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ENERGY CONSUMPTION IN AMMONIA PRODUCTION
INFLUENCE OF EXTERNAL CONDITIONS AND
KEY PROCESS PARAMETERS

TA/84/13

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SUMMARY

Recent developments in ammonia production processes have emphasized low energy consumption. A number of new process schemes have been introduced with energy consumption quoted in the range 6.4-7.2 Gcal/MT Ammonia. The figures given may, however, not be comparable, because the external conditions assumed in each case are different.

The purpose of the paper is to review -with the Topsøe Low Energy Ammonia Process as basis- the influence of external conditions on the overall energy consumption. Factors such as cooling water temperature, feedstock composition, required CO₂ production, required product delivery conditions are considered. Furthermore, the influence of certain key process parameters such as reforming conditions, synthesis pressure and refrigeration level are investigated, and it is discussed how the choice of such parameters can depend on factors such as plant location and plant capacity.

ENERGY CONSUMPTION IN AMMONIA PRODUCTION. INFLUENCE OF EXTERNAL CONDITIONS AND KEY PROCESS PARAMETERS.

1. Introduction.

During the last few years several new processes for energy efficient ammonia production have been announced. The energy consumption, which is normally expressed as lower heating value of feed + fuel, is most often claimed to be about 7.0 Gcal/MT of ammonia, although lower values have also been quoted. In some cases it is claimed that specific changes in process layout or operating conditions have resulted in significant improvements in the energy efficiency of the overall process. It is, however, not in all cases clear whether the figures refer to total energy

consumption, or whether e.g. only the process feed and fuel to the primary reformer are included. Furthermore information about the design basis is often not given, and for these reasons it may not be possible on the basis of published data to get a reliable impression of the relative energy consumption for different processes.

One purpose of the present paper is to illustrate - with the Topsøe Low Energy Ammonia Process as basis - to what extent the energy consumption is influenced by external factors such as cooling water temperature, feed gas composition, required production of CO_2 , and required product properties at battery limits. Another purpose of the paper is to discuss the influence of various operating parameters and process features on the overall energy consumption.

2. Theoretical Energy Consumption.

Throughout this paper, the practice of the industry will be followed, i.e. energy consumptions will be expressed in Gcal/MT of ammonia based on lower heating value of feed and fuel. In this context the term "fuel" means fuel for the primary reformer and fuel for the auxiliary boiler.

This is done in spite of the fact that the use of higher heating value would be more logical, which can be illustrated by considering the theoretical energy consumption by production of ammonia. Let us consider 4 different cases:

- a: reaction of methane with water and the stoichiometric amount of air
- b: reaction of methane with water and an excess of air
- c: reaction of methane with water and air with coproduction of an excess of synthesis gas
- d: reaction of n-butane with water and the stoichiometric amount of air.

The theoretical overall reactions are shown in Table 1, and the corresponding consumptions expressed in higher and lower heating values are given in Table 2. It is seen that when higher heating values are used, consistent results are

obtained with the consumption equal to the higher heating value of ammonia. With lower heating values, different results are obtained in each case, and a meaningful comparison is difficult to establish. (It may be noted that if the consumptions expressed in lower heating value are corrected with the heat of evaporation of the water consumed in the reaction, consistent results are obtained also in this case).

Case b and c illustrate two different approaches to the case where all ammonia must be converted to urea. In case b an excess of N_2 is produced which must at some point be separated from other components. In case c an excess of synthesis gas (H_2+N_2) is produced which must be extracted from the process and used as fuel.

For case a and d it is interesting to note that when comparing on the basis of higher heating values, feed consumption is slightly lower for butane than for methane. When using lower heating values, the opposite result is obtained.

It must be emphasized that the energy consumptions shown in Table 2 are of course quite unrealistic since they assume that all reactions are carried out with full conversion and without any losses, and since heat of reaction is credited at full value including the heat of condensation of ammonia.

3. Process Layout

A generalized block diagram for Haldor Topsøe's Low Energy Ammonia Process with natural gas as feed is shown in fig. 1. Three versions of this process are normally considered. The choice between them depends mainly on the relative cost of energy and capital and on the required production of CO_2 as discussed in paragraph 4.4.

Main features of the Process alternatives are summarized in Table 3. A more detailed discussion of the various process steps may be found elsewhere (Dybkjaer, ref 1). Energy consumption for the three cases is given in Table 4. Case 1, with the lowest energy consumption, will be a reference case for the discussion in the rest of the paper.

The energy consumptions for the three cases were evaluated on a consistent basis using whenever possible the same temperature approaches in exchangers, pressure drop in equipment, efficiencies of machineries, etc. A steam system as shown in Fig 2 was assumed in all cases. Only major rotating machines are driven by steam turbines. All other drives are electric motors. Electricity is generated in a generator driven by steam. Additional steam is produced in an auxiliary boiler as required to fulfil the energy requirements of the plant. Fuel to the auxiliary boiler is assumed to be identical to the process feed, and the efficiency of the boiler is 92% based on lower heating value. The same basis was used to evaluate all cases discussed below.

4. Influence of External Conditions.

Let us now, after these introductory remarks, analyse the influence of some external factors on energy consumption. The following will be considered:

- Cooling water temperature
- Feed gas composition (H/C ratio, content of N_2 and CO_2)
- Required CO_2 production
- Required temperature of product ammonia

4.1 Cooling water temperature

The temperature of the cooling water has an influence on the power consumption for make up gas compression and more importantly on the power required for the refrigeration compressor because a larger amount of the ammonia is condensed in the water cooler when the cooling water temperature is low. The overall effect is shown in fig 3. It is seen that the effect especially at low cooling water temperature is quite significant.

4.2 Feed Gas Composition

Natural gases can be characterized by their content of "bulk" impurities (most often CO_2 and/or N_2), "trace" impurities (sulfur compounds, etc), and by the mix of hydrocarbons, which can for the purpose of this paper be expressed as the H/C ratio of the gas.

- Trace impurities will normally have no influence on the energy consumption. They may be catalyst poisons in which case they must be removed carefully before the gas is introduced into the process.
- CO_2 may be present in certain natural gases in significant concentrations. If the concentration is very high it will normally be reduced in a separate purification step before the ammonia process proper. But natural gas with up to 15-20% CO_2 can without problems be used as feed gas for an ammonia plant. The presence of such amounts of CO_2 will have an effect on the operation in several sectors of the plant, but the effect on overall energy consumption is modest, when a physical CO_2 -removal process is used. When steam to carbon ratio at inlet primary reformer, methane leakage from secondary reformer, and equilibrium temperature exit low temperature shift converter are all equal to the values in the reference case, the increase in energy consumption caused by the presence of 10% CO_2 is about 0.05 Gcal/MT ammonia.
- N_2 could in contrast to CO_2 be regarded as a desirable impurity since it is a component required for the synthesis of ammonia. It has, however, only very little influence on energy consumption. A concentration of 10 vol% N_2 in methane will cause a decrease in the overall energy consumption of about 0.02 Gcal/MT ammonia and a slight change in the operating conditions in the reformer due to the decrease in the amount of process air.

- The H/C ratio of the gas has an influence on the overall stoichiometry of the theoretical synthesis reaction as illustrated in Table 1 with n-butane as an example. More water is consumed in the reaction and more CO₂ is produced together with the ammonia.

The effect of the H/C ratio on the energy consumption of the process is modest. For pure n-butane as feed, the consumption expressed in lower heating value will increase by about 0.1 Gcal/MT of ammonia. If higher heating value is considered, the same consumption corresponds to a decrease in energy consumption of about 0.07 Gcal/MT of ammonia compared to the base case with methane as feed.

The H/C ratio of the feed will also have an influence on the acceptable feed preheat temperature due to the risk of steam cracking in the preheat coil when higher hydrocarbons are present.

4.3 Ammonia Supply Temperature

A certain part of the energy consumption in the base case is used to condense the product ammonia. If the ammonia could be supplied as a gas, this energy could be saved. On the other hand extra energy must be used if the product must be supplied to an atmospheric storage as liquid at -33°C. The differences compared to the base case are:

- Ammonia supplied as gas at 3 atm,
saving: 0.14 Gcal/MT
- Ammonia supplied as liquid at -33°C,
extra consumption: Depends on cooling water temperature, 0.07 Gcal/MT with cooling water at 30°C.

4.4 Required CO₂ production

The base case (case 1 in Table 2 and 3) is based on the use of a physical CO₂ removal process with air stripping. In such processes, only about 70% of the available CO₂ is recovered in a form suitable for urea production. In many cases a situation exists where all or a major part of the product ammonia must be converted to urea, and in these cases one of the other alternatives in Table 3 and 4 must be used.

It is seen from Table 4 that the requirement for full CO₂ recovery has a significant influence on energy consumption. It should be mentioned that case 2 and 3 represent only one solution to the problem of full CO₂ recovery. It could also be possible to modify case 1 so that all CO₂ were recovered while maintaining the energy saving features. This alternative would have an energy consumption somewhat below case 2; but the extra investments required would in most situations render the economics unattractive.

5. Energy Balances in the Process

It is not the intention of the present paper to analyse in detail how changes in any of a large number of process parameters will affect the overall energy consumption of the plant. Instead, an attempt will be made to show how the process conditions in two main areas of the plant, the reforming section and the synthesis loop can, without basically changing the overall process layout, be adjusted to change the internal energy balances and to accommodate different external conditions. First it may, however, be illustrative to analyse the main energy flows in the ammonia process.

5.1 Energy Flows

Main energy flows in Topsøe's Low Energy Ammonia process are shown schematically in fig 4. Natural gas enters the process as feed and fuel for the reformer, and if required, as fuel gas for the auxiliary boiler. The feed gas is in the various front end processes converted to synthesis gas which is compressed and further converted to ammonia in the synthesis loop. Some unconverted methane and other inerts are extracted as purge gas from the loop and used together with the primary fuel as fuel for the primary reformer. About half of the energy content of the combined fuel is transferred through the reformer tubes to heat the reaction mixture and to sustain the endothermic reforming reaction. The remaining energy is available in the flue gas and mainly used to preheat the feed streams to

the reformer. Waste heat from the various process steps is recovered mainly downstream secondary reformer, shift, and synthesis converter, and transferred to the steam system for production of high pressure steam. The steam is in turn used to generate power for rotating machines or, after expansion through back pressure turbines, returned to the process as process steam. Depending on the balance between the amount of waste heat generated in the process and the power requirements in the plant, an excess of steam may in some cases be generated which must be credited at a value which depends on the local conditions. In most cases, however, an import of energy to the plant is required to satisfy the needs of the various consumers. This import may be in the form of electricity or fuel to an auxiliary boiler.

5.2 Use of process feed

In the conversion of process feed to final product a yield corresponding to 85-94% of the theoretical is obtained, depending on process layout. In Topsøe's Low Energy Ammonia Process, with a yield of 94%, the "losses" are distributed as shown in Table 5. These "losses" are almost qualitatively recovered, mainly as purge gas from the loop, and recycled to the fuel system, where they can be credited at fuel value. There has, however, on the way through front end and synthesis loop, been invested extra energy in the purge gas (for chemical conversion, compression, etc). This energy - equal to 25-30% of the heating value - must be considered lost, and an increasing amount of purge gas is therefore always equivalent to a decreasing overall energy efficiency, although the sum of process feed and reformer fuel may decrease.

Figures in Table 5 are based on the reference case, where a purge gas recovery unit is foreseen. If a purge gas recovery is not included, the hydrogen losses increase, and the overall yield may drop to about 85-90% of the theoretical mainly depending on the inert level chosen in the loop. The extra loss

is fully utilized as fuel, but as explained above an increase in overall energy consumption will be seen.

5.3 Steam and power generation

In order to obtain a low overall energy consumption it is necessary not only to reduce the consumption of natural gas for reformer feed + fuel, but also to reduce the power requirements of the plant and to increase the efficiency of the steam system and the power generation.

Ideally, the layout should, at least for an isolated ammonia plant, be such that the amount of steam which can be generated from the waste heat is equal to or less than the amount required to generate the power required to satisfy the needs of the plant. If an excess of waste heat is available, a steam export is generated which may be difficult to utilize. If too little waste heat is available, an energy import is required, and this may in many cases give a desirable flexibility in the fuel requirements and in the operation of the plant.

In order to generate as much power as possible from the waste heat it should as completely as possible be transferred to the high pressure steam system. This means that waste heat should be available at such levels that steam superheating, steam boiling, and boiler feed water preheat to a reasonable temperature can all be satisfied.

In Topsøe's process, the steam production and demand are in good balance, and the waste heat is either used for preheat and thus returned to the process or transferred to the high pressure steam system. There are no medium or low pressure steam boilers in the process, and no feed gas saturation or other special installations to utilize low grade is used. Total high pressure steam production is 3.0-3,5 T/MT of ammonia, and the power production (and consumption) in the plant is about 500-650 kwh/MT.

6. Influence of Key Process Parameters

6.1 Operating Conditions in the Reforming Section

The layout of the reforming section in the Topsøe Low Energy Process is shown in fig 5. Main design parameters in this area are:

- Steam to carbon ratio inlet primary reformer
- Preheat temperatures for process feed, process air, and combustion air
- Methane leakage exit secondary reformer
- Consumption of process air

Although these parameters are to a large extent inter-related an attempt will be made to discuss their influence separately.

6.1.1 Steam to Carbon Ratio

Topsøe's proprietary reforming technology is based on the use of a side-fired reformer and of highly active catalyst. With this technology and with the improved understanding available today of the mechanism of catalyst deactivation and carbon formation (Rostrup-Nielsen ref 2) reformers in ammonia plants may now be designed for a steam/carbon ratio of 2.5 at an average heat flux of about 80,000 kcal/m²/hr. Operation at even more severe conditions could be accomplished and is in fact used in other industries.

The steam to carbon ratio has an important influence on process layout including choice of shift and CO₂ removal technology as illustrated in Table 3.

The minimum steam/carbon ratio at inlet reformer can be determined by considering the amount of steam which is necessary, for equilibrium reasons, to obtain a reasonably low carbon monoxide concentration out of the low temperature shift converter. At an equilibrium temperature of 200⁰C, and a carbon monoxide leakage of 0.3 vol% (dry) it corresponds to a steam/dry

gas ratio of about .16 or a steam/carbon ratio at inlet reformer of about 2.3. This would then be the practical minimum steam/carbon ratio, if no other restraints were present. With the classical high temperature shift catalyst, which consists of iron oxide promoted with chromia, it is, however, not possible to operate at such low steam contents in the gas. The reason is that at low steam/dry gas ratio, the iron oxide will be converted to iron carbide (ref 1). The carbide is a Fischer-Tropsch catalyst and will cause formation of significant amounts of hydrocarbons, especially methane and lower olefins. The limit imposed by the risk of carbide formation depends on a number of parameters. In most cases it corresponds to a steam/carbon ratio at inlet primary reformer slightly above 3.0, and this then becomes the practical lower limit when classical catalysts are used. At lower levels new catalysts, such as those developed by Topsøe, must be used.

Low steam to carbon ratio combines best with a physical CO₂ removal process with no requirement for low grade heat for regeneration of the solvent. Since such processes recover only part of the CO₂ available in the process gas, "high" steam to carbon ratio is indicated when full recovery of CO₂ is required (cfr. table 3 and 4, and paragraph 4.4).

6.1.2 Preheating Temperatures.

The preheating temperatures of combustion air and process feed will mainly have an influence on the amount of fuel required for the reformer and thereby on the amount of heat available in the flue gas. Heat available in the flue gas is used for steam superheating and for preheating of reformer feed, process air, and combustion air. If no combustion air preheat is used there will, with other parameters unchanged, be an excess of heat available, and the plant becomes steam exporting. When high preheat temperatures are used, the heat available in the flue gas decreases, and some steam superheating must be done in a combined boiler/steam superheater downstream the secondary reformer. If all steam superheating

is done at this point, a consumption of natural gas for reformer feed + fuel of slightly below 6.6 can be obtained. If changes in the process design are introduced which further reduce the heat available in the flue gas, it may not be sufficient to satisfy all preheat duties. In such cases some preheating duty must be found in other parts of the plant. The solution suggested by Topsøe is to preheat the feed in a gas/gas heat exchanger after the secondary reformer.

If nothing else is changed in the plant, high preheat temperatures will have no influence on the overall energy consumption in cases where steam production and power consumption are in balance. In such cases the effect will only be to move consumption from the reformer to the auxiliary boiler.

6.1.3. Methane Leakage and Process Air/Process Feed Ratio

A high methane leakage exit secondary reformer will reduce the duty of the primary reformer and consequently the fuel consumption to such an extent that the reformer feed + fuel will decrease slightly in spite of a significant increase in the consumption of process feed. The total energy consumption of the plant will, however, increase because of the extra energy required to carry the methane through the process train before it is recovered in the purge gas and used as fuel.

A decrease in the reformer duty can also be obtained by changing the consumption of process air. If the amount of process air increases, a significant part of the duty of the primary reformer will be transferred to the secondary reformer. This effect is to some extent obtained by the introduction of a purge gas recovery unit, which increases the yield of product from a given amount of feed and thereby increases the process air/process feed ratio required to produce stoichiometric feed gas at the synthesis reactor.

The effects of a change in the CH₄ leakage exit secondary reformer and of deletion of the purge gas recovery unit foreseen in the base case are illustrated in Table 6.

6.2 Conditions in the Synthesis Loop

In a separate study on the energy efficiency of ammonia synthesis (Dybkjaer and Gam, ref 3) the effect of various loop design parameters including catalyst activity on the energy consumption in an ammonia synthesis loop was investigated. It was concluded that for a given loop layout such as the Topsøe S-200 Low Energy Loop the energy consumption depends only marginally on the actual combination of design parameters such as synthesis pressure, refrigeration level, etc. The energy consumption is mainly a function of the temperature at which the thermodynamic equilibrium of the ammonia synthesis reaction is established in the reactor system. It has especially been found that when the optimum levels of operating parameters are selected, the overall energy consumption does not change significantly when the synthesis is varied in the range 80-220 kg/cm²g.

In a specific case, the operating conditions including the pressure must then be chosen so that the heat recovered in the reactor system can all be utilized in the high pressure steam system, and so that reasonable sizes are obtained for the main equipment, especially the synthesis reactor and the large compressors. With Topsøe's low energy technology, a loop pressure of 140 kg/cm²g has been found to give a satisfactory balance between heat recovered in the boiler and the BFW preheater and also to give realistic sizes of all equipment even for very large capacities.

The layout of the Topsøe S-200 ammonia synthesis loop is shown in fig 6. The loop contains two refrigeration stages and the condensation of the product ammonia is done in a one casing refrigeration compressor. The make-up gas is added to the loop upstream the last chiller, where most of the ammonia has already

been condensed. Traces of CO_2 and H_2O are removed by co-condensation in the chiller so that the risk of poisoning of the synthesis catalyst with these compounds is eliminated without the use of molecular sieves for drying of the make-up gas. This arrangement is highly energy efficient. The extra energy savings which could be obtained by installation of a molecular sieve unit and by adding make-up gas downstream the separator is only about 0.03 Gcal/MT NH_3 (with no consumption assumed for refrigeration of the molecular sieves). These savings cannot justify the extra investment and the added risk of dust carry over from the molecular sieves to the synthesis reactor.

7. Conclusions

The influence on energy consumption in ammonia production of external conditions and certain process parameters has been analysed with Topsøe's Low Energy Process as basis. The following has been found:

- An increase in cooling water temperature of 10°C will cause an increase in total energy consumption of 0.15-0.20 Gcal/MT NH_3 depending on the temperature level.
- Feed gas composition (H/C-ratio and content of CO_2 and N_2) has only marginal influence on the energy consumption.
- Required CO_2 -production has an important influence on energy consumption. Full recovery of all CO_2 in the process gas will increase energy consumption by about 0.3 Gcal/MT. Production of sufficient CO_2 for conversion of all NH_3 to urea will require an additional 0.2 Gcal/MT NH_3 consumption, when pure methane is the feed.
- Cooling of NH_3 product to -33°C require an extra consumption of 0.07 Gcal/MT compared to supply at ambient temperature. If NH_3 could be supplied

as a gas, 0.14 Gcal/MT could be saved.

- The operating conditions in the reformer area have an important influence on overall energy balances in the process. Preheat temperatures for feed streams have no direct influence total on energy consumption, but on the distribution of fuel for reformer or auxilliary boiler. Increased methane leakage exit secondary reformer will reduce duty of primary reformer and reformer feed + fuel but increase total energy production. An increase of the Process air/Process feed ratio (obtained e.g. by installation of a purge gas ratio) will reduce duty of primary reformer and reformer feed + fuel and also reduce the total energy consumption.
- The energy consumption in the synthesis loop is mainly a function of synthesis equilibrium temperature (or installed catalyst volume). Operating conditions such as synthesis pressure, separator temperature etc. are chosen to allow integration of recovered heat into the steam system and reasonable sizing of major equipment.

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Table 1

Theoretical Overall Reactions for Production of Ammonia

- a: $0.4422 \text{ CH}_4 + 0.6155 \text{ H}_2\text{O}(l) + 0.6407 \text{ AIR} \longrightarrow$
 $\text{NH}_3(l, -33^\circ\text{C}) + 0.4424 \text{ CO}_2 + 0.0060 \text{ Ar} + 36.595 \text{ J}$
- b: $0.4997 \text{ CH}_4 + 0.5006 \text{ H}_2\text{O}(l) + 1.1896 \text{ AIR} \longrightarrow$
 $\text{NH}_3(l, -33^\circ\text{C}) + 0.50 \text{ CO}_2 + 0.4284 \text{ N}_2 + 0.0112 \text{ Ar} + 87.747 \text{ J}$
- c: $0.4998 \text{ CH}_4 + 0.6956 \text{ H}_2\text{O}(l) + 0.7241 \text{ AIR} \longrightarrow$
 $\text{NH}_3(l, -33^\circ\text{C}) + 0.50 \text{ CO}_2 + 0.1952 \text{ H}_2 + 0.0651 \text{ N}_2 + 0.0068 \text{ Ar} + 32.052 \text{ J}$
- d: $0.1361 \text{ C}_4\text{H}_{10} + 0.8196 \text{ H}_2\text{O}(l) + 0.6407 \text{ AIR} \longrightarrow$
 $\text{NH}_3(l, -33^\circ\text{C}) + 0.5445 \text{ CO}_2 + 0.0060 \text{ Ar} + 34.385 \text{ J}$
 $\text{AIR} = 0.2099 \text{ O}_2 + 0.7804 \text{ N}_2 + 0.0003 \text{ CO}_2 + 0.0094 \text{ Ar}$

Table 2

Theoretical Energy Consumption Expressed in Higher and Lower Heating Value (Gcal/MT Ammonia)				
	a	b	c	d
Higher Heating Value:				
Feed Consumption	5.52	6.24	6.24	5.49
Heat of Reaction	0.51	1.23	0.45	0.48
Byproduct Credit	-	-	0.78	-
Net Consumption	5.02	5.01	5.01	5.01
Lower Heating Value:				
Feed Consumption	4.98	5.63	5.63	5.07
Heat of Reaction	0.51	1.23	0.45	0.48
Byproduct Credit	-	-	0.66	-
Net Consumption	4.46	4.40	4.52	4.59

Process Sequence for Ammonia Manufacture

Fig. 1

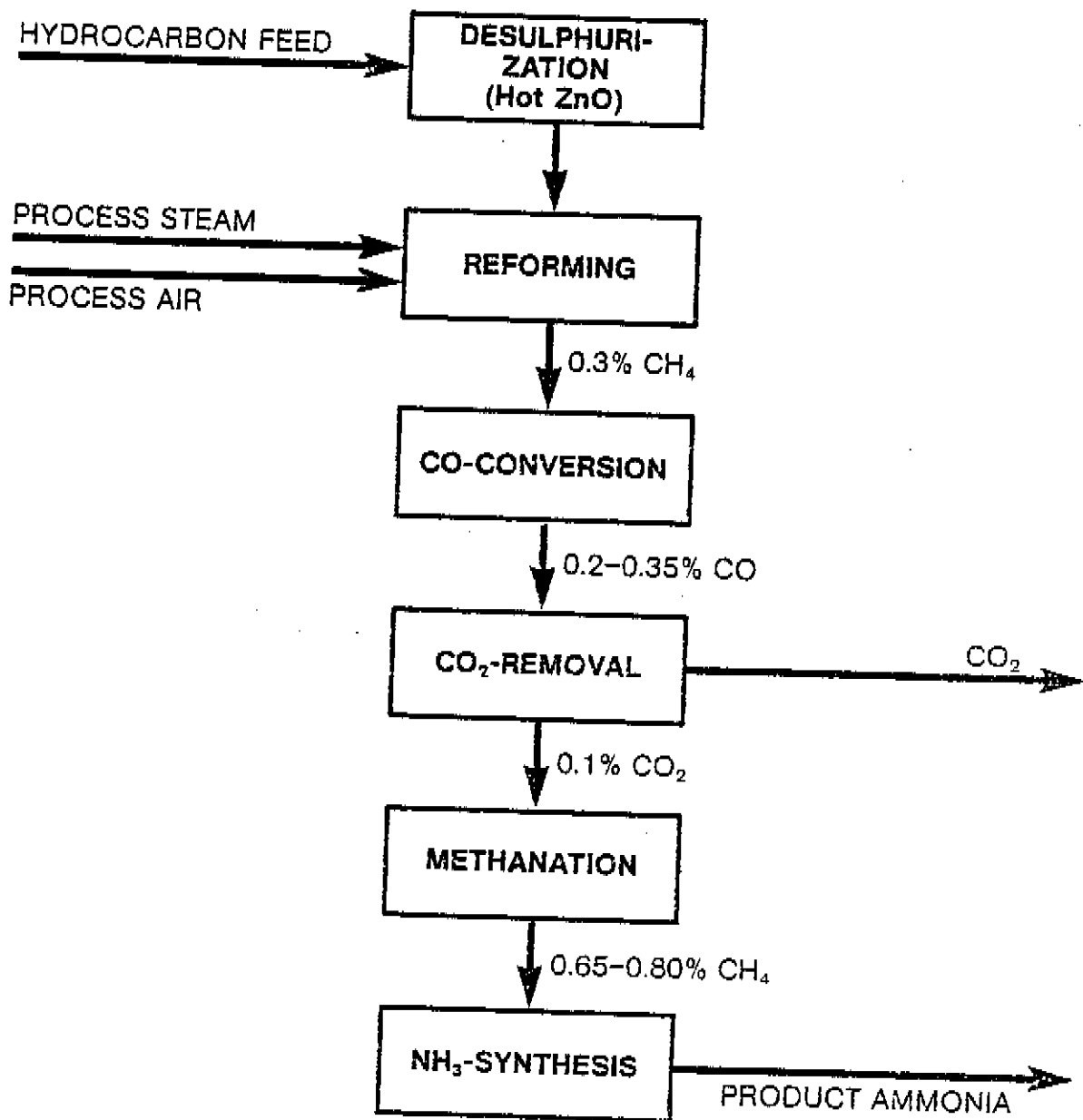


Table 3

Alternatives of Topsøe Low Energy Ammonia Process			
Alternative	1	2	3
Steam/Carbon Ratio	2.5	3.3	3.3
Shift Lay-out	New	Conventional	Conventional
CO ₂ -Removal	Physical	Chemical	Chemical
Synthesis Loop	S-200	S-200	S-200
Synthesis Pressure (kg/cm ² g)	140	140	140
Flue Gas Temperature (°C)	100	100	100
Comb. Air Preheat Temperature (°C)	350	350	350
Purge gas Recovery Unit	Yes	Yes	No

Table 4

Consumption Figures for Topsøe Low Energy Ammonia Process			
(Figures per MT of Ammonia)			
Alternative	1	2	3
Net energy Consumption, Gcal (LHV)	6.7	7.0	7.2
Cooling water, m ³ (delta T = 10°C)	200	230	255
Major Basis for Energy Balances:			
Natural gas			
- Delivery pressure	42 kg/cm ² g		
- Composition	100% CH ₄		
Cooling water temperature	20°C		
Product ammonia	Liquid NH ₃ at ambient temperature		

Topsøe Low Energy Ammonia Process Simplified Layout of Steam System

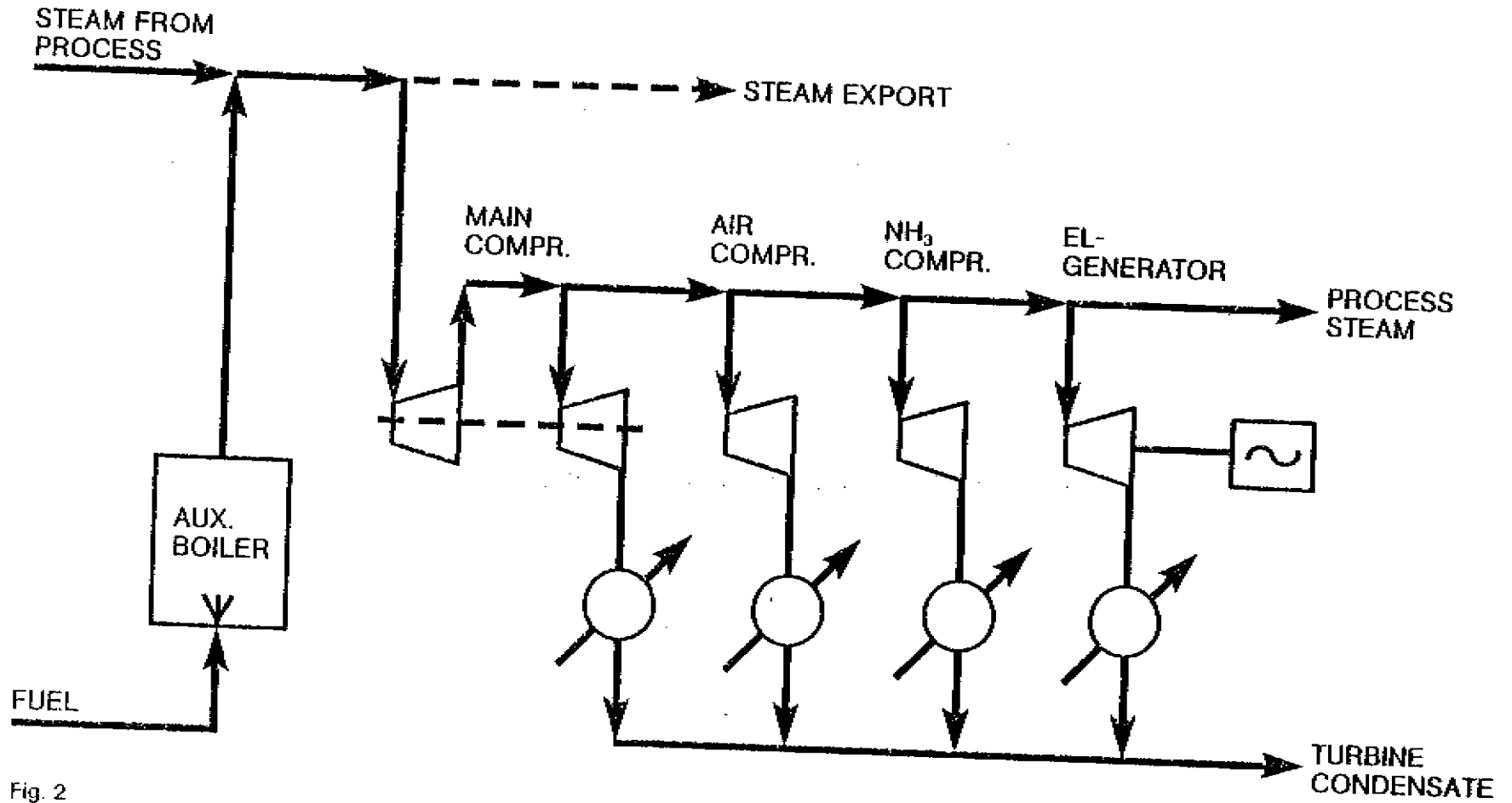


Fig. 2

Energy Consumption vs Cooling Water Temperature

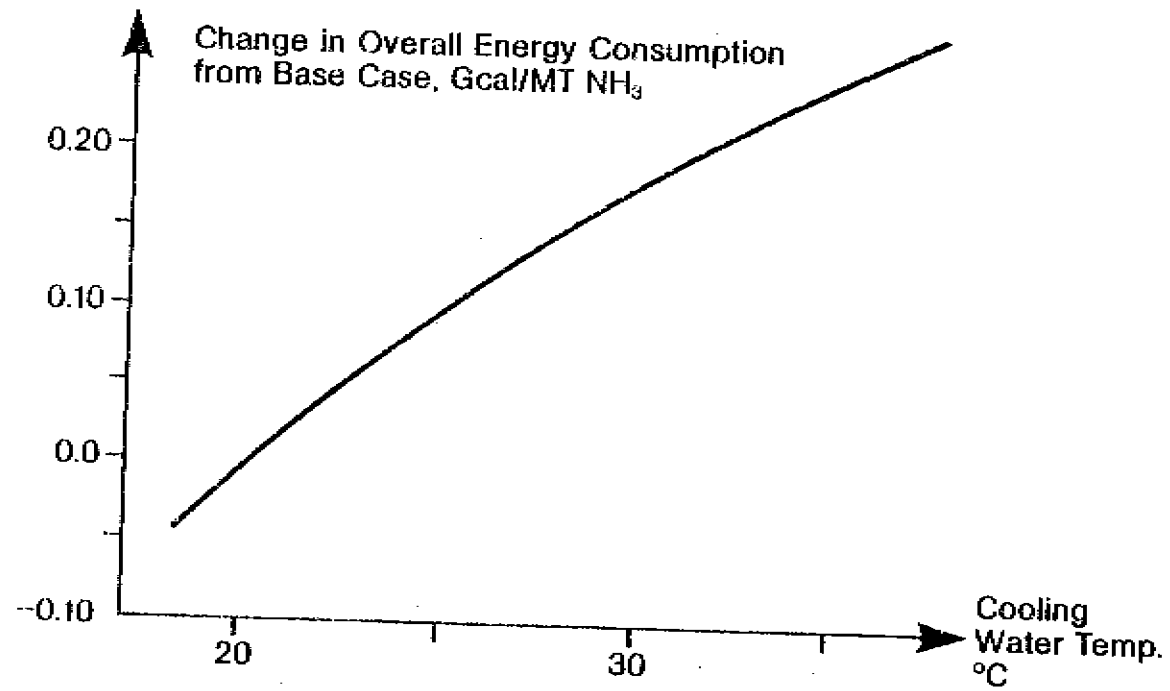


Fig. 3

Main Energy Flows in Topsøe's Low Energy Ammonia Process.

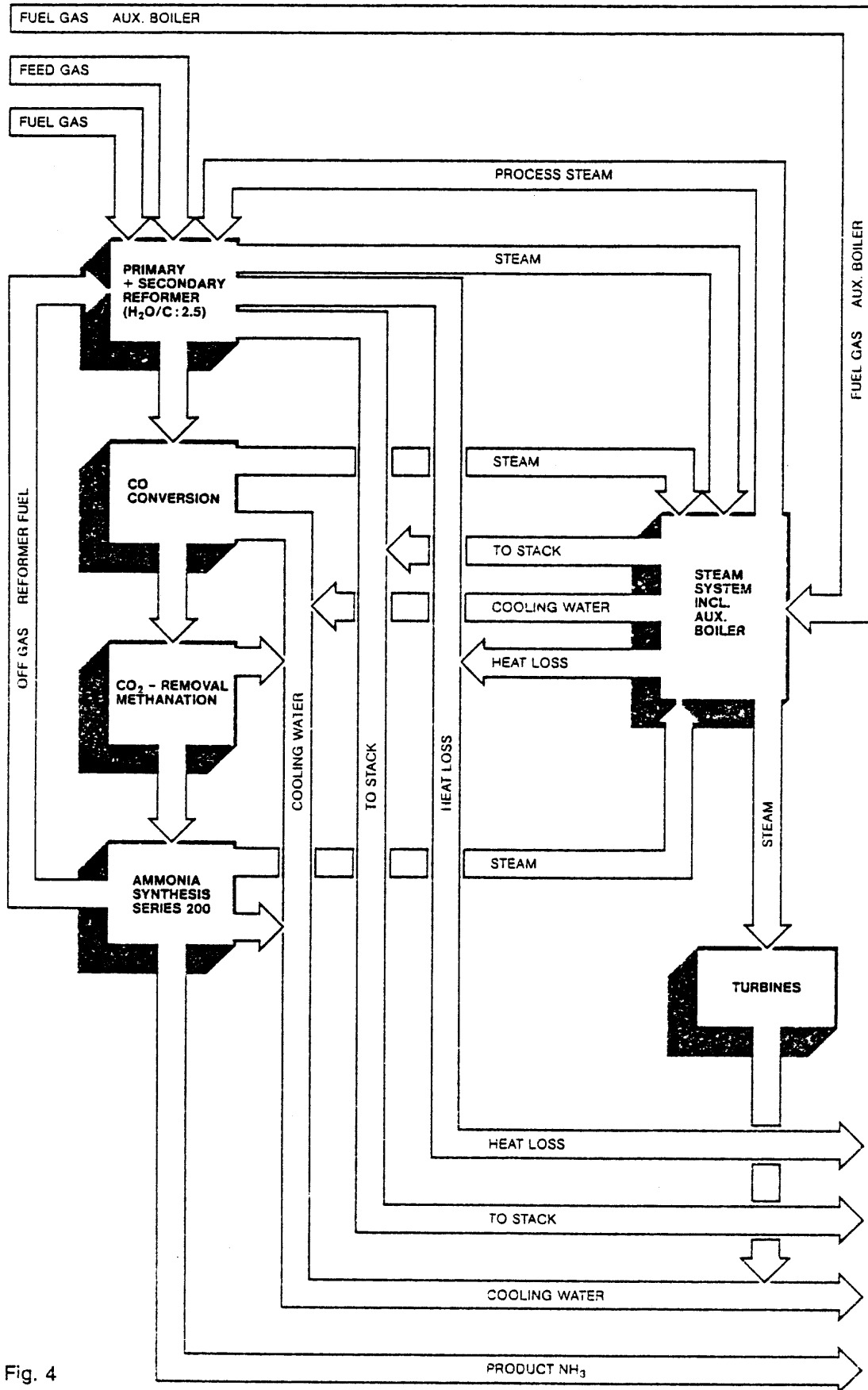


Fig. 4

Topsøe Low Energy Ammonia Process Layout of Desulphurization and Reforming

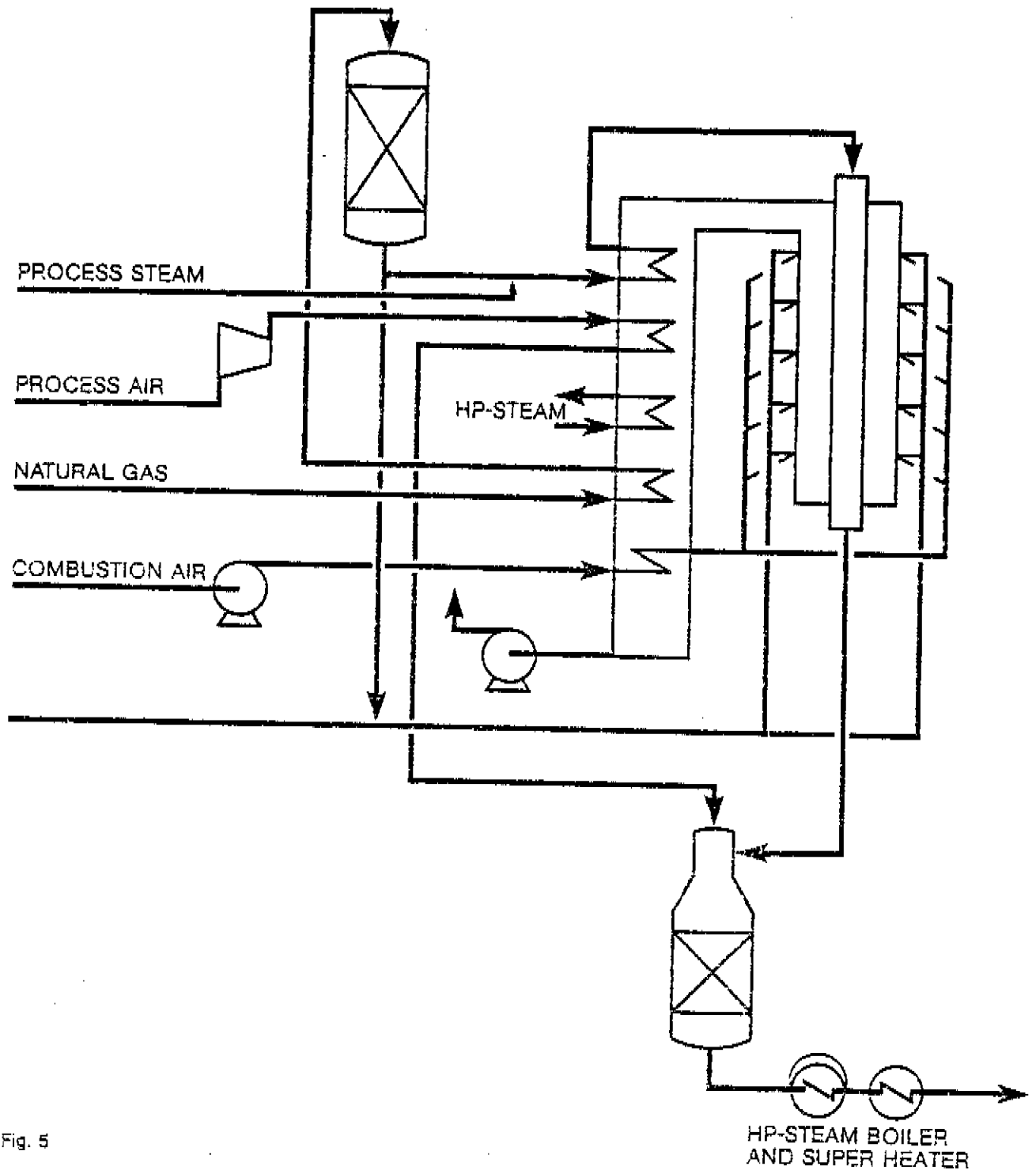


Fig. 5

Table 5

Conversion of Process Feed to Product Distribution of Losses	
Non-Converted CH ₄	25%
Non-Converted CO	8%
Methanation of CO and CO ₂	34%
Loss with purge gas	22%
Misc. Losses	<u>11%</u>
Total	100%
Total Losses correspond to 6% of Feed.	

Table 6

Effect of Methane Leakage and Purge Gas Recovery			
	Base Case	High CH ₄ -leak	No PGR-Unit
CH ₄ -leakage	0.3	1.0	0.3
P.G. Recovery	Yes	Yes	No
Process Air/Feed gas, Nm ³ /Nm ³	1.47	1.47	1.43
Relative Duty, Prim. Ref.	1.0	0.87	1.10
Relative Consumption: Ref. Feed + Fuel	1.0	0.98	1.02
Total	1.0	1.01	1.02

Topsøe Low Energy Ammonia Process Layout of Synthesis Loop

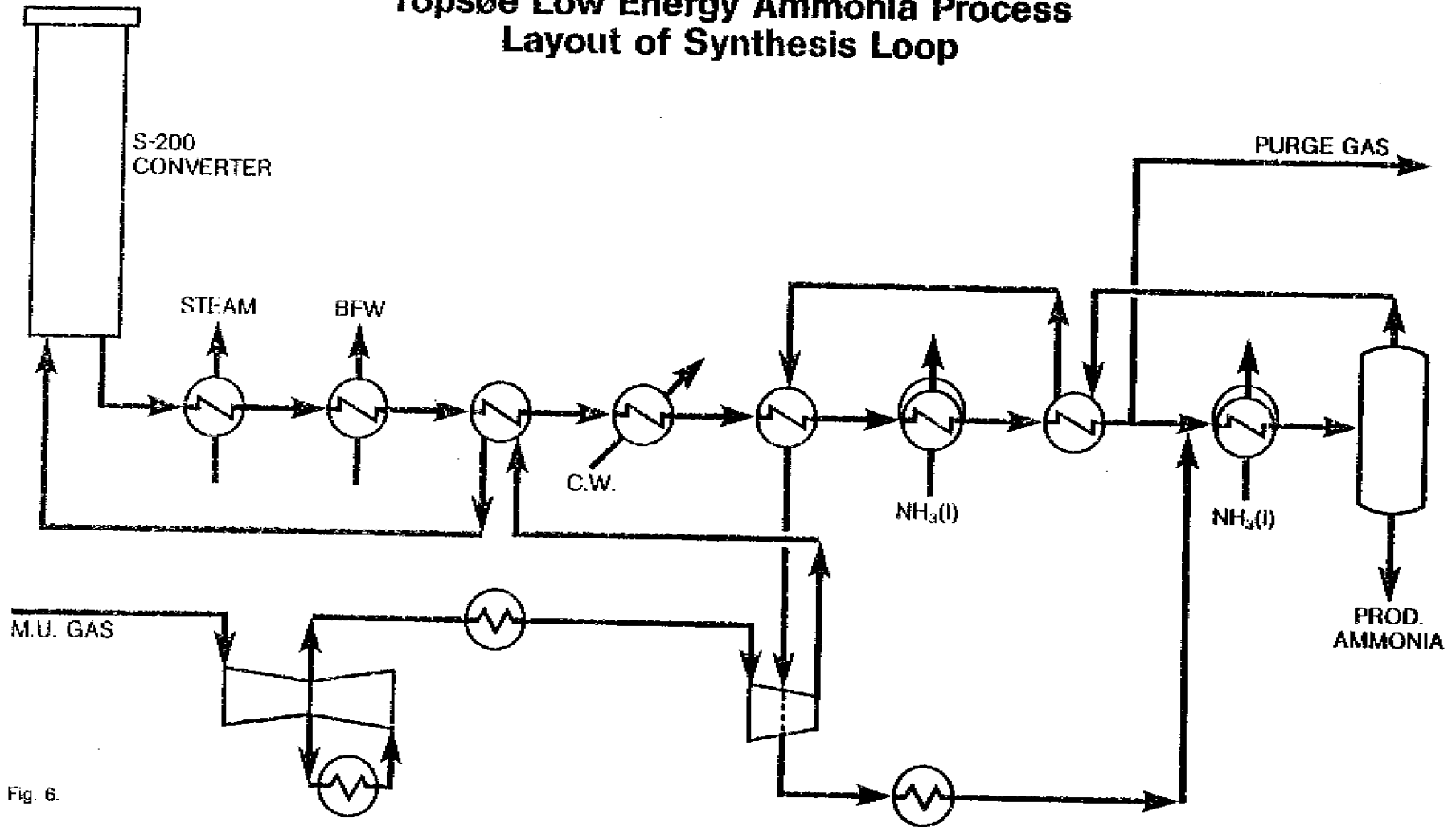


Fig. 6.

TA/84/13 Energy consumption in ammonia production. Influence of external conditions and key process parameters by Ib. Dybkjaer, Haldor Topsoe, Denmark

DISCUSSION: Rapporteur N.D. WARD, Norsk Hydro Fertilizers Ltd, United Kingdom

Q - Mr. B.K. JAIN, FAI, India

It is mentioned that production of sufficient CO₂ for conversion of all the NH₃ to urea will require some additional energy consumption.

I suppose the endeavour should be to produce maximum CO₂ and utilize this in the production of urea even, if need be, by importation of additional NH₃ to consume all the CO₂. The incremental urea production would perhaps offset the marginal increase in energy consumption when it is apportioned to the total urea production. Your comments please.

A - The problem of production of CO₂ and the match between the CO₂ and the NH₃ in the urea plant can be tackled in different ways. The problem is only relevant when you have a rather clean natural gas. If you run on naphta or heavy natural gas, the balance between H₂ and C in the feedstock is such that sufficient CO₂ is produced automatically.

If you have dry NG or pure methane, as we have assumed, then it is necessary in some way to increase the production of CO₂, if it is not possible in the local conditions in which the plant is built to coproduce NH₃ and to export that out of the plant.

It can be done, as you can see from some theoretical equations included in the paper, in two different ways: one is to add an excess of air to the secondary reformer. In this way you will upset the balance between the H₂ and N₂ in the synthesis gas and you will have to spend energy and investment to take the excess N₂ out of the synthesis gas. But you will be able to achieve the balance between NH₃ and CO₂. It cannot be done in this way without additional energy consumption.

It can also be done, and that is what we have assumed and realized in plants built in various parts of the world, by producing an excess of synthesis gas. This is done simply by having a somewhat larger front end which produces sufficient CO₂ for the conversion of the ammonia. The excess synthesis gas is then taken out before the compression, i.e. after the CO₂ removal, and used as feed in the reformer. The extra energy consumption, by doing it this way, is marginal. In such a case, it does not pay to install a purge gas recovery unit, which is installed to reduce the production of synthesis gas.

Q - Mr. M. SIVONEN, Kemira Oy, Finland

Is the steam/carbon ratio of 2.5 already used continuously in any conventional ammonia plant?

A - No, this steam/carbon ratio is not yet used in any conventional

ammonia plant. Plants are under construction where operation at this ratio will be possible. But similar operation conditions are used in conventional steam reforming plants for other productions, such as H₂, where the conditions in the primary reformers are very similar to what they are in an ammonia reformer plant. Even lower steam to carbon ratios are used in steam reformers producing oxo-synthesis gas or gas for other purposes.

Q - Mr. D. BANQUY, Foster Wheeler, France

Did you evaluate the effect of the synthesis pressure on the overall investment cost?

A - Yes, we did. It is of course not a simple thing to do. We have studied the effect of the synthesis pressure. Since we found that we could not by analyzing the energy consumption alone find any optimum pressure, we selected it on the basis of other criteria, like:

- Size and cost of equipment
- Reasonable layout of the compressor section
- Reasonable integration of the heat recovery system.

I would like to emphasize that we see no problem whatsoever in supplying a synthesis loop at 200 atm or at 80 atm, as we have the technology for it and we are quite willing to do that. For our own complete process scheme, we have found that a pressure of approx. 140 atm is advantageous when we consider the investment cost and the overall integration of the energy system.