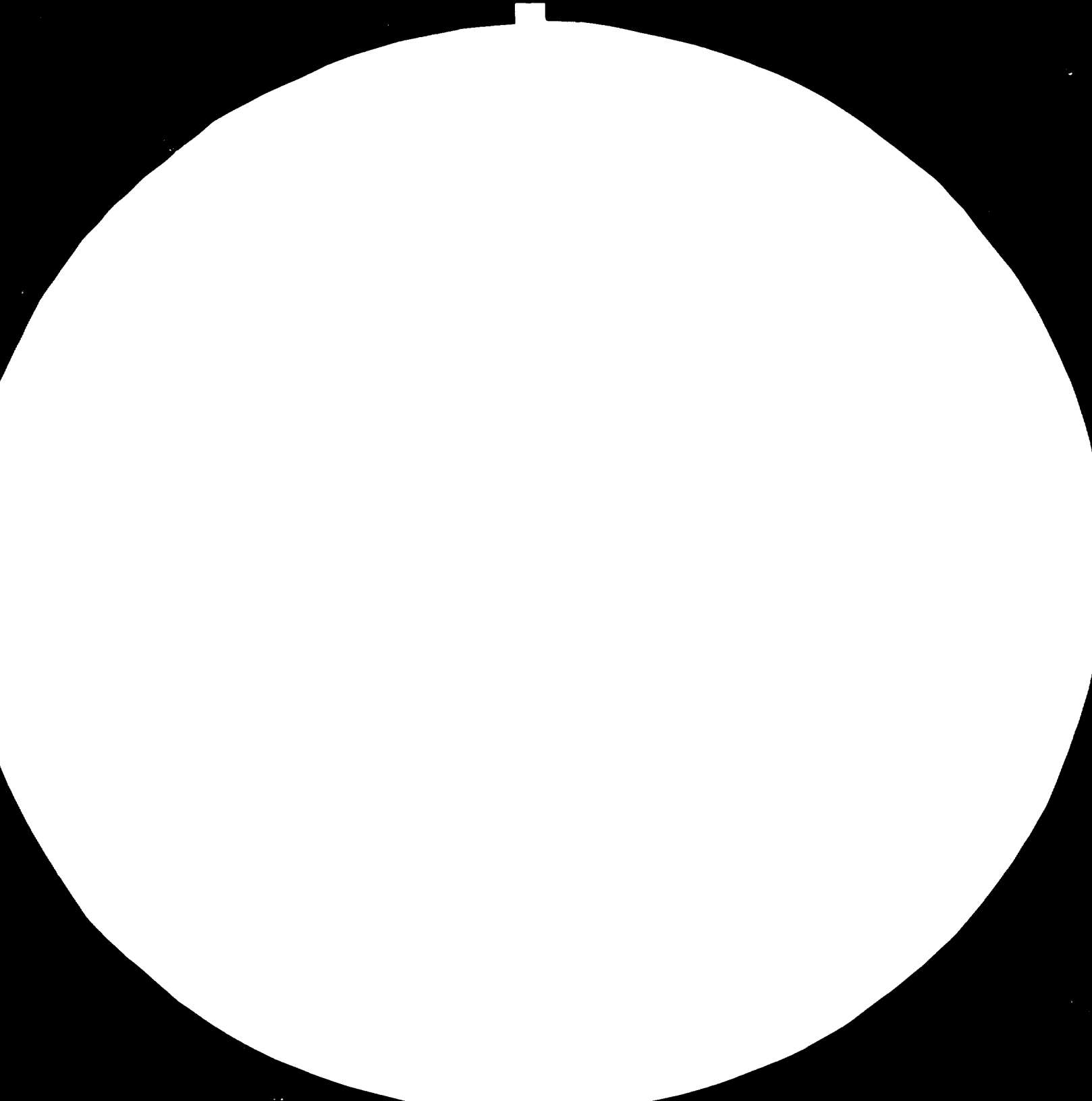
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Microcopy Resolution Test Chart
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MODEL BASIC ENGINEERING DESIGN
FOR A
100 MTPD AMMONIA PLANT

FINDINGS & CHARACTERISTICS OF
100 MTPD AMMONIA PLANT

UNIDO CONTRACT No. 82/48
PROJECT No. UD/INT/82/101
ACTIVITY CODE: UD/INT/30.3

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UNITED NATIONS INDUSTRIAL DEVELOPMENT ORGANISATION

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PROCESS FLOW DIAGRAM

COAL GASIFICATION SECTION	9484-237-11121
STEAM REFORMING SECTION	9484-237-11120
AMMONIA SYNTHESIS SECTION	9484-237-11107

CHAPTER-I

1. INTRODUCTION

- 1.1 At the NFC-UNIDO Seminar on MINI-FERTILIZER PLANTS, held at Lahore on November 15-20, 1982, FEDO had presented a design document for a 100 Metric Tonnes/day Ammonia Plant. A modular concept had been adopted for the design using established process steps. In line with the logic of long-term availability of raw-materials in the majority of countries, two designs based on Coke and Fuel Oil as raw-materials and with many common elements were presented in the document. The reasons for the choice of these two raw-materials had also been explained.
- 1.2 The discussions at the Seminar favoured a modified approach for the final design study. Based on current plant experience and the expected availability of small pockets of Natural Gas in many developing countries, it was decided that designs based on Coal and Natural Gas also be done to supplement the available study on Fuel-oil. This has now been done, adopting fresh approaches from those previously followed. Two basic designs, one based on Natural gas and another based on Anthracite Coal are presented in this document. In effect, taking both documents together, UNIDO have four different designs and using the modular principle, many possible variants.
- 1.3 The very conception of mini-plants has been born out of the recognition of fertiliser as an important input in the economic evolution of Developing Countries and the uncertainties affecting supplies and prices following on the oil price crisis of 1973-75. The heavy impact of higher transport charges on prices favours local manufacture. To many countries the introduction to sophisticated industry could be through local mini-plants which will help to up-grade technical capabilities and promote better agriculture.

- 1.4 The rising price of energy has focussed attention on energy consumption per unit of fertiliser output. Large single stream plants are favourably placed on this score, since the effective gathering of fugitive waste heat is more easily done in these units. Most plants are self-contained in power requirement by adoption of designs which use process heat for meeting this need. A matching of available heat with the imperatives of design of principal machinery confine these achieved high efficiencies to the larger sizes of plants. A limitation on energy efficiency is placed on Mini-plants by the need to depend on purchased or generated electric power for the principal drives. Direct comparisons of energy consumption per unit of product are therefore not in order. However, it may be noted that a 1979 Study on Ammonia plants in North America*, where a large number of the most advanced Ammonia synthesis plants are installed, shows an energy consumption of 11.209×10^6 kilo calories per Metric Tonne of Ammonia. This figure may be taken as a fair index for comparison of new designs.
- 1.5 In presenting the first document the logic which governed the selection of preferred raw-material for a design of plant for wider application had been discussed (para 3.1 to 3.4). The assumption of the availability of local pockets of Natural gas based on the fairly large number of countries with potential (listed by BEICIP, a Bureau of the Institute Francaise du Petrole in 1980) is the logic for the Natural gas alternative presented here. It is assumed that many of these deposits are recoverable economically for local consumption and a price of about \$4 per 1000 cft. has been assumed in the costing. Anthracite Coal is a commodity in abundant supply and is the raw-material assumed for the second design.

[*ENERGY AND FERTILIZER : IFDC TECHNICAL BULLETIN IFDC-T-19)

2. PROCESSES

2.1 Ammonia Synthesis

The design of the Ammonia Synthesis module is common for both alternatives. From the receipt of purified Synthesis Gas till the delivery of liquid Ammonia into storage is the compass of this module. In order to reduce the number of rotating machinery, the simplest Ammonia process has been used, namely, the Casale Process which requires only the Synthesis Compressor. The system operates at 400 kg/cm² which does not present any problems of availability of equipment at the small size of Plant being considered. No refrigeration machinery is involved. Circulation is by ejector which is trouble-free. Maximum recovery of Ammonia is effected by scrubbing the Purge gases and distilling off the Ammonia. The residual purge gases are available as fuel for process or off-sites use.

2.2 Synthesis Gas - based on Natural Gas Reforming

2.2.1 The design adopted is based on the orthodox flow-sheet of Natural Gas Reforming Plant. Natural Gas at 40 kg/cm² is preheated before passing to sulfur removal by Zinc Oxide. The gas is then mixed with steam and reformed in two steps, primary and secondary. Heat recovery is effected from the flue gas by heating process streams and combustion air. The heat in the process gases is used to raise steam for process purposes. There will be a surplus of steam which can be used in off-sites.

2.2.2 CO conversion in two steps, High Temperature & Low Temperature reduces CO in the process gas stream by combining CO with steam to give volume for volume conversion to Hydrogen. The gas stream which contains Hydrogen and Nitrogen as well as CO₂ is then processed for removal of CO₂.

- 2.2.3 The Benfield Process, which is widely accepted has been adopted in this case for the CO₂ removal step. Physical absorption processes like Selexol Process are available; but are yet to make an impact and hence choice of the more widely adopted Benfield Process. The bulk of CO₂ in the gas stream is removed in this step. The output gas stream is essentially in the ratio needed for Ammonia Synthesis and needs only the removal of the small fraction of Carbon Oxides still left. This is done by methanation which converts carbon oxides to process-neutral methane. The gas is now ready for compression which is the first step in the synthesis module.
- 2.3 Synthesis Gas - based on Anthracite Coal
- 2.3.1 Even though Coal and Coke were the raw-materials used at the beginning of the Ammonia industry, these widely available raw-materials were replaced by the more easily handled Petroleum derivatives, Naphtha and Fuel Oil where Natural Gas was not available. The changing pricing structure of fuels provokes a revision of attitude to traditional fuels. There are several popular processes for their use in gas making which have made their strong presence in improved versions in recent years. For use in Mini-Ammonia Plants, where simplicity and dependability are important criteria, the Wellman Gasifier using enriched air is being proposed. The use of PSA unit for providing the enriched air is a simpler alternative to the cryogenic units, normally used for Partial Oxidation Processes.
- 2.3.2 The PSA unit uses molecular sieves and the pressure swing principle to provide a supply of enriched air with oxygen content of 46% which will ensure a properly proportioned Synthesis Gas at the end of the gas making module. Since the Gasifier itself operates at near atmospheric pressure the enriched air is delivered at low pressure.

- 2.3.3 The Wellman Gasifier is a well established equipment which has proved itself in many installations. Multiple units are used (three in this case) so that even in the event of maintenance of one unit, Gas supplies to the synthesis plant is not interrupted. The gas leaving the producer is scrubbed free of dust and higher hydrocarbons and sent to a gas holder which serves to even out process flows.
- 2.3.4 The gas has to be scrubbed free of any sulphur content before further processing. The Benfield Hi-pure Process is used after compressing the gas to about 35 kg/cm². Since the Benfield Process is used after CO-conversion to remove CO₂ from the gas stream, the Hi-pure Process fits well into the general scheme. The compressing of gas to the intermediate pressure enables the purified gases to be delivered to the synthesis module at the same pressure as in the case of the Natural Gas alternative.
- 2.3.5 CO-conversion is carried out in High Temperature and Low temperature Stages. The Carbon Di-oxide formed is removed in the absorption towers of the Benfield Process and the gas passes on to Methanation. With carbon oxides reduced by conversion to Methane to a residual of less than 10 ppm. the gases enter the suction of the synthesis compressor.
- 2.3.6 The Gasification module is tailored around the Wellman Gasifier which is considered simpler and easier to operate than other coal gasification equipment. It will be noted that unit production costs are higher in this case in spite of lower raw-material costs largely on account of electricity and steam costs as well as depreciation. The electricity and steam consumptions

suggest a Co-generation cycle Power Generation at the site itself thereby making substantial savings due to higher energy efficiency.

3.0 DOWNSTREAM PLANTS

The recommendation made regarding down-stream plants in Chapter-I - para-7 of the earlier document is retained. Product costs have been presented based on new costing of Ammonia.

4.0 GENERAL

4.1 In presenting revised designs the following requirements have been met.

4.1.1 Alternate designs for the use of Coal and Natural Gas have been prepared as required. The modular pattern enables the substitution of part of the Fuel Oil design earlier presented to combine with the new Ammonia Module.

4.1.2 The design for Natural Gas processing follows classical steps since none of the newer process steps publicised have been established yet and the savings indicated are marginal and still to be proven. The design for coal has used the simplest and fairly widely accepted Gasifier in combination with a standard PSA oxygen enrichment unit. Benfield Process has been adopted for both routes as it is presently the most widely accepted purification step. The Benfield Hi-pure Process for removal of sulfur incorporates easily into the system.

4.1.3 The generation of power has to be done in an off-sites unit. The surplus steam from the Natural Gas Reforming process and the requirement of power and steam at low and

medium pressure in the Coal Gasification Process can be suitably integrated with the off-sites unit. This may be done more effectively in specific cases since locational factors are involved.

4.1.4 Investment and production costs based on a location in India have been presented.

4.1.5 Two Gasification modes and a Synthesis Process have been provided in this document. Each of these processes can be standardised with limited modifications made for local factors. Moreover the modules within each group can be replaced if an improved process becomes available. Particular attention has been paid to choosing the most easily operable process units and schemes.

4.2 Energy Efficiency

4.2.1 Quoted below is an extract from IFDC Bulletin "ENERGY & FERTILIZER":

"There do not seem to be any commercially feasible technological breakthroughs in Nitrogen manufacturing that would result in major energy saving".

What this means is that while further energy savings are desirable, those publicised are at best marginal and unlikely to affect the perceived advantages of economy of local manufacture. Quoting again from the same source "For example, operation of an Ammonia Plant at 60% capacity may use 25% more energy per tonne of product than continuous operation at full capacity", it is clear that the utmost emphasis needs to be paid to design features which make uninterrupted operation possible.

4.2.2 In the larger plants with complex process linkages, it has been observed that a single interruption in a month increases the monthly average consumption of energy by as much as 10%. The ability to operate for long periods without interruption, engineered with comparative ease into the simpler designs presented in this document may therefore be their best recommendation of Mini-Plants for developing countries.

CHAPTER II

MINI AMMONIA PLANT

PROCESS

1. Natural Gas steam reforming

1.1 Design basis: 100 MT per day of Anhydrous Ammonia

1.2 Raw Materials: Natural gas of following composition by volume % equivalent to make up gas required for 100 MT per day of Anhydrous Ammonia:

CH ₄	98.02
C ₂ H ₆	1.35
C ₃ H ₈	0.4
C ₄ H ₁₀	0.12
N ₂	0.01
O ₂	0.1
Total S	10 ppm V
Lower Heating Value)	8707 Kcal/MM ³ GAS
)	

1.3 The Plant

1. The feed natural gas of above composition is assumed to be available at the battery limit at 40 Kg/cm²g pressure and ambient temperature. Natural gas is used for fuel use also.
2. Auxiliary steam raising facility has been provided as a buffer to meet the steam requirement in the plant during start-up and other abnormalities. The plant is self sufficient in steam requirement at normal plant load.
3. Although only Zinc Oxide has been, at present, provided for the feedstock sulphur removal, provision has been made for Comox catalyst bed also for use, if required.
4. Benfield process has been adopted for the decarbonation of the make-up gas.

1.4 Plant Description (Drg. No. 9484-237-11120)

The natural gas at 40 Kg/cm²g pressure at the battery limit is freed of liquid condensate in the knock out drum 301 and is preheated to 400°C in the preheater 581 for the feed gas desulphurisation. The natural gas for fuel requirement is tapped from the main stream prior to the preheater but after the liquid knock-out drum. The feed gas at 400°C from the preheater is passed through the ZnO bed in the desulphurisation vessel 302. The desulphurised feed is then mixed with the required quantity of process steam to make up a steam carbon ratio of 3.5:1 at the Primary Reformer inlet, preheated again to 510°C in the Primary Reformer convection train and fed to the primary reformer 582. The Reformer is a conventional top fired furnace with 64 numbers of centrifugally cast tubes of HK 40 alloy and 40 numbers of gas fired burners in the radiant box. Following equipments are housed in the convection train of the reforming furnace :

- 1) Feed Preheater
- 2) Process Air Preheater
- 3) Process Steam Superheater
- 4) Flue Gas Boiler and
- 5) Combustion air preheater

The reformer effluents at about 820°C and a methane slip of 10.78% is further reformed in the Secondary Reformer 303. This is an adiabatic reactor wherein required quantity of the preheated process air at 470°C is admitted and further reforming of the Methane takes place at a higher temperature than in the Primary Reformer.

The Secondary Reformer effluent at 1000°C with a methane content of 0.33% is then cooled to 380°C in the Reformed Gas Boiler 502. The hydrogen content in the gas is further enhanced by conversion of the carbon monoxide to carbon dioxide in presence of steam in the high and low temperature shift reactors 305 and 306 respectively. The gas is fed to the high temperature Shift Converter 305 at 380°C.

The exothermic heat in the HTS effluent is removed in the converted gas boiler 503 and process heat exchangers 504 and 505 and the temperature reduced from 440°C to 223°C at the inlet of the low temperature Shift Reactor (LTS) 306.

The carbon monoxide content in the gas exit of the LTS is 0.3% and carbon dioxide about 17.9%. The heat content in the gas is used for the reboiler duty in the decarbonation section based on the Benfield process.

The carbon dioxide in the synthesis make up gas is reduced to 0.1% in the decarbonation section. The solution is 30% potassium carbonate, 3% DEA and Vanadium Pentoxide as inhibitor. SS 304 slotted rings are used as the tower packing.

The carbon oxides in the gas exit the decarbonation section are converted into Methane in the Methanator 311. The total inerts in the make-up gas after the Methanator will be 1.15% consisting of Argon 0.32% & Methane 0.83%.

The gas at 24 kg/cm^2 pressure and 45°C is sent to the ammonia synthesis section as make-up gas.

2. COAL GASIFICATION SECTION

2.1 Design Basis: 100 MT per day of Anhydrous Ammonia.

2.2 Raw Materials: Anthracite coal of following proximate analysis :

H ₂ O	:	6.4 %
Volatiles	:	4.5 %
Fixed Carbon	:	78.9 %
Ash	:	10.2 %
Calorific Value	:	6800 Kcal/kg.

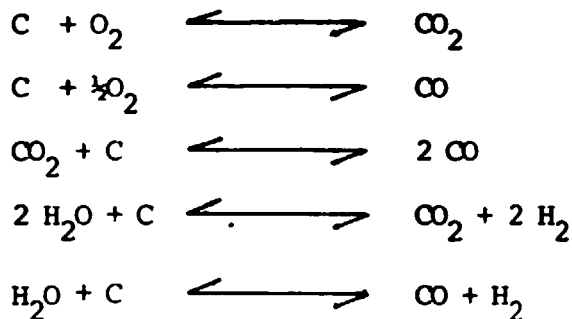
2.3 The Plant:

Pressure Swing Adsorption (PSA) System has been included in the scheme for meeting the requirement of enriched air with 46% oxygen for the gasifiers and also to ensure required quantity of nitrogen in the synthesis make-up gas.

2. The gasifier is a fixed bed type operating at near atmospheric pressure.
3. A gas compressor has been incorporated to boost the gas pressure from atmospheric to medium pressure of 35 kg/cm²g prior to the gas purification section.
4. Hi-Pure Benfield process has been considered most suitable for the gas desulphurisation and decarbonation processes.
5. Final purification of the gas is by Methanation where the carbon oxides are reduced to less than 10 ppm level.

2.4 Process Description [Drg.No.9484-237-11121]

The production of raw gas is by using Wellman Producer. The gas is produced by controlled partial oxidation of coal in the fixed bed type producer in an atmosphere of oxygen provided by enriched air. Steam is used to control the combustion. The gas generated contains mostly carbon monoxide and Hydrogen along with Nitrogen, carbon dioxide as well as solid impurities. Major reactions in the producer are



Three gasifiers, 583, are used in parallel for the production of the raw gas. The coal is fed from a hopper situated over the gasifier.

The gasifier consists of a water jacketed retort, a two compartment fuel bin with segmental gates, force feed pipes, rotating grates, ash bin etc. The upper compartment of the fuel bin serves as a storage bin while lower compartment functions as sealed continuous fuel feed system. Coal from the gas tight lower bin flows continuously through the feed pipes to fill the retort. Oxygen required for the gasification is supplied through the enriched air from the PSA system and the steam, supplementing steam that is generated in the jacket is supplied from Offsites. The gas leaves the gasifier from the top at about 300°C. The ash as and when formed is discharged below through the rotating grates.

The gas leaving the gasifier is sent to the water scrubber 312 to remove solid impurities. Scrubber circulating pumps 602 maintain water circulation in the scrubber. The condensate in the cooled gas from the scrubber is knocked out in the knock out drum 313 and stored in the gas holder 402.

The gas from the gasholder is compressed to 30 to 35 kg/cm² pressure in the gas compressor 754 prior to its further purification processes. About 10% of the carbon dioxide and 0.2% of sulphur in the raw cooled and compressed gas is removed in the Hi-pure Benfield section where the gas is washed with DEA & Potassium Carbonate solution in separate columns. The gas exit the section will be free of Carbon Dioxide and Sulphur.

The carbon monoxide in the gas exit the desulphurisation section is converted into carbon dioxide in the High (HTS) 316 and low (LTS) 317 temperature shift converters, which operate respectively at 370°C and 210°C. Additional steam at 14.4 TPH rate will be required for the shift reaction in both the HTS & LTS, which together bring down the Carbon Monoxide content from 43% to 0.2% and the Carbon Dioxide builds upto 30%.

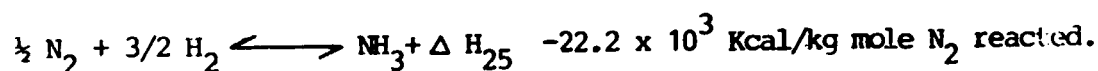
The Carbon Dioxide is removed in the decarbonation section, where 30% carbonate solution with 3% DEA and Vanadium Pentoxide constituting the Benfield solution is used. The carbon dioxide in the gas exit, this section will be 0.1%.

The carbon oxides in the gas are finally removed in the Methanator 320 where they are catalytically converted into Methane. Make up synthesis gas containing about 1.92% inerts, consisting of 0.89% Methane and 1.03% Argon, is then sent to the ammonia synthesis section.

3. AMMONIA SYNTHESIS SECTION3.1 Process Description [Drg. No.9484-237-11107]

The synthesis gas after the final purification in the Methanator is compressed in the synthesis gas compressor (755) to 450 kg/cm² pressure. The gas after oil separation in the oil filter (325) is then sent through the ejector (921) where part of its pressure energy is used up in boosting the recycle gas to the converter inlet pressure of 400 kg/cm².

The synthesis system used is as per Casale design. The synthesis make up gas mixed with the recycle gas in the ejector enters the converter (326). Inlet gas heating is by exchange of heat with the converter effluent which leaves the converter at about 260°C. The gas enters the catalyst bed at 370°C. The exothermic reaction takes place to produce ammonia



The converter effluent at 260°C is cooled with cooling water in the Ammonia condenser (530). The bulk of the Ammonia in the gas stream is condensed in a single stage in this condenser.

The condensed Ammonia is separated in the separator (327) and the recycle gas after a purge, is drawn to the recycle ejector, mixed with the synthesis make up gas and fed to the converter.

The ammonia in the separator is let down to 20 kg/cm² in the let down vessel (328). The dissolved gas released in the let down vessel joins the purge gas and goes to the ammonia recovery section for ammonia recovery. The gases scrubbed free of Ammonia is used as fuel.

The purge gas from synthesis loop and the flash gas from let-down vessel are sent to the scrubber (329) where the Ammonia is scrubbed with water. A part of the ammonia solution is fed to the middle section of the scrubber through a solution cooler (531) as circulating solution and the remaining solution is fed to the stripper (330).

In the Ammonia Stripper, the Ammonia solution is counter-currently heated with steam admitted at the bottom of the tower. The overhead gaseous product of the stripper is 99.5% NH_3 in vapour phase and this is condensed by a reflux condenser (534) and collected in a separator (331). A part of it is used as reflux and the remainder is sent to the let down vessel as recovered ammonia. The bottom product of the stripper is sent to the scrubber via solution exchanger (533) as absorbing fluid.

CHAPTER III

MINI AMMONIA PLANT

TECHNICAL DATA

List of Equipment with outline specifications

1. Natural Gas Steam Reforming Section (Ref: Drg. No. 9484-237-11120)

<u>Sl. No.</u>	<u>Item</u>	<u>Description.</u>	
1	301	Natural Gas KO Drum	1 No.
		Diameter	500 mm
		Height	1000 mm
		Material	Carbon Steel
		Internal	Demister pad made of AISI 304 material.
2	302	Desulphurisation Reactor	2 Nos.
		Diameter	1000 mm
		Height	3830 mm
		Material	Carbon Steel
		Zn O	1.7 M ³
3	303	Secondary Reformer	1 No.
		Shell diameter	1450 mm
		Total height	5000 mm
		Material	Carbon Steel Shell refractory lined inside. Water jacketed outside.
		Internal	Process Air Burner
		Catalyst	2.9 M ³

Sl No.	Item	Description	
4.	304	Steam Drum Common for all Waste Heat Boilers. Total Steam output 12 TPH at 40 kg/cm ² pressure.	1 No.
5.	305	HTS Reactor Diameter Height Material Catalyst	1 No. 1760 mm 1550 mm ½ M0 C Steel 3.53 M ³
6.	306	LTS Reactor Diameter Height Material Catalyst	1 No. 1870 mm 1825 mm Carbon Steel 4.76 M ³
7.	307	Knock Out Drum Diameter Height Material Internal	1 No. 780 1450 Carbon Steel with SS cladding inside. Demister pad AISI 304.
8.	308	CO ₂ Absorber Top Diameter Top Height Bottom Diameter Bottom Height Material Packing Internal	1 No. 750 mm 12000 mm 1500 mm 15000 mm Carbon Steel 40 mm 18-8 SS Slotted rings 21 M ³ Demister Pad of 18-8 SS

Sl. No.	Item	Description	
9.	309	Solution Regenerator	1 No.
		Diameter	2000 mm
		Height	36000 mm
		Material Top	18-8 Clad Carbon Steel
		Bottom	Carbon Steel
		Internal	Distributor
			Demister 18-8 SS
		Packing	50 mm 18-8 SS
			Slotted rings 31 M ³
10.	310	CO ₂ Knock Out Drum	1 No.
		Diameter	800 mm
		Height	1500 mm
		Material	18-8 SS Clad Carbon Steel
		Internal	18-8 SS Demister pad
11.	311	Methanator	1 No.
		Diameter	1550 mm
		Height	1600 mm
		Material	½ MO C Steel
		Catalyst	3.04 M ³
12.	401	Stack	1 No.
		Diameter	750 mm
		Height	20,000 mm
13.	501	Combustion Air Heater	1 No.
		Heat Duty	1.31 MM Kcal/Hr.
14.	502	Reformed Gas Boiler	1 No.
		Heat Duty	3.3 MM Kcal/Hr.
15.	503	Converted Gas Boiler	1 No.
		Heat Duty	0.613 MM Kcal/hr.
16.	504	Converted Gas Exchanger	1 No.
		Heat Duty	0.56 MM Kcal/Hr.

Sl. No.	Item	Description	
17.	505	BFW Heater No.3 Heat Duty	1 No. 0.068 MM Kcal/hr.
18.	506	BFW Heater No.2 Heat Duty	1 No. 0.21 MM Kcal/hr.
19.	507	BFW Heater No.1 Heat Duty	1 No. 0.8 MM Kcal/hr.
20.	508	Solution Reboiler Heat Duty	1 No. 3.17 MM Kcal/hr.
21.	509	Solution Exchanger Heat Duty	1 No. 3.78 MM Kcal/hr.
22.	510	Solution Cooler Heat Duty	1 No. 3 MM Kcal/hr.
23.	511	CO ₂ Gas Cooler Heat Duty	1 No. 1.38 MM Kcal/hr.
24.	512	Methanation Exchanger Heat Duty	1 No. 0.88 MM Kcal/hr.
25.	513	Make up Gas ^C ooler Heat Duty	1 No. 0.165 MM Kcal/hr.
26.	581	Natural Gas Preheater Heat Release No. of Burners Natural draft	1 No. 0.6 MM Kcal/hr. 2 Upshot type
27.	582	Primary Reformer Process Duty Type Heat Release	1 No. 6.1 MM Kcal/hr. Rectangular box type, refractory lined, down-fired. 13.5 MM Kcal/hr.

Sl. No.	Item	Description	
		No. of Burner	40
		No. of Tubes	64
		Inside Diameter of Tube	97 mm
		Fired length of Tube	10,000 mm
28.	601 AB	Solution Pump	2 Nos.
		Capacity	150 M ³ /Hr.
		Type	Centrifugal electric power driven
		Discharge Pressure	30 Kg/cm ² g.
29.	751	Process Air Compressor	1 No.
		Type	Reciprocating electric power driven
		Capacity	4250 NM ³ /hr.
		Discharge Pressure	40 kg/cm ² g
30.	752	Combustion Air Blower	1 No.
		Type	Centrifugal, Electric power driven.
		Capacity	13,000 NM ³ /hr.
		Discharge Pressure	550 mm WG.
31.	753	Induced Draft Fan	1 No.
		Type	Centrifugal, electric power driven.
		Capacity	18,000 NM ³ /hr.
		Suction Pressure	(-) 550 mm WG.

2. COAL GASIFICATION SECTION (Ref.Org.No.9484-237-11121)

Sl. No.	Item	Description	
1.	312	Water Scrubber	3 Nos.
		Diameter	3000 mm
		Height	7500 mm
		Material	Carbon Steel
		Packing	27 M ³ of 50 mm SWRR
2.	313	Gas K O Drum	1 No.
		Diameter	1500 mm
		Height	3000 mm
		Material	Carbon Steel
		Internal	18-8 SS Demister Pad
3.	314	Sulphur Absorber	1 No.
		Diameter	750 mm
		Height	10,000 mm
		Packing	40 mm 18-8 SS Slotted rings
74.	315	Sulphur Stripper	1 No.
		Diameter	650 mm
		Height	12,000 mm
		Packing	40 mm 18-8 SS Slotted rings
5.	316	Decarbonating Tower	1 No.
		Diameter	900 mm
		Height	20,000 mm
		Packing	40 mm 18-8 SS Slotted rings
6.	317	Carbonate Regeneration Tower	1 No.
		Diameter	1000 mm
		Height	15,000 mm
		Packing	40 mm 18-8 SS Slotted rings

Sl. No.	Item	Description	
7.	318	HTS Reactor	1 No.
		Diameter	1600 mm
		Height	8000 mm
		Material	$\frac{1}{2}$ MO C Steel
		Catalyst	9.5 M ³
8.	319	LTS Reactor	1 No.
		Diameter	2000 mm
		Height	2500 mm
		Material	Carbon Steel
		Catalyst	4.75 M ³
9.	320	CO ₂ Absorber	1 No.
		Top Diameter	750 mm
		Top Height	8000 mm
		Bottom Diameter	1500 mm
		Bottom Height	10,000 mm
		Packing	40 mm 18-8 SS Slotted rings
10.	321	CO ₂ Stripper	1 No.
		Diameter	2000 mm
		Height	20,000 mm
		Packing	50 mm 18-8 SS Slotted rings
11.	322	Methanator	1 No.
		Diameter	1550 mm
		Height	1600 mm
		Material	$\frac{1}{2}$ MO C Steel
		Catalyst	3.1 M ³

Sl. No.	Item	Description		
12.	402	Gas Holder Capacity Material	5000 M ³ Carbon Steel	1 No.
13.	514	Gas Cooler Heat Duty	1.25 MM Kcal/hr.	1 No.
14.	515	Solution Reboiler Heat Duty	1.4 MM Kcal/hr.	1 No.
15.	516	Reboiler 1 Heat Duty	1.33 MM Kcal/hr.	1 No.
16.	517	Converted Gas Exchanger 1 Heat Duty	1.024 MM Kcal/hr.	1 No.
17.	518	Converted Gas Exchanger 2 Heat Duty	1.008 MM Kcal/hr.	1 No.
18.	519	Converted Gas Exchanger 3 Heat Duty	1.02 MM Kcal/hr.	1 No.
19.	520	Reboiler 2 Heat Duty	3.2 MM Kcal/hr.	1 No.
20.	521	Methanator Exchanger Heat Duty	0.33 MM Kcal/hr.	1 No.
21.	522	Make-up Gas Cooler Heat Duty	0.65 MM Kcal/hr.	1 No.
22.	523	Acid Gas Cooler Heat Duty	3.2 MM Kcal/hr.	1 No.

Sl. No.	Item	Description	
23.	583	Coal Gasifier Fixed Bed Type Size 3000 mm I.D. Capacity 6000 NM ³ /hr. of gas containing CO + H ₂ 65 % by Vol minimum.	3 Nos.
24.	602	Scrubber Water Pump Capacity Discharge Pressure Type	6 Nos. 110 M ³ /hr. 2 kg/cm ² g. Centrifugal, Electric power driven
25.	754	Gas Compressor Capacity Suction Pressure Discharge Pressure Type	1 No. 13,500 NM ³ /hr. 400 MM Hg. 40 kg/cm ² g. Reciprocating, Electric power driven.
26.	851	PSA Unit Capacity Duty	1 No. 6500 NM ³ /hr. of enriched air. One complete module to produce oxygen enriched air containing 46 % V/V O ₂ minimum, with air compressor, Molecular sieves, valves, instruments interconnecting piping etc.

3. AMMONIA SYNTHESIS SECTION (Ref: Dwg.No.9484-237-11107)

Sl. No.	Item	Description	
1.	325	Make-up Gas Oil Separator	1 No.
		Diameter	350 mm
		Height	2000 mm
		Internal	filter elements.
2.	326	Ammonia Converter	1 No.
		Operating Pressure	350 - 400 kg/cm ² g
		Internals	Heat Exchangers, Electric Heater, Cooling Tubes & Catalyst. 3.2 M ³
3.	327	Ammonia Separator	1 No.
		Diameter	500 mm
		Height	4000 mm
4.	328	Ammonia let down Vessel	1 No.
		Diameter	1250 mm
		Height	3800 mm
5.	329	Scrubber	1 No.
		Diameter	350 mm
		Height	16500 mm
		Packing	25 mm SWRR 1.2 M ³
6.	330	Stripper	1 No.
		Diameter	750 mm
		Height	20000 mm
		Packing	25 mm SWRR 4.5 M ³
7.	331	Separator	1 No.
		Diameter	650 mm
		Height	1200 mm

Sl. No.	Item	Description	
8.	530	Ammonia Condenser	1 No.
		Heat Duty	1.53 MM Kcal/hr.
9.	532	Water Cooler	1 No.
		Heat Duty	0.08 MM Kcal/hr.
10.	533	Solution Heater	1 No.
		Heat Duty	0.135 MM Kcal/hr.
11.	534	Reflux Condenser	1 No.
		Heat Duty	0.16 MM Kcal/hr.
12.	605 AB	Scrubber Circulating Pump	2 Nos.
		Capacity	5 M ³ /hr.
		Medium Handled	Aqueous Ammonia
		Discharge Pressure	25 kg/cm ² g
13.	606 AB	Reflux Pump	2 Nos.
		Capacity	5 M ³ /hr.
		Medium	Liquid Ammonia
		Discharge Pressure	25 kg/cm ² g
14.	755	Synthesis Compressor	
		Capacity	7500 NM ³ /hr.
		Suction Pressure	24 kg/cm ²
		Discharge Pressure	450 kg/cm ²
		Reciprocating electric power driven. The unit to include instruments, interstage coolers, molecular sieve dryer, interconnecting pipes, valves etc.	
15.	921	Ejector	
		Duty	To boost recycle gas 45,200 NM ³ /hr. from 388 kg/cm ² to 400 kg/cm ² with make up gas 12,550 NM ³ /hr. at 450 kg/cm ² g.

CHAPTER-IV

100 TPD MINI-AMMONIA PLANT:

Capital and Operating Costs

Capital cost and operating cost for 100 MTPD Ammonia and matching end-product plant Calcium Ammonium Nitrate have been worked out and given below on the basis of Anthracite Coal and Natural Gas as feed-stocks for Ammonia production. In both cases, the cost of associated facilities have been added to the main plant costs. The plant site assumed is a typical Indian location (Udyogamandal). It is assumed that the Site is a developed one with infra-structural facilities like road/rail access, power and water supply upto battery-limits. The rates given at the end of the estimate list out the various assumptions underlying the estimate.

Table-1	:	100 MTPD Ammonia Plant - Summary of Capital Cost..
Table-2	:	-do- -do- - Cost of Production and Sensitivity Study.
Table-3	:	End-product Plant (CAN)- Summary of Capital Cost..
Table-4	:	-do- -do- (CAN)- Cost of Production and Sensitivity Study.

Table-1100 MTPD AMMONIA PLANTSummary of Capital Cost

Scope in this battery-limits:

- a) Coal based plant - Feedstock gasification, Benfield de-sulfurisation, CO conversion, Benfield CO₂ removal, Methanation and Ammonia synthesis, PSA for enriched oxygen supply.
- b) Natural Gas based plant - Steam reforming, CO conversion, CO₂ removal, methanation and Ammonia synthesis.

	\$ Million	
	With Coal Gasification	With Natural Gas Steam Reform- ing
1. Equipment & Machinery (Battery-limits)	18.112	13.820
2. Spares	1.811	1.382
3. Total F.O.B./F.O.R. Cost	<u>19.923</u>	<u>15.202</u>
4. Marine & Inland transport	2.988	2.280
5. Licence, Engg. & Management Expenses	5.434	4.146
6. Local Erection cost	1.358	1.546
7. Civil Construction Cost	<u>0.430</u>	<u>0.269</u>
8. Sub total 3 to 8	30.133	23.443
9. Preliminary & Pre-production Expenses	<u>1.507</u>	<u>1.172</u>
10. Erected Battery-limits plant cost	31.640	24.615
11. Erected cost of off-sites & services	<u>6.328</u>	<u>3.938</u>
12. Sub total (10+11)	37.968	28.553
13. Contingencies	<u>3.797</u>	<u>2.855</u>
14. Total Fixed Capital	41.765	31.408
15. Working Capital	<u>0.325</u>	<u>0.210</u>
16. Total Investment	42.090	31.618
	*****	*****

100 MTPD AMMONIA PLANTTable-2Cost of production and Sensitivity Study

Base Case - 90% Production level

	<u>Coal Gasification</u>	<u>N.G. steam Reforming</u>
1. Production '000 TPA	28.8	28.8
2. Cost of production in \$ million/year		
2.1 Feed Stock	1.107	3.272
2.2 Utilities	2.809	0.896
2.3 Maintenance, Labour & overheads	2.053	1.542
2.4 Depreciation	4.172	3.141
	<hr/>	<hr/>
Total Cost	10.141	8.851
Cost / M.T. \$	352.12	307.33

3. Sensitivity analysis

	<u>Cost of Prodn. \$/M.T.</u>	<u>Net Change \$/M.T.</u>	<u>Cost of Prodn. \$/M.T.</u>	<u>Net change \$/M.T.</u>
3.1 At 80% level of production	379	27	328	21
3.2 At 100% level of prodn	331	- 21	291	- 16
3.3 With +10% in feed- stock price	356	4	319	12
3.4 With +10% in capital cost	371	19	322	15
3.5 With +10% return on investment	498	146	417	110

Table-3END-PRODUCT PLANT - CALCIUM AMMONIUM NITRATESummary of Capital Cost

Scope inside Battery Limit:

- 1) Nitric acid plant - 180 MTPD
 2) Amm. Nitrate /CAN Plant - 220 MTPD (AN) - 280 MTPD (CAN)
 3) Lime Grinding Unit,
 Product Storage & Handling

\$, Million

1. Erected Battery Limit Plant cost	19.700
2. Off-sites & Service facilities	<u>3.940</u>
3. Sub total (1) and (2)	23.640
4. Contingencies	<u>2.364</u>
5. Total Fixed Capital	26.004
6. Working Capital	<u>0.756</u>
7. Total Investment	<u><u>26.760</u></u>

Table-4END PRODUCT PLANT (CAN)Cost of production and Sensitivity studyBase Case - Production from NH₃ availability at 90% level

	Coal Gasifi- cation	N.G. Steam Reforming		
1. Production				
Ammonia available '000 TPA	28.8	28.8		
CAN Production '000 TPA	87.3	87.3		
2. Cost of production				
			\$ Million/year	
2.1 Ammonia	10.141	8.851		
2.2 Limestone	0.170	0.170		
2.3 Utilities	0.357	0.357		
2.4 Maintenance, Labour & overheads	1.020	1.020		
2.5 Depreciation	2.600	2.600		
Total cost	14.288	12.998		
Cost/Te	\$ 163.67	148.89		
	Say (164)	(149)		
3. <u>Sensitivity analysis</u>				
	Cost of prodn.	Change \$/M.T.	Cost of prodn.	Change \$/M.T.
3.1 At 80% level Ammonia availability	178	14	161	12
3.2 At 100% " "	152	-12	139	- 10
3.3 With 10% increase in Capital Cost	174	10	157	8
3.4 With 10% increase in Ammonia cost	175	11	159	10
3.5 With 10% return on investment	243	79	216	67

100 MTPD MINI AMMONIA PLANTAssumptions on Cost Estimate

1. Capital Cost
 - 1.1 Equipment cost and materials are assumed at Indian Prices (Jan. 83)
 - 1.2 Spares @ 10% of equipment cost have been included.
 - 1.3 Transport costs include 15% towards transportation and associated expenses (marine and inland).
 - 1.4 Site erection and civil works costs are based on prevailing rates at Udyogamandal (India).
 - 1.5 No duties or taxes are included in the estimates.
 - 1.6 Off-site facilities included are Power Receiving and Distribution within plant area. Water Treatment, Steam Generation, Cooling towers, Raw-material receiving storage and handling, Effluent Disposal, Workshops, Laboratory, General Stores and Administrative buildings.
 - 1.7 It is assumed that developed land with infrastructural facilities is available and hence no cost has been provided towards land and development, utility supply upto battery limit, rail road access and township.
 - 1.8 Contingency has been provided @ 10% of fixed capital.
 - 1.9 As method of financing is not decided, interest during construction has not been considered in the capital costs.
 - 1.10 The effect of inflation during construction has also not been considered. The impact of a 10% increase in cost has been shown in the sensitivity study.
 - 1.11 In working capital, the following are included:
 - i) 30 days' stock of coal feed for Ammonia (no storage of N.G.)
 - ii) 5 days' stock of Ammonia
 - iii) 15 days' stock of end-product fertilizer CAN
2. Cost of production
 - 2.1 Plant availability:

Number of stream days considered for this study is 320. Base case is taken at 90% production level of Ammonia. End-product capacities are fixed on the basis of Ammonia availability.
 - 2.2 For this study, the following feed-stock prices are assumed:

Coal	:	\$ 30/Tonne
Natural Gas	:	\$ 16 for 10 ⁶ K.Cal.

- 2.3 Utility prices assumed are \$0.05/M³ for water and \$0.05/kWh for power commensurate with the prevailing rates in Udyogamanda (India).
- 2.4 Maintenance is taken @ 3% of erected plant cost.
- 2.5 Average salary is taken at \$2000.- per year per person. Overheads have been added at 100% of salaries.
- 2.6 Straight-line depreciation @ 10% is taken for computing cost of production.
- 2.7 No interest on capital is assumed.
- 2.8 Sensitivity study gives the cost of production with 10% return on total investment.
- 2.9 Sensitivity study also gives the impact of 10% change in production level, 10% increase in capital cost and 10% increase in feed-stock cost.
-

ANTHRACITE COAL 5330 KG/HR

STEAM 5300 KG/HR

COAL GASIFIER
583

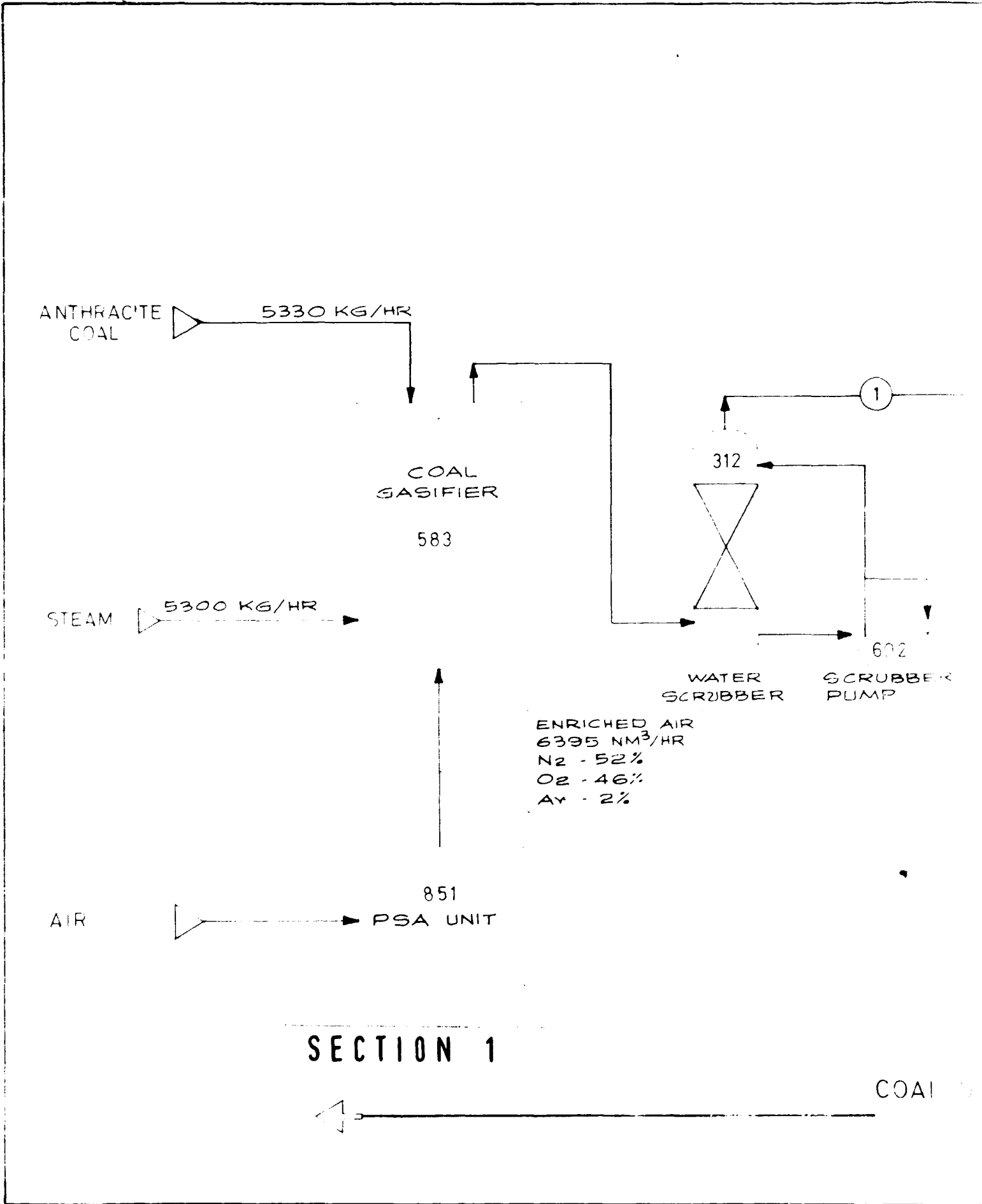
AIR 851 PSA UNIT

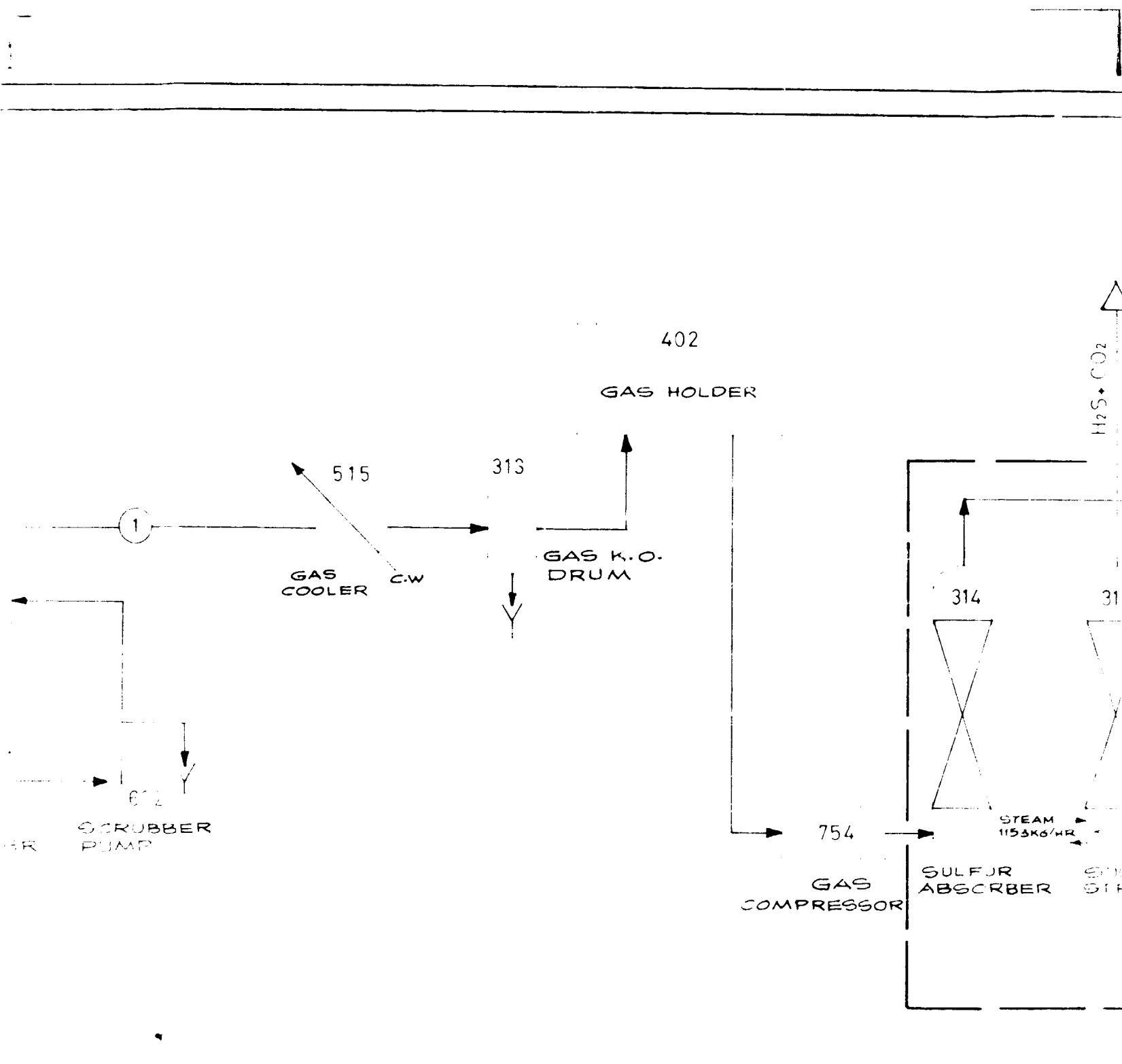
ENRICHED AIR
6395 NM³/HR
N₂ - 52%
O₂ - 46%
Ar - 2%

WATER SCRUBBER
SCRUBBER PUMP

SECTION 1

COAL

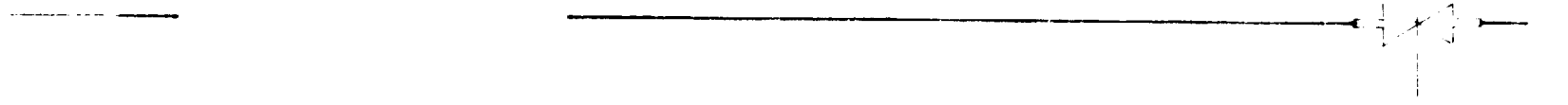


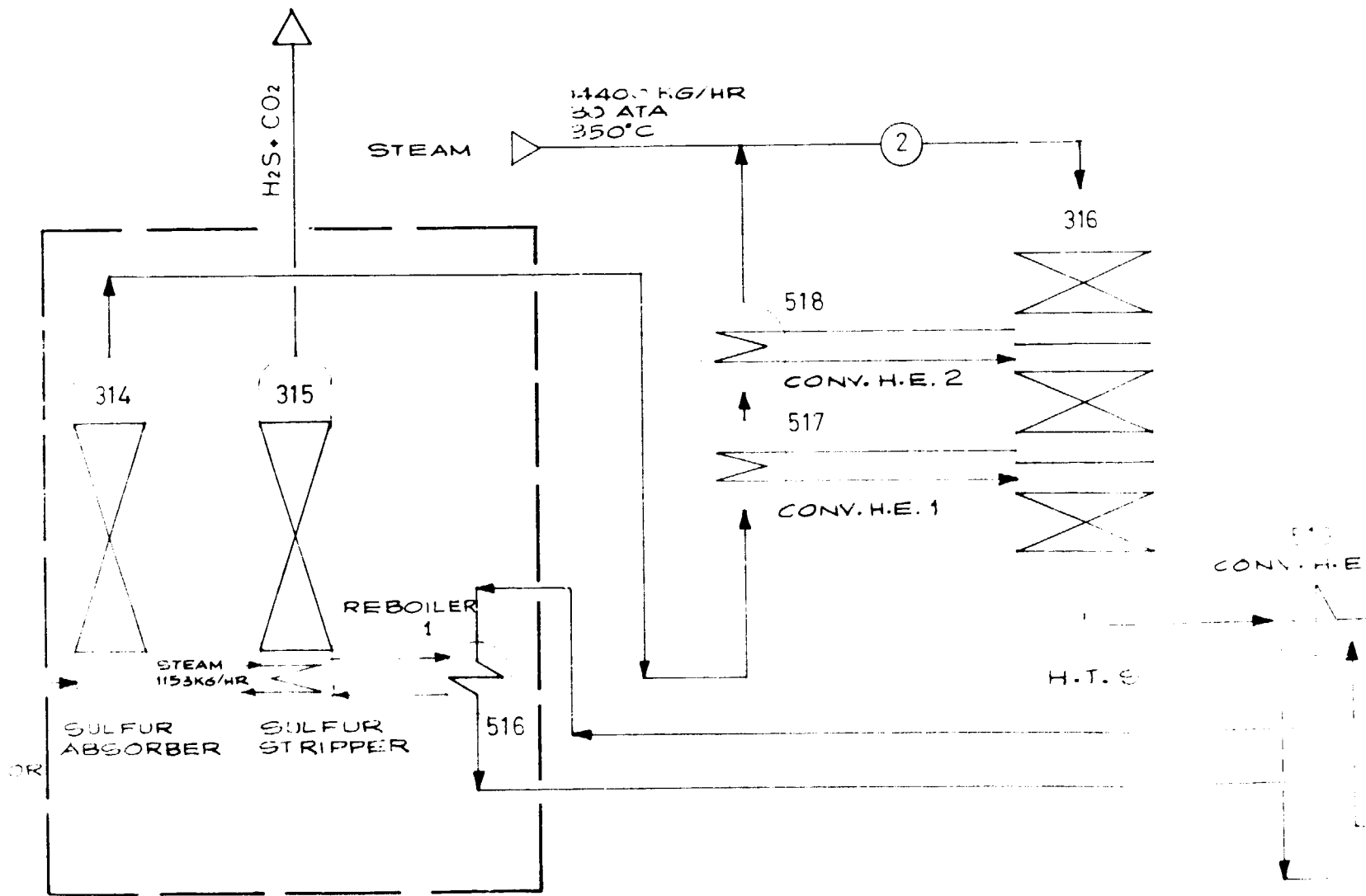


SECTION 2

COAL GASIFICATION

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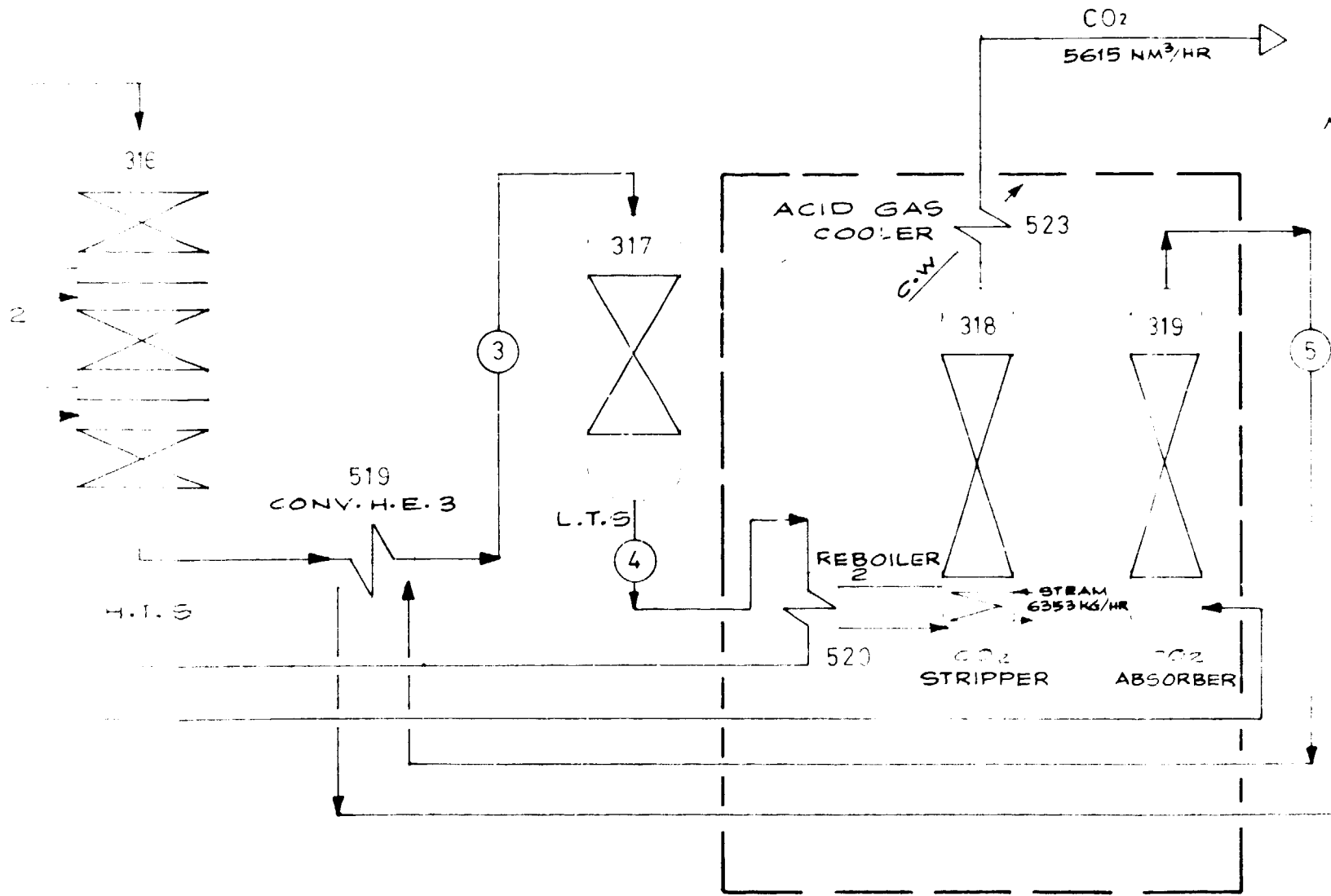




SECTION 3

DESULFURISATION

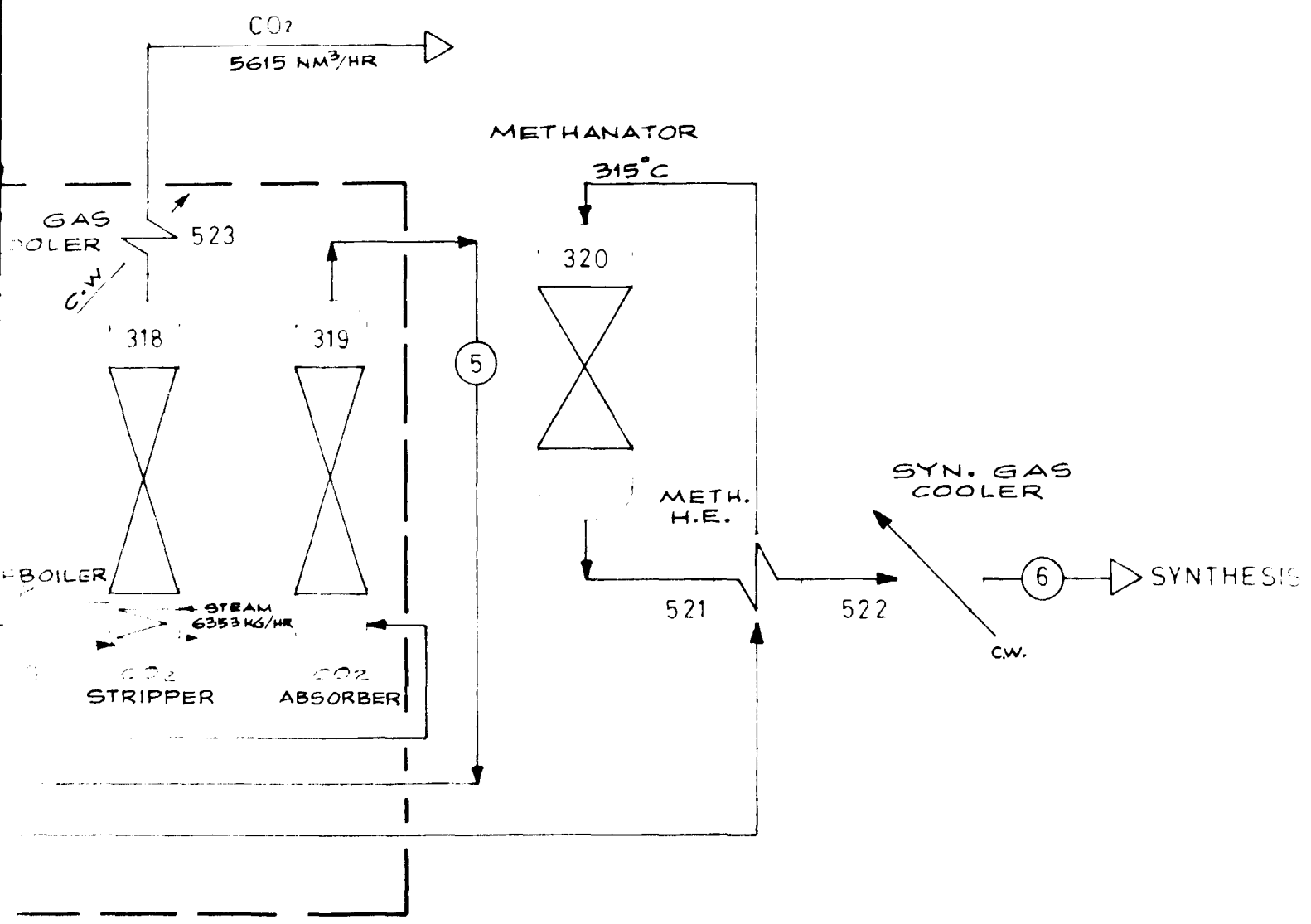
SHIFT CONVERSION



SECTION 4

SHIFT CONVERSION

DECARBONATION



SECTION 5

DECARBONATION

METHANATION



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STREAM	1	2	3	4	5	6
CO %	39.37	43.94	3.00	0.20	0.29	
H ₂ %	26.38	29.45	49.60	50.98	73.18	73.12
CO ₂ %	10.17		28.50	30.44	0.1	
N ₂ %	21.92	24.47	17.55	17.07	24.54	24.96
O ₂ %	0.24	0.27				
CH ₄ %	0.78	0.87	0.63	0.61	0.88	0.89
Ar %	0.9	1.0	0.72	0.7	1.01	1.03
H ₂ S %	0.2					
COS %	0.04					
FLOW NM ³ /HR	14300	12800	17865	18365	12750	12535
TEMP °C	65	370	210	230	60	40
PRES. KG/CM ² G	0.5	31	28.6	28.3	27.4	27
S/G	0.21	1.4	0.71	0.66	0.007	0.003

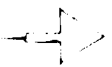
GAS
ER

⑥ → SYNTHESIS

L.W.

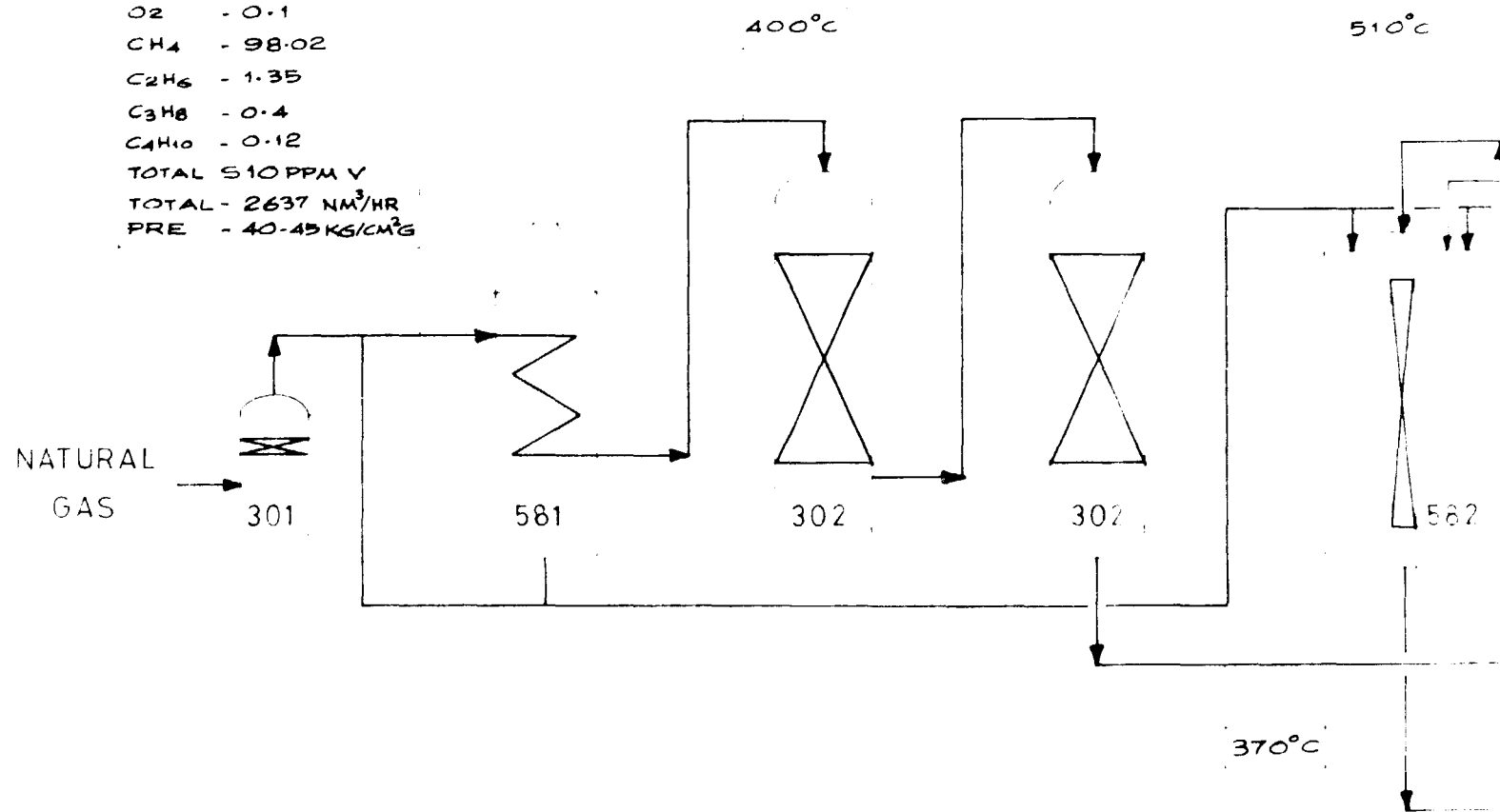
SECTION 6

FACT ENGINEERING & DESIGN ORGANISATION			
THE FERTILISERS AND CHEMICALS TRAVANCORE LIMITED			
TIRUYOGAMANDAL		KERALA	
DESIGNED		TITLE	
DRAWN	<i>[Signature]</i>	PROCESS FLOW SCHEME	
CHECKED	<i>[Signature]</i>	COAL GASIFICATION SECTION	
APPROVED		MINI AMMONIA PLANTS FOR UNIDO	
DATE	30.3.1983	DRG.No. 9484-237-1121	R
SCALE	NTS		



← SATURATED STEAM
40 KG/CM²

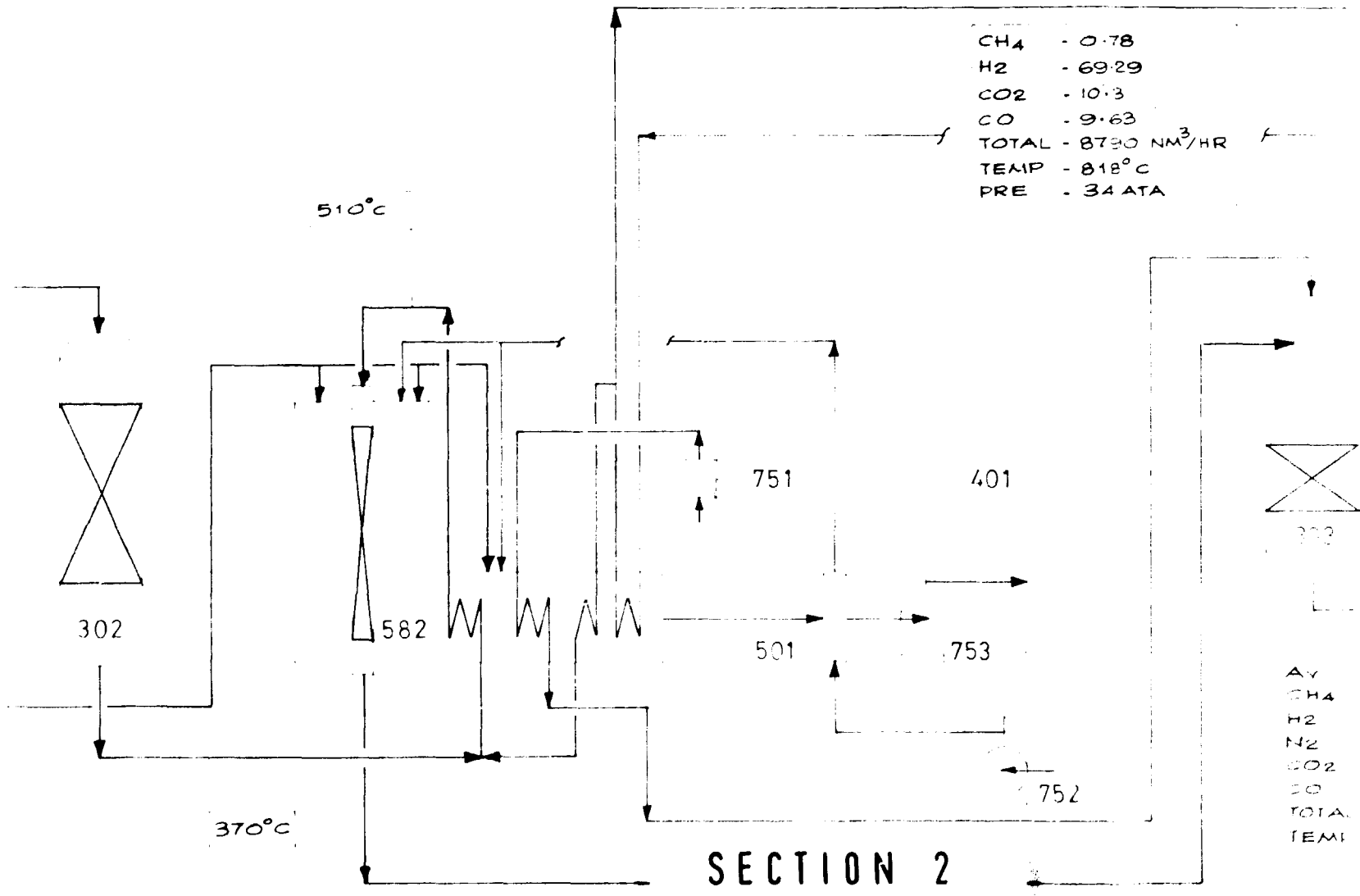
N₂ - 0.01
O₂ - 0.1
CH₄ - 98.02
C₂H₆ - 1.35
C₃H₈ - 0.4
C₄H₁₀ - 0.12
TOTAL SIO PPM V
TOTAL - 2637 NM³/HR
PRE - 40-45 KG/CM²



SECTION 1

DESULFURISATION

← SATURATED STEAM FOR PLANT USE
40 KG/CM²G



CH₄ - 0.78
H₂ - 69.29
CO₂ - 10.3
CO - 9.63
TOTAL - 8790 NM³/HR
TEMP - 818°C
PRE - 34 ATA

B.F.W. →

PRI. REFORMING

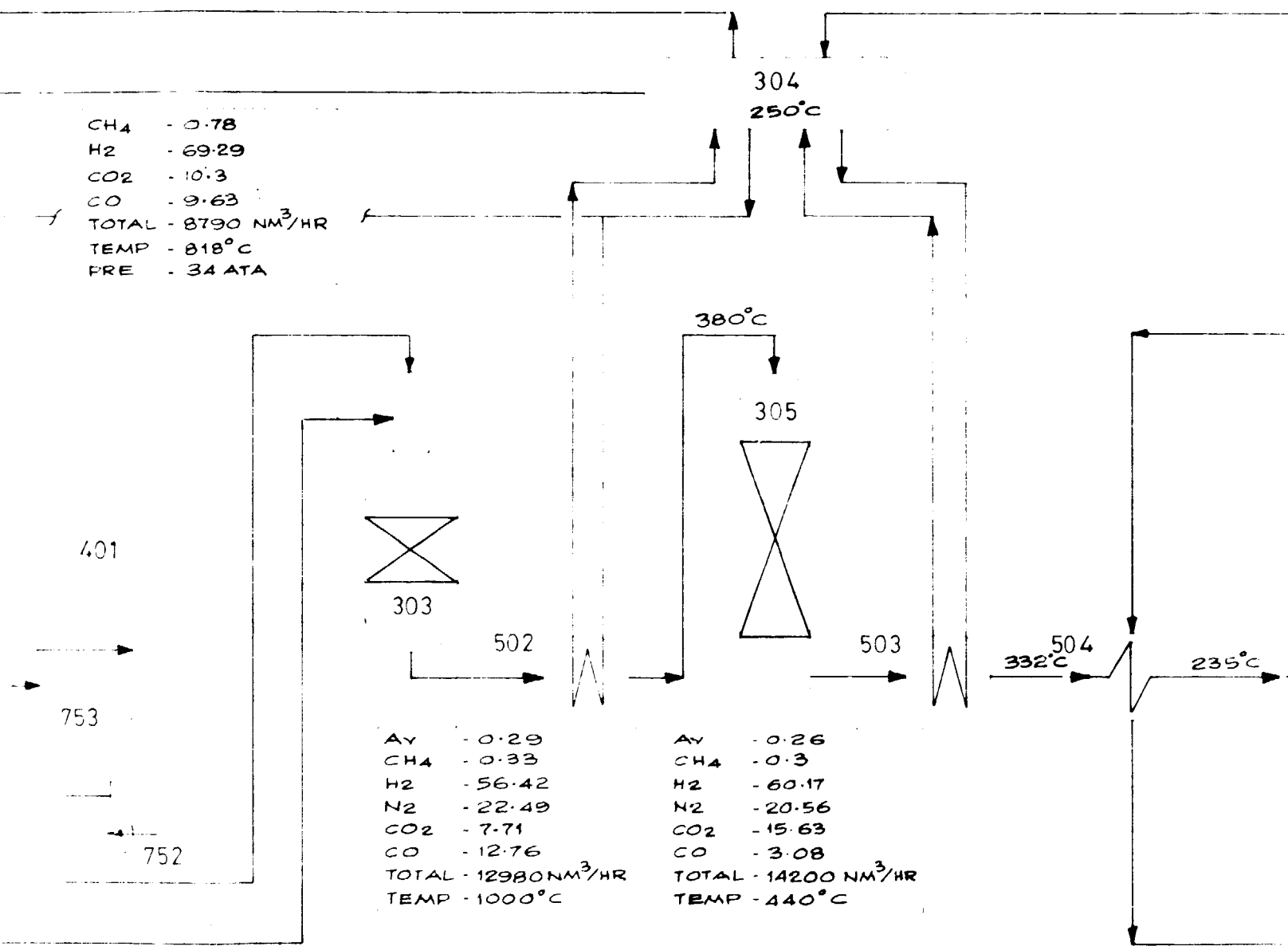
PROCESS STEAM
TEMP - 371°C
PRE - 40KG/CM²G
TOTAL - 7590KG/HR

PROCESS AIR
TEMP - 470°C
PRE - 40KG/CM²G
TOTAL - 3750NM³/HR

SEC. REFORMING



CH₄ - 0.78
 H₂ - 69.29
 CO₂ - 10.3
 CO - 9.63
 TOTAL - 8790 NM³/HR
 TEMP - 818°C
 PRE - 34 ATA



Ar - 0.29
 CH₄ - 0.33
 H₂ - 56.42
 N₂ - 22.49
 CO₂ - 7.71
 CO - 12.76
 TOTAL - 12980 NM³/HR
 TEMP - 1000°C

Ar - 0.26
 CH₄ - 0.3
 H₂ - 60.17
 N₂ - 20.56
 CO₂ - 15.63
 CO - 3.08
 TOTAL - 14200 NM³/HR
 TEMP - 440°C

PROCESS AIR
 TEMP - 470°C
 PRE - 40KG/CM²
 TOTAL - 3750 NM³/HR

SEC. REFORMING

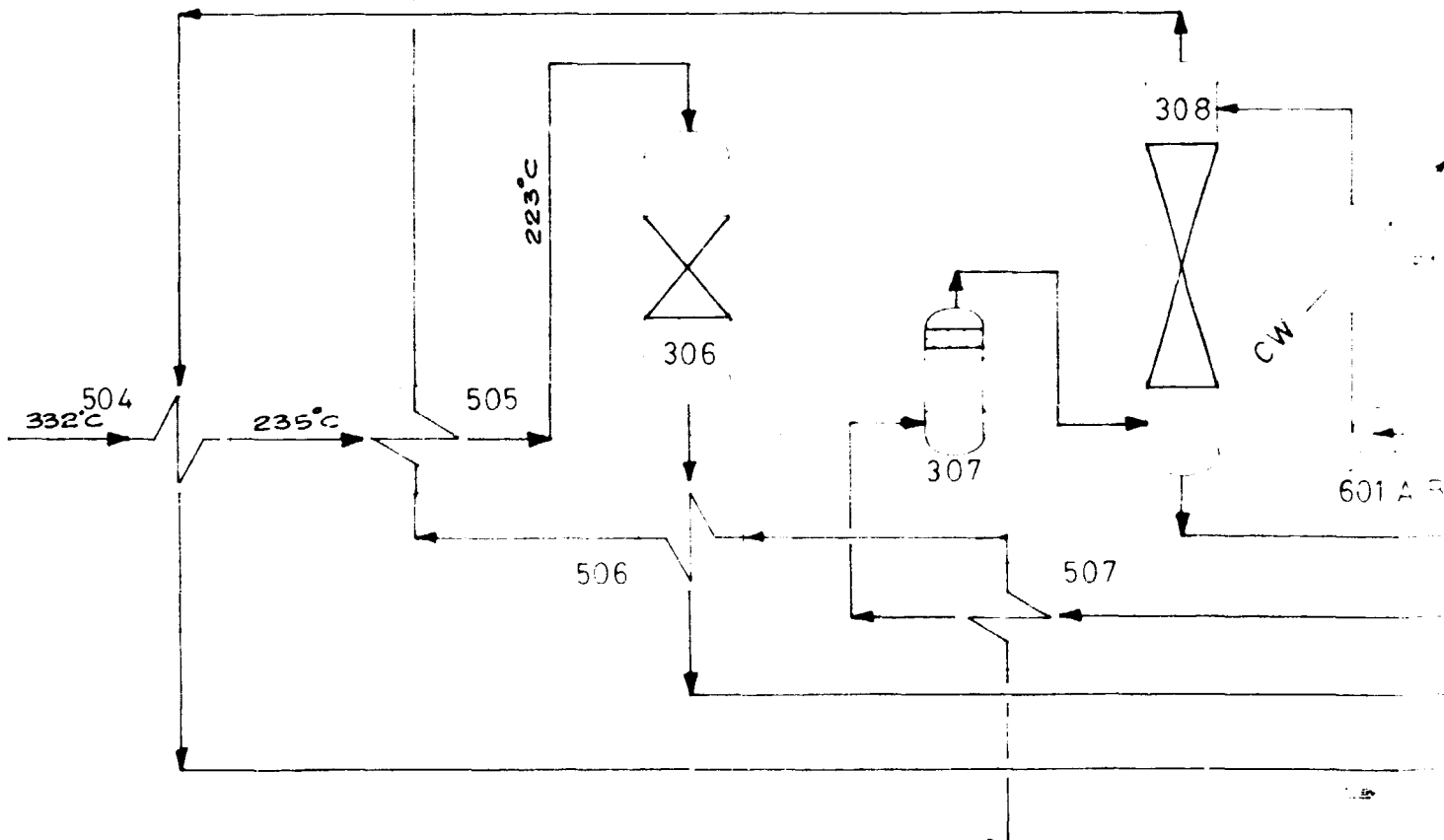
HTS

SECTION 3



Ar - 0.26
 CH₄ - 0.3
 H₂ - 61.24
 N₂ - 20
 CO₂ - 17.9
 CO - 0.3
 TOTAL - 14595 NM³/HR
 TEMP - 235°C

Ar - 0.31
 CH₄ - 0.35
 H₂ - 74.53
 N₂ - 24.34
 CO₂ - 0.1
 CO - 0.37
 TOTAL - 11990 NM³/HR
 TEMP - 50°C



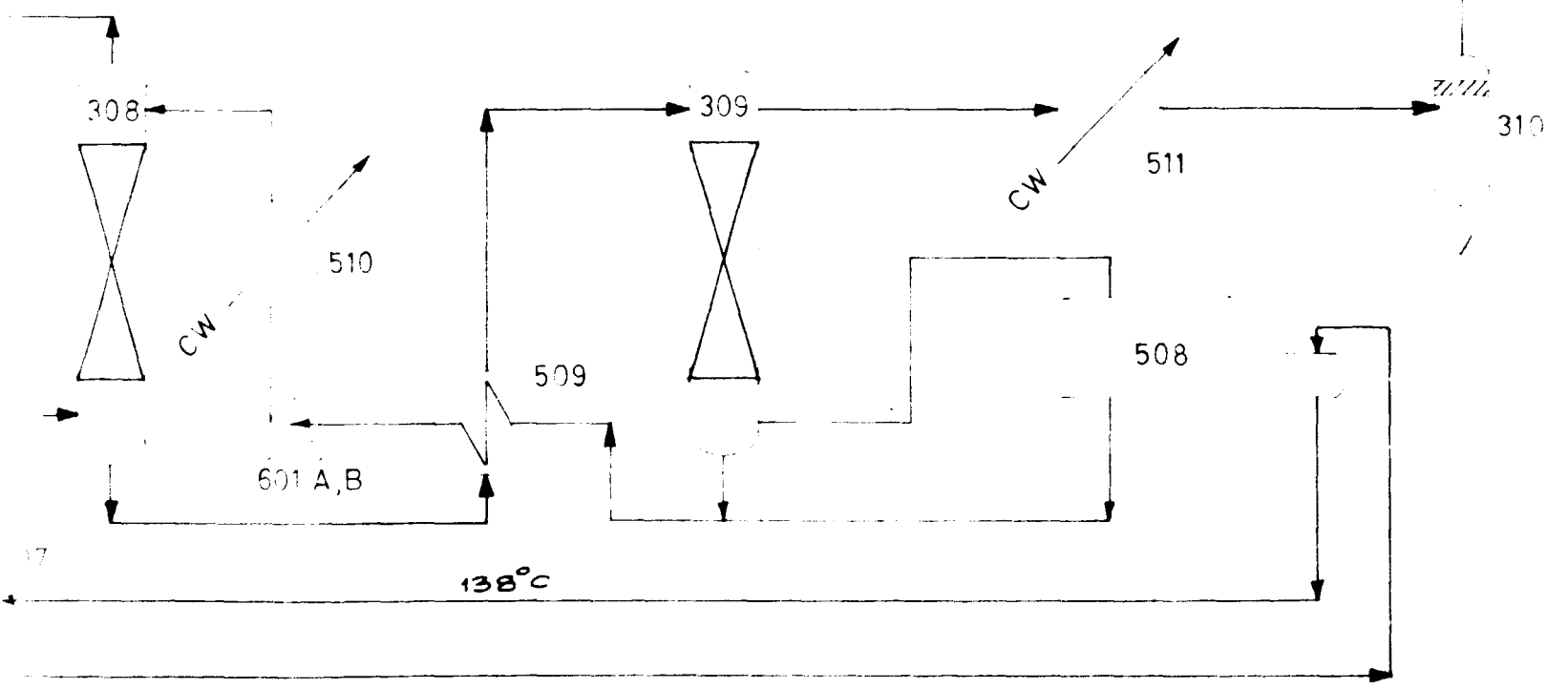
LTS

CO₂ REMOVAL

SECTION 4



51
 35
 53
 34
 37
 90 NM³/HR
 °C

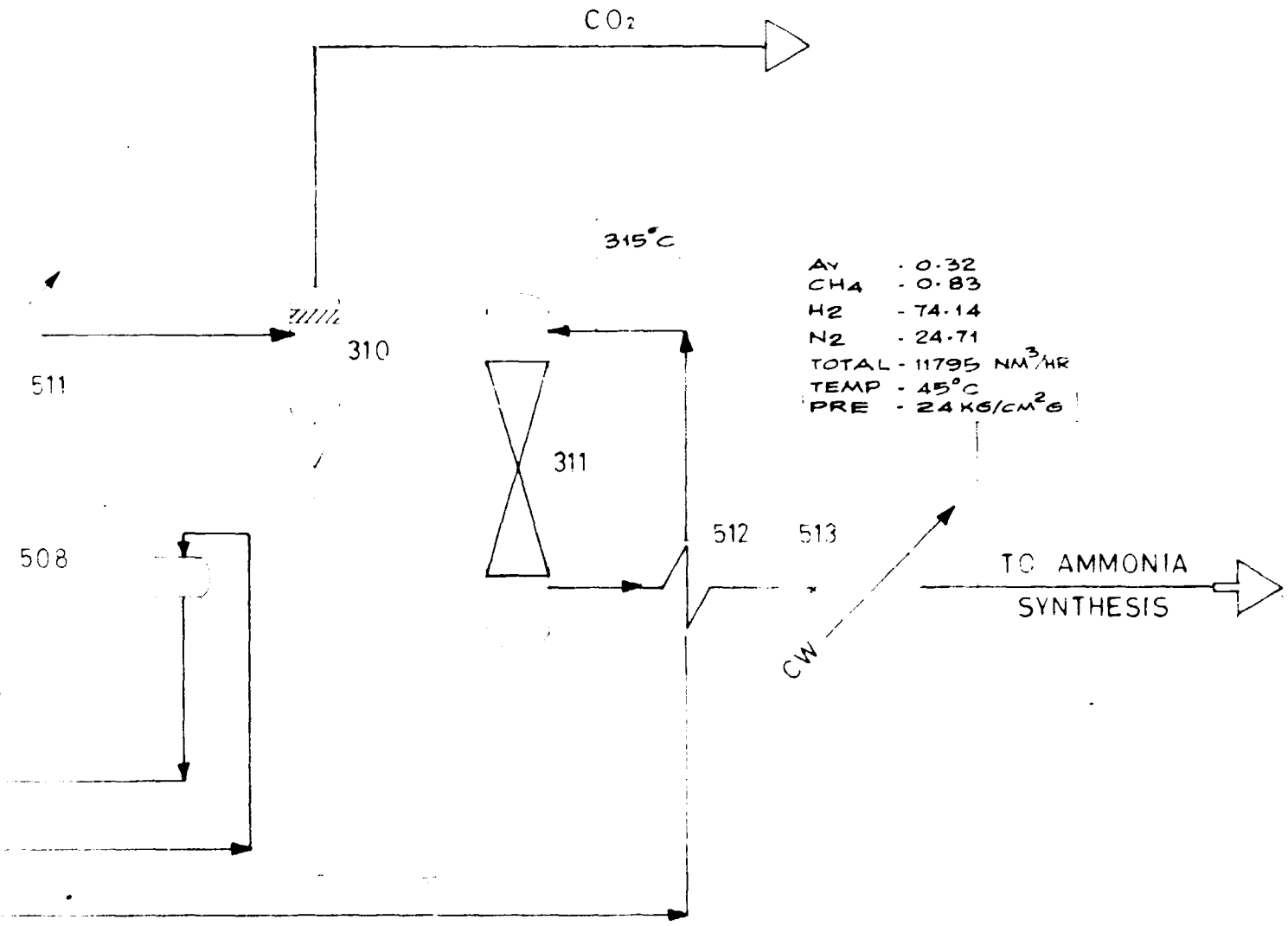


CO₂ REMOVAL

SECTION 5

ME





METHANATION
SECTION 6

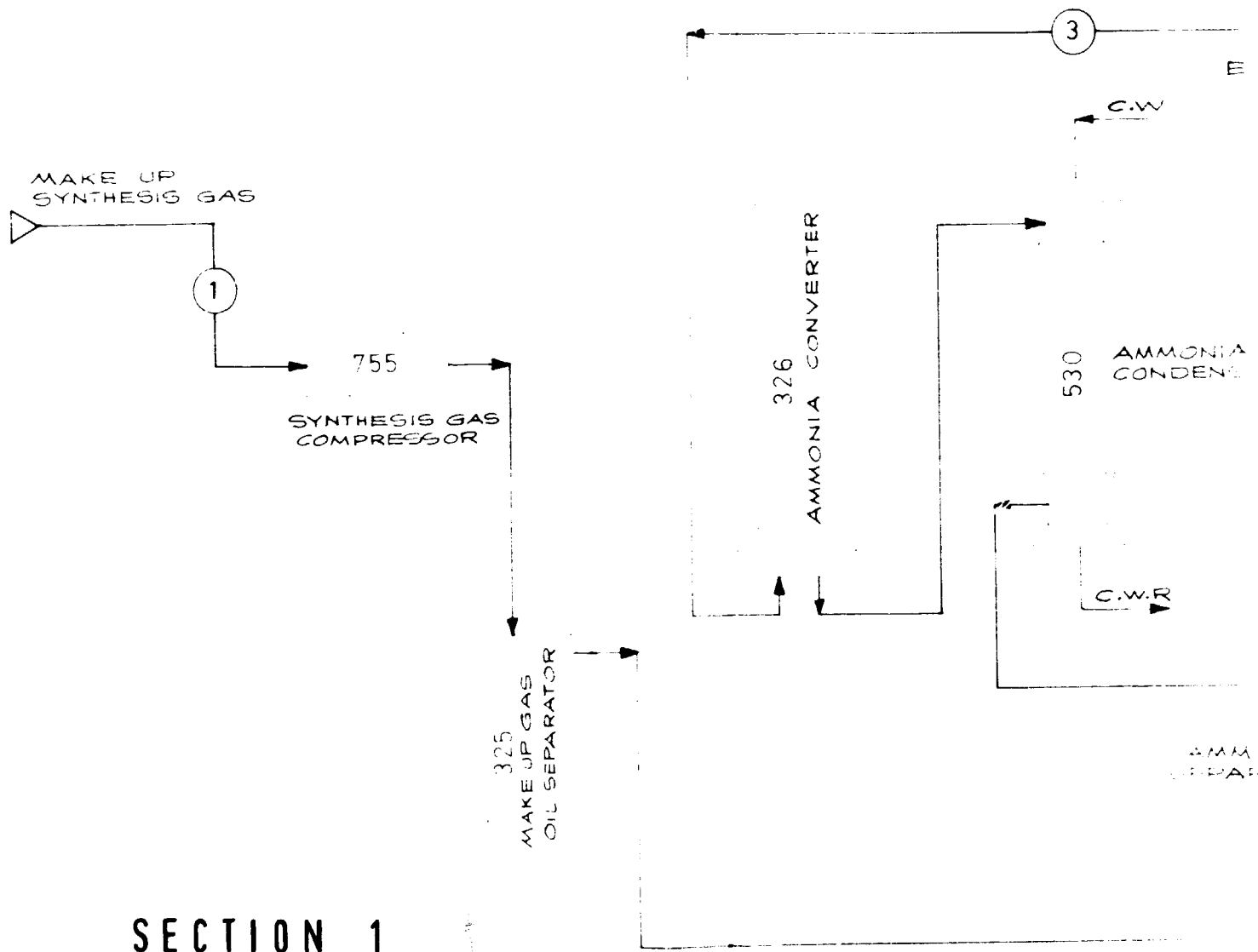
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ITEM No	DESCRIPTION
301	NATURAL GAS K.O.DRUM
302	DESULFURISATION VESSEL
303	SEC. REFORMER
304	STEAM DRUM
305	HTS
306	LTS
307	K.O.DRUM
308	CO ₂ ABSORBER
309	REGENERATOR
310	CO ₂ K.O.DRUM
311	METHANATOR
401	STACK
501	COMB. AIR HEATER
502	R.G. BOILER
503	CONVERTED GAS BOILER
504	CONVERTED GAS EXCHANGER
505	B.F.W. HEATER No. 3
506	B.F.W. HEATER No. 2
507	B.F.W. HEATER No. 1
508	SOLUTION REBOILER
509	SOLUTION EXCHANGER
510	SOLUTION COOLER
511	CO ₂ GAS COOLER
512	METH. EXCHANGER
513	MAKE-UP GAS COOLER
581	N.G. PRE HEATER
582	PRI. REFORMER
601A,B	SOLUTION PUMP
751	PROCESS AIR COMPRESSOR
752	F.D. FAN
753	J. D. FAN

AMMONIA
SYNTHESIS →

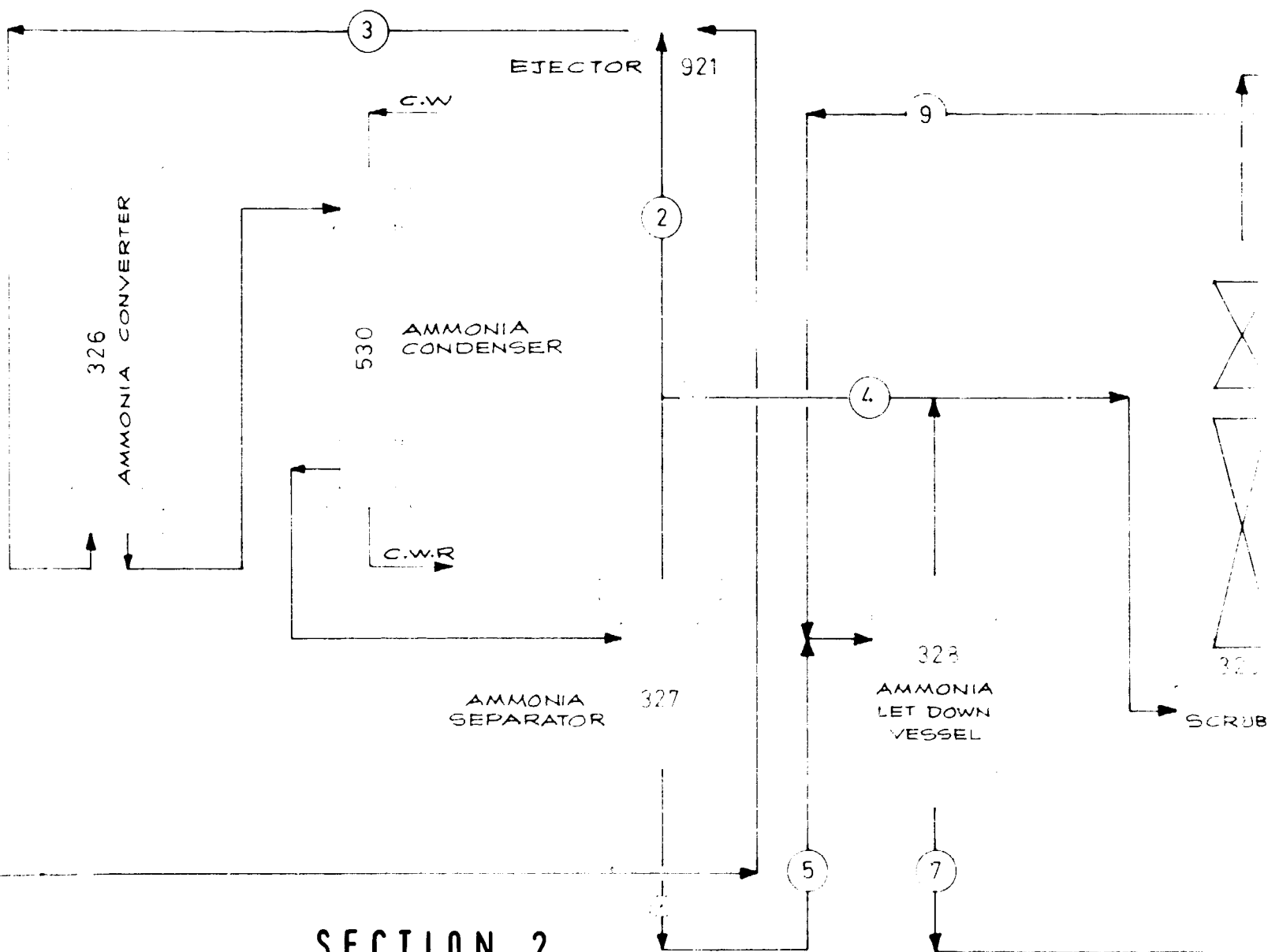
SECTION 7

FACT ENGINEERING & DESIGN ORGANISATION			
THE FERTILISERS AND CHEMICALS TRAVANCORE LIMITED			
UDYOGAMANDAL		KERALA	
DESIGNED		TITLE	
DRAWN	<i>Shanmughan</i>	PROCESS FLOW SCHEME	
CHECKED	<i>Krishnakant</i>	N.G. STEAM REFORMING SECTION	
APPROVED		MINI AMMONIA PLANTS FOR UNIDO	
DATE	10.6.1983	DRG.No. 9484-237-11120	
SCALE	NTS		RC



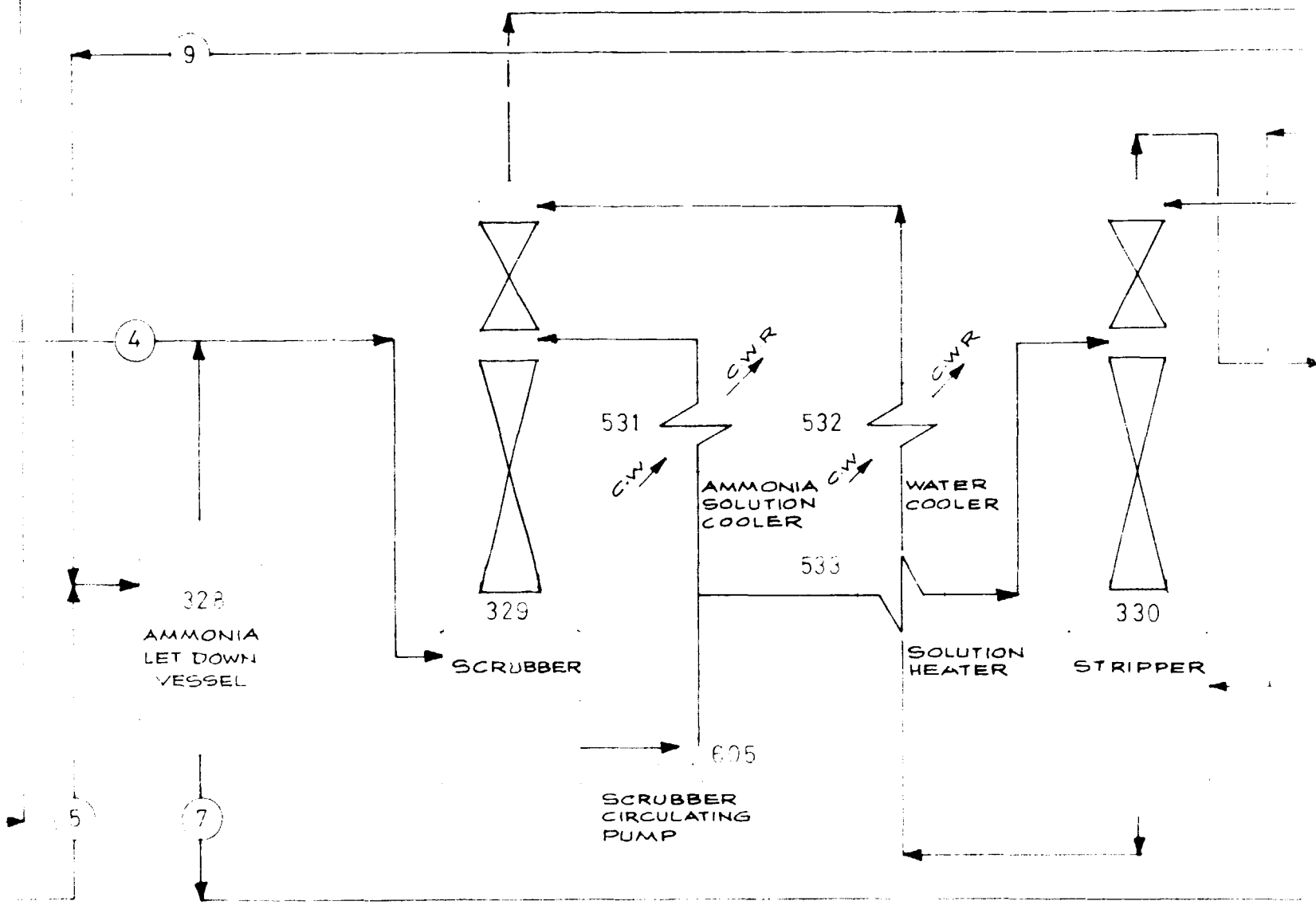
SECTION 1

STREAM	1	2	3	4	5	6	7
MEDIUM	GAS	GAS	GAS	GAS	LIQUID	GAS	LIQUID
H ₂ KMOL/HR	409.32	1189.71	1507.42	130.59	4.55	39.6	
N ₂ KMOL/HR	137.53	405.18	542.71	2.13	1.00	13.78	
NH ₃ KMOL/HR	0.00	90.80	90.80	2.72	43.5	0.0	45.7
CH ₄ KMOL/HR	4.98	112.12	117.10	1.16	1.62	4.98	
A KMOL/HR	5.76	129.75	135.51	1.88	1.88	5.76	
DRY TOTAL KMOL/HR	557.25	1926.65	2483.00	57.68	133.14	64.12	45.7
H ₂ O KG/HR	38.5	0.0	0.0	0.0	0.0	0.4	
PRESSURE KG/CM ² G	24.0	388.0	400.0	100.0	20.0	20.0	18.0
TEMPERATURE °C	40.0	40.0	40.0	400.0	26.0	40.0	26.0



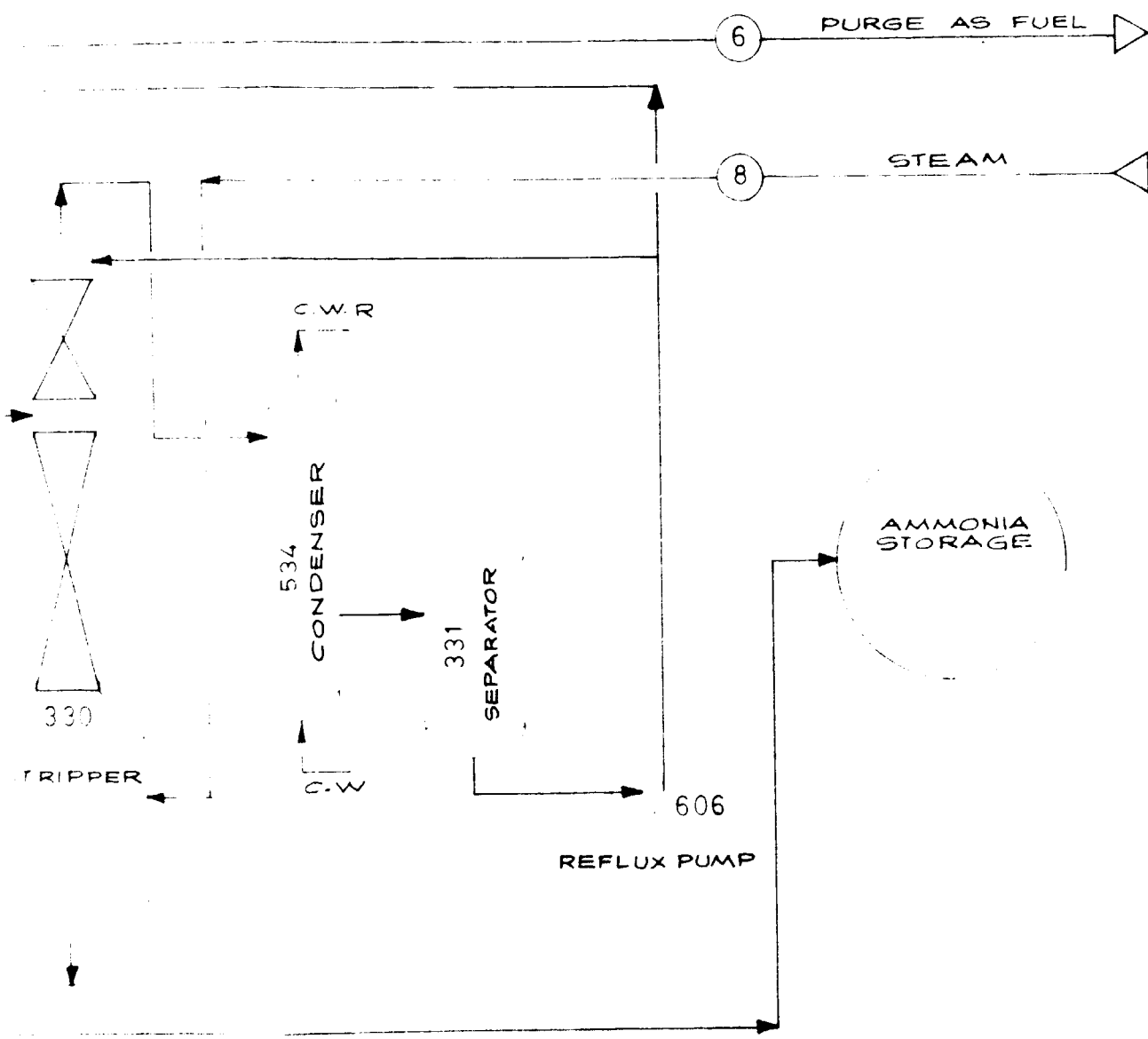
SECTION 2

	4	5	6	7	8	9
	KG	LIQUID	KG	LIQUID	STEAM	LIQUID
1	30.59	4.55	39.6			
2	2.13	1.00	13.78			
3	2.72	42.5	0.0	45.73		11.87
4	1.26	1.62	4.98			
5	1.28	1.88	5.76			
6	57.68	53.14	64.12	45.73		11.87
7	0.0	0.0	0.4		120.75	
8	0.0	20.0	20.0	18.00	35.00	20.0
9	40.00	26.0	40.0	26.00	243.00	38.0



SECTION 3

9
 5010
 11-87
 11-87
 20-0
 38-0



SECTION 4

FACT ENGINEERING & DESIGN ORGANISATION THE FERTILISERS AND CHEMICALS TRAVANCORE LIMITED UDYOGAMANDAL KERALA			
DESIGNED		TITLE	
DRAWN	<i>P. J. S. Nair</i>	PROCESS FLOW SCHEME	
CHECKED	<i>W. S. S. Nair</i>	AMMONIA SYNTHESIS SECTION	
APPROVED		MINI AMMONIA PLANTS FOR UNIDO	
DATE	3 3 1983	DRG.No. 9484-237-11107	R
SCALE	NTS		

