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

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TECHNOLOGY PROFILE ON MINI FERTILIZER PLANTS*

Prepared by

UNIDO Secretariat

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* This document has not been edited.

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TECHNOLOGICAL PROFILE ON MINI FERTILISER PLANTS

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1. INTRODUCTION

1.1 Global review of fertilizers

A review of world agriculture during the last 20 years reveals minimal expansion of cultivated land and a 50 per cent increase in world population. Food supply per head has increased by five per cent since 1961. But regional disparities exist. Productivity in developed countries, where consumers have accepted high food prices, has been consistently high. In contrast, although some spectacular results have been achieved, production in 70 per cent of developing countries has been inadequate.

The FAO states that the world has the potential to feed a population of 6.2 billion in the year 2000 moderately better than it fed 4.4 billion in 1984. It also stated that lack of adequate food could be the energy crisis of the 1990's.

By the 1990's there will be 100 million more to feed each year.

Feeding such numbers means that world food production must be increased by 60 per cent over the next 15 years. Developing countries will have to double their food production just to keep pace with their population growth. What if this cannot be achieved? One result will be that the numbers of undernourished and starving will increase.

By the year 2000, the numbers of seriously undernourished in the world could be as high as 600 million, the Far East and Africa being the worst hit.

In order to improve agricultural performance to the point of reaching food self - sufficiency, crop production must be increased.

There are three ways to do this:

- increase arable land,
- increase the harvested proportion of that area,
- above all, increase yields from each unit of land harvested.

The possibility of cultivating new land varies from region to region, and only a quarter of the required increase in crop production can come from growth in arable and harvested areas. The rest must come from more intensive use of existing land, through increased inputs and improved management.

Fertilizers are generally considered the key input to agriculture and food production. Several independent estimates suggest that, in the absence of fertilizers, expected crop yields could be halved. Effects on agricultural output would go farther than this, because fertilizer use is now an essential component of farming systems.

Consumption of the principal fertilizer nutrients - Nitrogen, Phosphate and Potassium - has risen sixteenfold in the last 40 years - from 7,5 million tons in 1945/46 to 120 million tons in 1984/85. This corresponds to around 350 million tons product.

By 1980, developed countries were using, on average, 127 kg of nutrients per hectare, whilst developing countries were using some 31 kg. Even within developing countries there are wide differences in application rates. For example, in Latin America it is 42 kg, while in Africa the average rate is only eight kg per hectare.

Clearly, considerable potential exists worldwide to increase fertilizer application rates. Indeed, in order to double food production by the year 2000 (FAO target), up to 75 per cent of which must come from higher yields, developing countries must use more and better fertilizers more efficiently.

Forecasts by the UNIDO/FAO/World Bank Joint Working Group on Fertilizers and by the industry itself suggest, that over the next ten years, fertilizer nutrient consumption will grow at an annual rate of between three and four per cent.

These forecasts represent percentage increases of 40 per cent for Nitrogen, 32 per cent for Phosphates and 35 per cent for Potash, for the period 1984 to 1995. The FAO believes that fertilizer consumption growth rates will have to be maintained through to the year 2000, and even increased, if the world's regional food production targets are to be achieved.

The correlation between population trends and Nitrogen use tends to support this FAO statement Fig. 1. The increasing world population and the increasing world nitrogen use over the last 20 years or so have been so closely and linearly related that population levels alone can be used to help determine future nitrogen demand. If this trend line is extrapolated to the estimated population of 6.2 billion in 2000, it indicates a Nitrogen use, in that year, of more than 120 million tonnes.

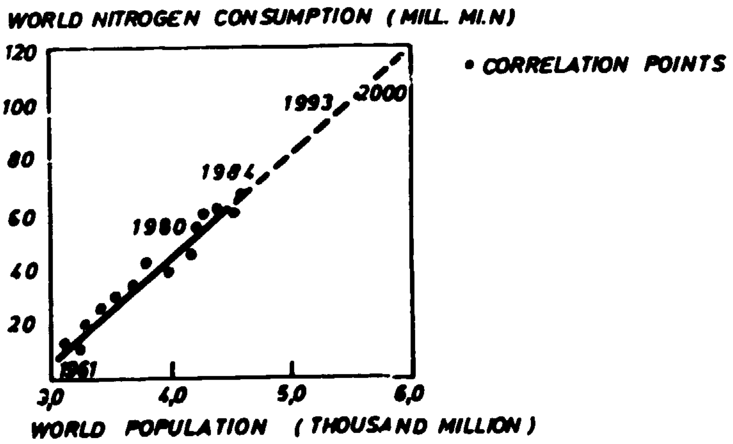
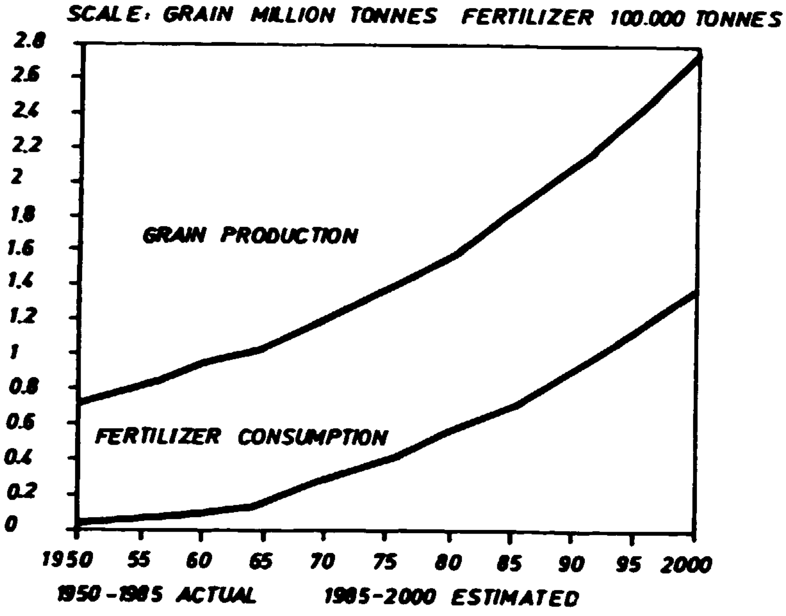


Fig. 1. WORLD POPULATION/NITROGEN FERTILIZER USE

(SOURCE: FAO)

A similar correlation seems to emerge from the diagram on Fig.2. between grain production and fertilizer consumption.



SOURCE: WORLD BANK

Fig.2. WORLD CONSUMPTION OF FERTILIZER AND GRAIN PRODUCTION

If, on a global basis, the quantity and quality of food produced in 2000 is to meet FAO targets, more then 120 million tonnes of Nitrogen will be consumed - possibly around 135 million. In short, this translates into a world wide increase in fertilizer consumption of around 80 per cent over the next 15 years - an annual consumption growth rate in excess of five per cent.

Can the fertilizer industry meet this demand? Only if it builds plants at a present day cost of around 108 billion USD - 66 billion USD for new capacity and 42 billion USD for replacement plants and refurbishing. This is the considered opinion of industry experts.

This would correspond to around 200 plants with an average capacity of 1000 ton/day ammonia, the usual size for today's big plants, situated possibly in the industrialised and in those developing countries where the necessary infrastructure, feedstocks, operational experience exist. All other developing countries would have to pay in foreign exchange for the fertilizer they would use.

Another possibility is to build a considerable number of small plants (miniplants) in the developing countries which do not have a big enough market, raw materials and infrastructure for the big capacities.

The choice or rather the ratio between the capacities to be realised through the building of big plants and those implemented in form of miniplants will depend on the competivity of the miniplants. Most developing countries would be willing and able to build small fertilizer plants if they could be convinced that this is the most economic way to cover their needs and at the same time this solution fits in well in their general industrialisation and technological development efforts. So the question of the size is of paramount importance for the future development of the fertilizer industry in the developing world.

1.2 Fertilizer plant size

Until the sixties, the fertilizer plants considered today as the lower end of the miniplant category, represented the maximum sizes realisable. A single-line ammonia plant with 200 tpd capacity was beyond the reach of most companies in the fifties. Technical developments opened wide the doors before the implementation of huge single-line plants. Some of these developments, like the use of the steam turbine - driven centrifugal compressors in the ammonia synthesis were not realisable below a rather high capacity (around 600 tpd by that time). The big units lowered the specific investment costs, reduced considerably the consumption figures and made possible the development of an export oriented industry in the industrialised countries, situated mostly near sea ports.

In many cases, the industrialisation of the developing countries tried to follow this trend. Jumbo plants were erected in remote locations without adequate infrastructure, industrial background, and logistic facilities. In many other developing countries, like China, India and Mexico, the fertilizer industry started with smaller units. Only when the necessary skills and experiences were acquired and the infrastructure created, did the building of big units begin, where the market allowed. Experiences of many new small plants built in the industrialized countries proved that the small-capacity plant was still needed.

At the request of several conferences, consultations and export meetings, UNIDO prepared a study: Minifertilizer Plant Projects (UNIDO/IS.416, Sectoral Studies Series No.7. Vol. 1-2). This study demonstrated the viability and importance of the mini fertilizer plants for the developing countries. It showed, that although the specific investment and the factory gate production costs will be always higher than those of a big plant in an industrialized country, the landed costs of the fertilizer at the farm gate will be competitive or lower, when delivered from a miniplant in the neighbourhood of the agricultural market. Many other factors contributed also to the renaissance of the miniplant concept. Scarcity of capital, mobilisation of local resources, the much less demanding implementation, operation and maintenance, the quicker realisation are some of the advantages which point in favour of the small capacities.

The 100 billion dollar investment mentioned earlier would certainly render futile all plans for an adequate and sufficient food supply for mankind, if only big investment, mobilizing state and international capital were possible.

The main question is: are there reliable and proven technologies available for the miniplants which could be competitive with the modern big plants. Is a scale - down possible which would conserve the technical and economic advantages of the processes and equipment developed for the big capacities but at the same time simple enough for the often difficult conditions in remote locations with poor infrastructural background?

In order to answer this question, let us first review the structure of the fertilizer industry.

1.3 The fertilizer industry

The fertilizer industry has a very complex structure. As can be seen from Fig. 3., there are many routes leading to the end products, which contain one or more nutrients. From this complex structure, potash is omitted, since it is more a product from mining and beneficiation, and not an industrial one. In the nitrogeneous fertilizer field, only ammonia and urea present real problems. All others have well proven technological processes which are realised constantly in developing and industrialized countries. It is the same with the phosphorous fertilizers. The commun granulation technics used for all fertilizers however have also some special features from the point of view of the miniplants. Therefore this profile reviews the ammonia and urea processes and the granulation technics only.

For the ammonia plant sizes from 100 to 350 tons NH3/day will be considered as miniplants with the corresponding sizes for the urea production and granulation.

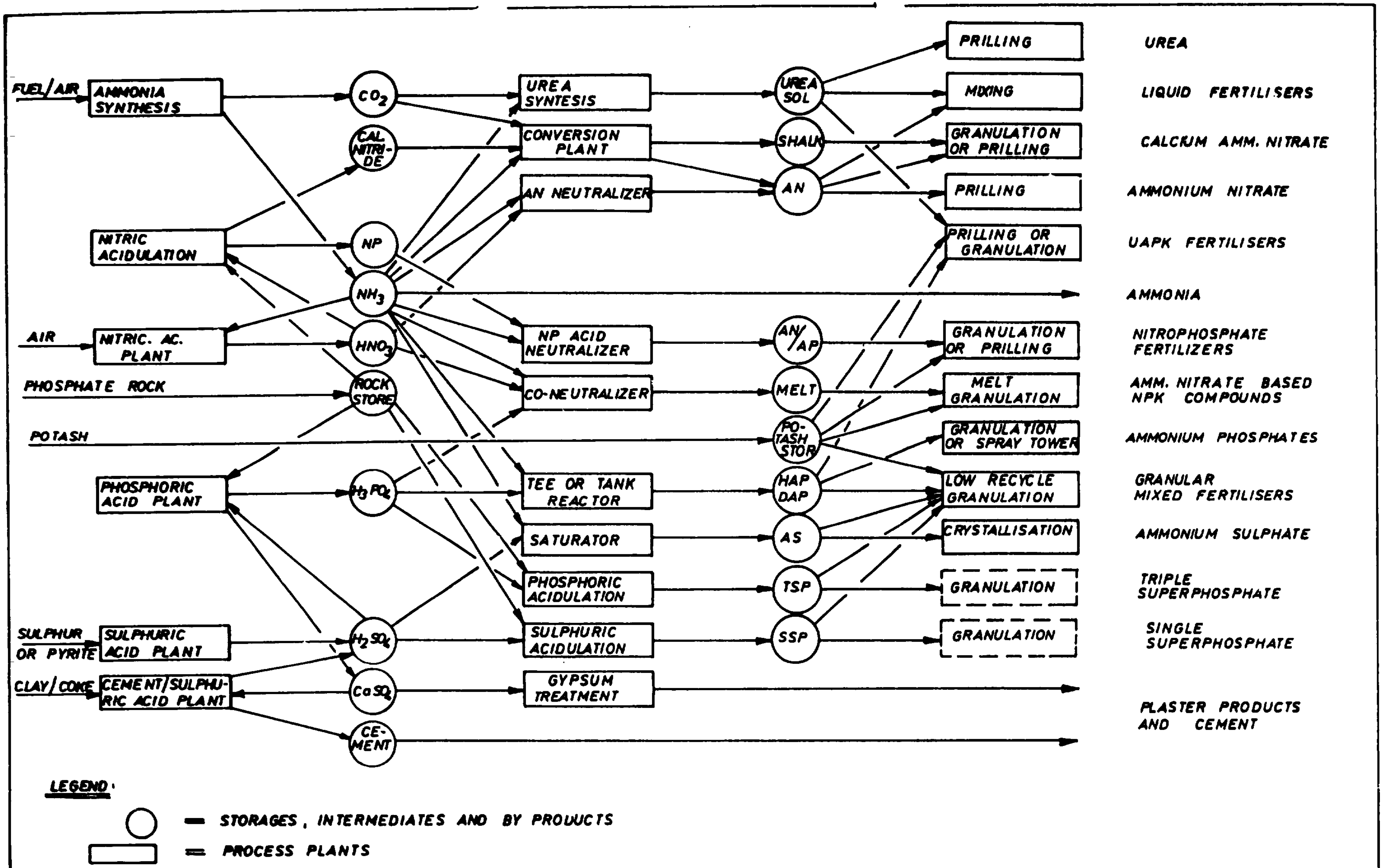


Fig. 3. MAIN PROCESSES OF THE FERTILISER INDUSTRY

The nitrogen industry

While most raw materials occur in nature bound in a compound form and must be separated for utilisation, nitrogen, a vital element of all life, is available in free form in illimited quantitites in the atmosphere. Living organisms, however, with the exception of a few bacteria, can use only fixed nitrogen. Over every hectare of the earth, there are 77,000 tons of free nitrogen, but to fix the amount necessary for a good crop on one hectare of arable land, about 8 Gigajoules are necessary. So the problem of the nitrogen fertilizer industry is really that of the fixation of the atmospheric nitrogen, a big energy consumer process.

From the various nitrogen fixation methods developed during the last century, today only the ammonia route has industrial importance, and no change in this respect is foreseen in the ne future. So the ammonia synthesis is the basic step of all nitrogen fertilizer production. The ammonia thus produced is further processed either to urea or to ammonium nitrate, the two most popular fertilizers. Ammonium sulphate was once widely used, but is constantly losing ground. Ammonium bicarbonate is a specific product developed in China for miniplants.

History of the ammonia processes

For many years ammonia was made in small units, 10-30 t/d, using coal - derived gas as the source of hydrogen. With the development of catalytic steam reforming in the USA, natural gas began its rise to the position of dominant feedstock. At first, plant size remained small and specific energy consumption remained high. By the 1950's reforming pressures were rising but it was at the beginning of the sixties that major technological breakthroughs were made in steam reformer design, centrifugal compressors, large quench type ammonia reactors and integrated steam and power systems. Many of these features needed large capacities and before long the minimum size of integrated plant was 600 t/d with the more normal capacity being 900-1000 t/d.

Apart from the steam reforming developments and the introduction of low temperature CO shift catalyst, most of the other catalysts were well proven. So the advance was not led primarily by the chemistry, but by the better mechanical engineering and metallurgy.

It was not until the oil crisis of the mid - seventies, which quadrupled the price of oil, that improvements in the design of CO2 removal units, ammonia synthesis loops and the inclusion of physical chemistry developments like cryogenic and membrane hydrogen recovery led to the next big reduction in specific energy consumption.

Since these large plants obtained a 6-7 % reduction in energy consumption by raising steam at 120 bar compared with the normal refinery practice of 45 bar, ammonia plant capacity generally settled out at 600 to 1350 t/d to make the best use of high pressure steam. Ammonia plants of 1000-1500 tpd capacity were built in developed countries using inexpensive natural gas or straight-run naphta. The ammonia or ready - made fertilizer were in great part shipped to markets situated usually far away. Low cost feedstock, booming fertilizer market, rapidly expanding industrial and economic growth, low cost credits and investment - all contributed to the world-wide concentration of the ammonia production in big units and a highly developed world market. Process technology was adopted to the technical level of the industrialized countries and gradually became more and more sophisticated. Highest possible temperatures, pressures and severity were aimed at in reforming, higher possible capacities in the whole line. No wonder then, that when developing countries with the necessary raw materials followed suit, they ran into trouble. Serious delays and cost overruns in project implementation, low on-stream factors, operational troubles defeated the economic advantages hoped for.

Developing countries, having had smaller ammonia plants before and thus possessing enough experienced technical staff and expertise, fared much better: when the big units followed the smaller plants more satisfactory records were achieved.

The oil crisis, the economic slump, the general shortage of foreign exchange, high transport costs together with the problems encountered with the jumbo plants created a completely new world market situation both in fertilizer products and fertilizer plants.

In this new situation, at the beginning of the 1980's, new trends have emerged in the nitrogen fertilizer industry and first of all in the manufacturing of ammonia:

- the price of energy increased dramatically, forcing the process owners to develop new, energy saving processes, however this technical innovation has been directed towards simpler, less sophisticated processes and equipment, easier to implement and less expensive to operate, and therefore more suitable for several developing countries

- while natural gas, remained the most suitable feedstock for ammonia production, long range perspectives as well as local conditions imposed a return to the old methods of synthesis gas production using other feedstocks, like coal or electrolytic hydrogen, where intensive R and D work is under way

- the experiences of several developing countries and the above factors focussed the attention on the mini-plant concept.

Process owners are engaged in an intensive development activity aimed at the reduction of both the investment costs and energy consumption. The individual situations may differ but the general tendency is simplification and some well known old methods are being renewed.

The main features of the new developments are:

- Reformer: low steam to carbon ratio, mild reforming conditions reduce considerably the heat load, and energy requirements, simplify construction, reduce investment. On the other hand, less waste heat is available and a higher methane leakage results. New energy recovery systems enter, e.g. the new version of the old saturator - dehumidifier loop.

- Gas purification: the higher methane content of the primary reformer exit gas opens two different routes: one separates the hydrogen in pure form (by PSA e.g.) and the other components are used as fuel gas in the primary reformer. Several purification steps are eliminated but pure nitrogen is needed. The second route uses excess air in the secondary reformer and after the normal purification cryogenic separation of the excess nitrogen either before or in the synthesis loop. For CO2 removal low-energy processes (mainly physical absorption) are used.

- Synthesis: New catalysts, converter constructions led to lower operating temperature, pressure and pressure drop, higher yield. Purge gas recovery systems are used to improve energy efficiency. Serious consideration is given to ammonia separation by absorption in water.

- Energy conservation: Gas turbine, absorption refrigeration, Rankine cycle are the most frequently encountered methods to make the process more efficient.

As a result of the above general tendencies, several new process schemes have been worked out. All of them are well suited in principle for implementation in the developing countries and particularly for mini-fertilizer plants, but only after having been proven on a commercial scale.

So the situation of the ammonia processes in the mid eighties is the following: several new flowsheets have been worked out for the big plants and some of them already implemented commercially. The spectacular achievements of the big plants have considerally reduced the interest of both contractors and investors for miniplants. The few miniplants realised in the last ten fifteen years have at least 50% higher energy consumption than the modern big plants and the specific investment cost is also much higher. The flowsheets are derived from these of the sixties for big plants or even older.

In the following chapter we worked out five possible flowsheets specifically adapted for miniplants. They are all composed from well known industrial process steps and differ from the big plants mainly in simplicity, ease of operation, less demanding process conditions and infrastructural requirements. Obviously, all of them have therefore somewhat higher specific energy consumption and investment cost than the most modern big plants, but the production cost difference is smaller than usual and allows competitive production, when the farm-gate costs in to remote locations are compared.

2. PROCESSES FOR MINIPLANTS

2.1 Process based on the ICI AMV Flowsheet (Fig.4.)

Natural gas from the battery limit is divided into two systems, feedstock and fuel. The feedstock is mixed with a small quantity of recycled hydrogen from the synthesis gas stream and the gas mixture is then heated in the Convection Section of the primary reformer prior to desulphurisation. The desulphurized gas is passed through the Feed Gas Saturator, before being sent to the primary reformer. Heat is supplied to the Saturator using 43 kg/cm2 steam and by interchange with the process gas between the two shift conversion vessels.

The temperature of the gas is raised to 480 deg C by heating it in the furnace convection zone. The heated mixture of natural gas, H_2 and steam is reformed in Primary Reformer to produce H_2, CO, CO_2, CH_4 and steam. Reforming occurs as the gas flows downwards through a number of heated catalyst-filled tubes made of nickel alloy, exiting at the bottom of the tubes. Mild reforming conditions (exit temperature below 800 C) lower the energy consumption and prolong tube life.

The heat required for the endothermic reforming reaction is provided by burning fuel in a number of burners.

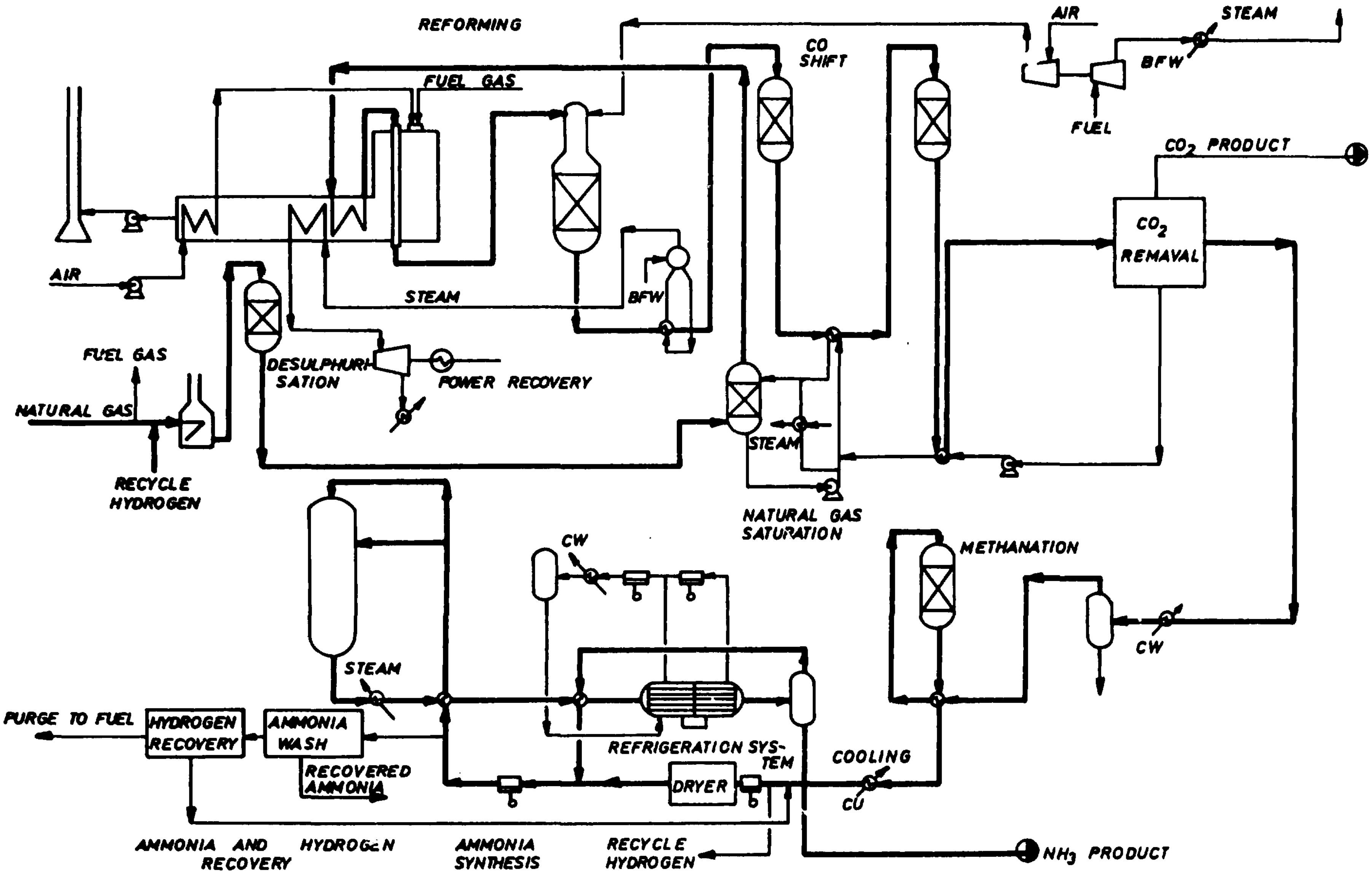


Fig. 4. AMMONIA - SYNTHESIS FROM NATURAL GAS. ICI - AMV.

Fuel is provided partly from waste gas from the hydrogen recovery unit and partly from fresh natural gas.

Combustion air is pre-heated by exchange with hot flue gas from the reformer.

Waste heat is recovered from the flue gas leaving the Primary Reformer radiant box by means of the following coils located in the convection section:

High Pressure Steam Superheater
Reactants Preheater
Combustion Air Heater

Reformed gas from the primary reformer enters the top section of the Secondary Reformer where it is blended with process air in a ceramic mixer section above the catalyst bed. The quantity of process air added is controlled to give the desired CH_4 content in the exit gas.

For that purpose excess air has to be introduced over the quantity needed for a H_2/N_2 ratio of 3:1 in the synthesis gas. This excess will raise the nitrogen content, which must be eliminated in the synthesis loop. Oxygen in the air reacts with some of the reformed gas. This generates a high temperature and provides the heat for further endothermic reforming of the residual natural gas as the gas passes down through the catalyst bed which it leaves at about 34 kg/cm2 and 930 deg C.

Filtered process air is delivered to the Secondary Reformer by a gas turbine driven compressor. The exhaust gas from the gas turbine is used for steam raising.

The secondary reformer is a refractory lined vessel with an external water jacket.

The reformed gas is cooled in Waste Heat Boiler before entering the CO shift section.

Cooled reformed gas enters the CO shift section in which the CO content of the gas is reduced to a low value by reaction with steam. The exothermic reaction takes place in two stages with heat removal between the stages. In the first reactor, the HT (High Temperature) Shift Converter, the bulk of the CO is converted at a high temperature over an iron oxide based catalyst. In the second reactor, the LT (Low Temperature) Shift Converter, the final CO conversion takes place over a copper based catalyst at a lower temperature. The CO Shift reaction equilibrium is favoured by low temperature.

The gas stream between the two shift reactors is cooled in the Saturated Water Heater.

In this example the CO2 removal is made by means of the Benfield Lo-Heat process. (Fig. 5.)

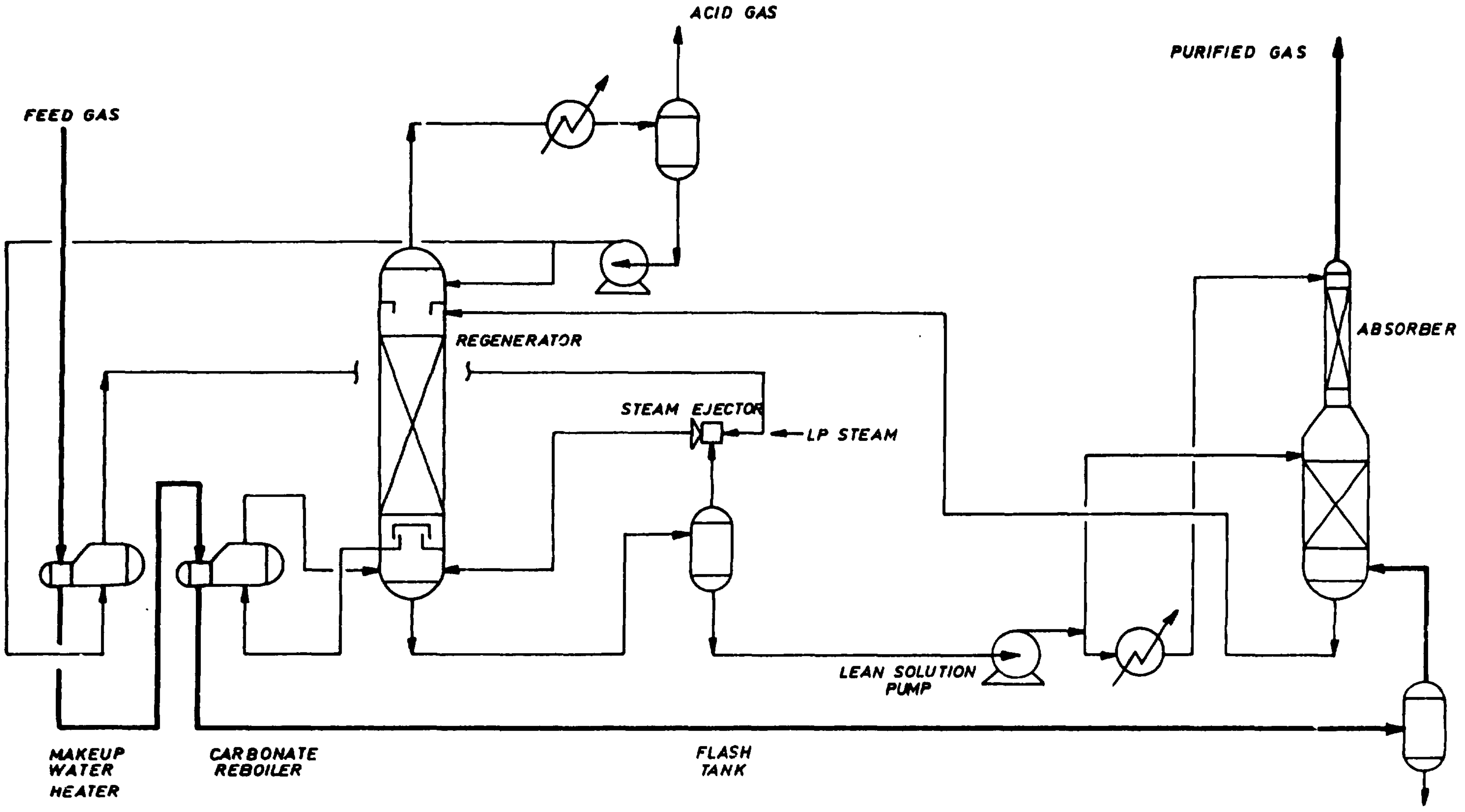


Fig.5. BENFIELD LO-HEAT PROCESS FOR CO_2 REMOVAL

This is an advanced activated hot carbonate process which derives part of its regeneration reboil heat from the make gas itself, and the rest from low pressure live steam. The process also allows most of the CO2 to be recovered in a condition suitable for urea manufacture. BASF MDEA uses a different arrangement. Any other process can be also used instead.

Gas from the CO shift section is cooled first in the Make-up Water Heater and then in the Benfield Reboiler. The cooled gas enters the base of the CO2 Absorber where it is washed by a counter current stream of carbonate solution. The column contains a number of packed beds. The top is fed with a cooled part of the lean carbonate solution, and the mid-point with the remaining semi-lean carbonate solution uncooled. The top of the lower is fitted with wash trays irrigated with BFW. Washed gas leaves the top with a CO2 content of 0.1 mol %. The gas passes to the methanation section.

Rich carbonate solution from the base of the absorption column flows to the top of the Carbonate Regenerator via a power recovery turbine. The solution flashes on entering the regenerator and then flows downwards over packed beds against a flow of stripping steam. Lean carbonate solution is flashed to a reduced pressure and pumped back from there to the Absorber via the motor and hydraulic turbine driven carbonate pumps.

Gas from the CO2 removal section is treated in a conventional methanating section. The gas is heated in the Methanator Interchanger before entering the Methanator. The gas passes over the catalyst and the CO and CO2 react exothermally with H2 to give CH4 and H2O. Residual carbon oxides at the outlet are less than 2 ppm v/v. The hot gas is cooled with feed gas in an exchanger. The gas is then further cooled by water, and then chilled before going to the Syngas Knock Out Drum, where condensed water is taken out and rejected. The gas is then dried in the Syngas Dryers. Recovered hydrogen from the loop is returned at this stage.

The dried gas goes to the Syngas Compressor. A small part of this syngas is recycled to the natural gas feed stream before hydrodesulphurisation. The rest is fed into the synthesis loop upstream of the NH3 Loop Circulation Compressor.

Ammonia is synthesised over an iron based catalyst at a pressure of about 85 bar. The synthesis reaction is exothermic and the catalyst is arranged in a number of stages with inter - stage cooling to keep the gas in the optimum temperature range.

Only a portion of the synthesis gas is converted to ammonia on each pass through the catalyst beds, so the unconverted gas is separated by cooling and condensation of the ammonia as a liquid, and recycled to the catalyst together with the fresh make-up gas.

Fresh make-up gas enters the synthesis loop at the suction side of the steam turbine-driven Loop Circulator. The gas discharged is heated in the Hot Interchanger by exchange with converter effluent gas and then flows to the NH3 Converter. A portion of the feed gas enters as quench gas for temperature control of the inlet to the second of the catalyst beds. The remainder of the feed gas is then heated to reaction temperature in an internal heat exchanger by exchange with hot gas leaving the second catalyst bed. The preheated feed gas then enters the top of the first catalyst bed. Hot gas leaving the first bed is quenched with cool feed gas and enters the top of the second catalyst bed. Hot gas leaving the second bed is cooled as it passes through the external exchanger. From the external exchanger the gas passes to the third catalyst bed which is an adiabatic reactor without quench gas addition.

Hot gas leaves the converter and is cooled successively by exchange with BFW, converter feed gas, recycle synthesis gas, and boiling ammonia refrigerant in the chiller. Liquid ammonia is condensed from the gas and separated. Unconverted gas is recycled to the circulator via the interchanger.

Refrigeration for the chillers is provided by an indirect ammonia refrigerant cycle. The cycle has two evaporation pressures and is driven by a two - stage reciprocating compressor. Ammonia vapour is condensed by cooling water and collected in the Refrigeration Receiver.

2.2 Process based on PSA gas purification (Fig.6.)

This scheme represents the technology of no particular licensor and should be generally available from most experienced contractors. However an ammonia converter capable of achieving the desired performance is necessary. It is anticipated that Ammonia Casale, Uhde, Topsoe and Kellogg converters can achieve this performance. PSA technology is available from Union Carbide Corporation and Linde AG.

Natural Gas from the battery limit is heated to 400 deg C in a heat exchanger. The gas is then desulphurised using ZoO before it is mixed with steam to give a steam to carbon molar ratio of 3:1. The mixture is heated to about 500 deg C in the Reactants Heater located in the convection zone of the primary reformer before entering the inlet system of the Primary Reformer. The reformed gas leaves the tubes at 20 kg/cm2.a. and 850 deg C.

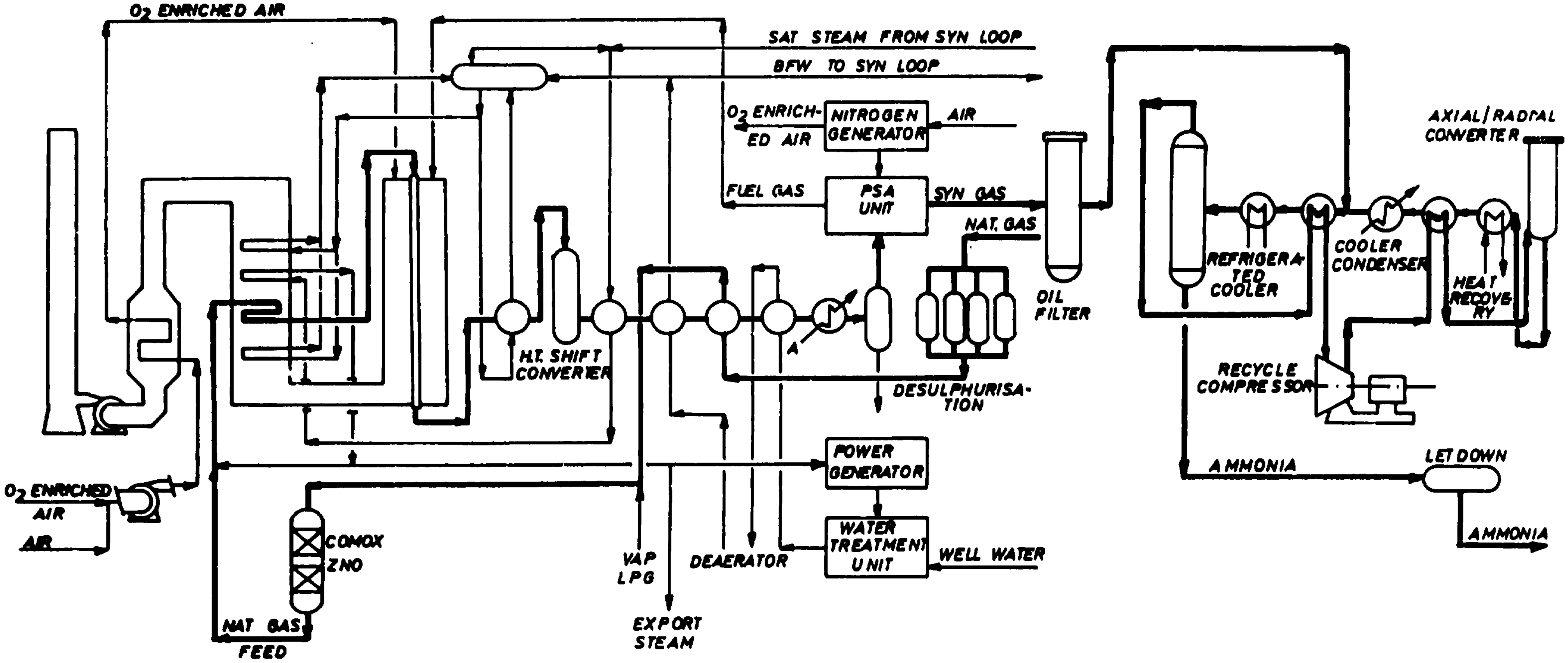


Fig.6. AMMONIA SYNTHESIS FROM NATURAL GAS. PSA BASED PROCESS

The reforming furnace is fired with PSA offgas and additional natural gas. In order to make sufficient steam for the plant to be self contained in power there is additional firing of natural gas in the convection section of the furnace. Heat from this additional firing and from the gas leaving the radiant section is recovered in the following coils:

Steam Superheater
Reactant Heater
ConZone BFW Heater
Natural Gas Heater
Combustion Air Heater
ConZone Boiler

The flue gas is discharged to the Flue Gas Stack by the Flue Gas Fan. Combustion air is provided for all burners by the Combustion Air Fan.

Hot gas from the reformer is cooled to 350 deg C in the Primary Make Gas Boiler when it enters the HT CO Shift Reactor. In this reactor about 70% of the carbon monoxide is converted in a single stage to CO2 and more H2. After the reactor heat is recovered in the following exchangers:

Steam Superheater
BFW Heater
Natural gas Heater

The cold gas enters the PSA unit where 87 - 88% of the incoming hydrogen is recovered in a very pure form. All the water vapour, carbon dioxide and methane are removed. Any helium present will pass through, as will part of the argon. A few ppm of carbon monoxide will slip but this can be controlled to under 3ppm when a few ppm of nitrogen may also slip through. The separated gases are let down into the surge and mixing system which operates at about 1.3 Kg/cm2.a. The mixed gas is used as fuel.

The process operates cyclically over a period of a few minutes. At 350 ton/day NH3 capacity 8 or 10 beds will be used and the hydrogen product is available at a steady flowrate. The surge and mixing system evens out the flow and composition of the fuel gas to a calorific value variation of less than 2 1/2% over the cycle.

Nitrogen is produced in an air separation unit of standard design. The Air Compressor is likely to be a centrifugal type with electric motor drive. The nitrogen is compressed in a reciprocating Nitrogen Compressor to about 27 Kg/cm2.a. at which pressure it joins up with the hydrogen from the PSA unit to make a 3:1 hydrogen:nitrogen mixture.

The nitrogen plant must make a pure gas containing less than 10 ppm v/v of oxygen in the nitrogen. A small amount of liquid nitrogen storage is provided to speed restart of the unit after any warm-up that may occur.

The hydrogen-nitrogen mixture is compressed to 120 Kg/cm2 using 2 reciprocating compressors. The compressed gas is passed through an oil filter, as is necessary, and then mixed before the cold exchanger.

The loop gas is then successively cooled by:

Ammonia Loop Cold Interchanger
Ammonia Chillers

After the second stage of chilling liquid ammonia at 4 Deg C is separated from the gas in the Ammonia Catchpot. The unreacted gas then returns via the Ammonia Loop Cold Interchanger to the Circulator.

The combined gas is compressed by the Circulator to 125 kg/cm2. The gas is then heated to about 250 deg C by the Loop Interchanger. It enters the Ammonia Converter where about 23% of the hydrogen is converted to ammonia. The gas leaves the converter at about 430 deg C and enters the Ammonia Loop Boiler where it raises 48 kg/cm2 steam. It is then further cooled in the

Ammonia Loop Interchanger
Product Gas Cooler

and receives the make-up gas.

Most of the inerts present in the feed gas will dissolve in the ammonia product and be removed from the loop in this way. Others will build up and it may be necessary to purge either continuously or intermittently.

The refrigeration for the ammonia loop is done in two stages of chilling with a two-stage reciprocating Refrigeration Compressor.

2.3 Coal based flowsheet (Gasification with oxygen) (Fig.7.)

This scheme includes licensed technology from Texaco Development Corporation of the USA, Selexol of Norton of the USA, and may use CO shift technology licensed by EXXON. Some designers may use ammonia loop or reactor technology subject to license agreements. For this description we have adopted an ammonia loop with an ammonia reactor designed by Ammonia Casale of Switzerland. Large PSA units are available from Union Carbide Corporation or Linde AG. The Air Separation (ASU) plant can be obtained from a vendor with a proven track record in such plants.

Coal will be received at site into stockpiles of up to 40,000 t if some constance from the mine. If the mine is alongside then the stockpile can be much smaller.

Care needs to be taken with the stockpile to prevent spontaneous combustion taking place. Recovery from the stockpile will be by front loader. This will tip the coal into a conveyor hopper system feeding the coal mills.

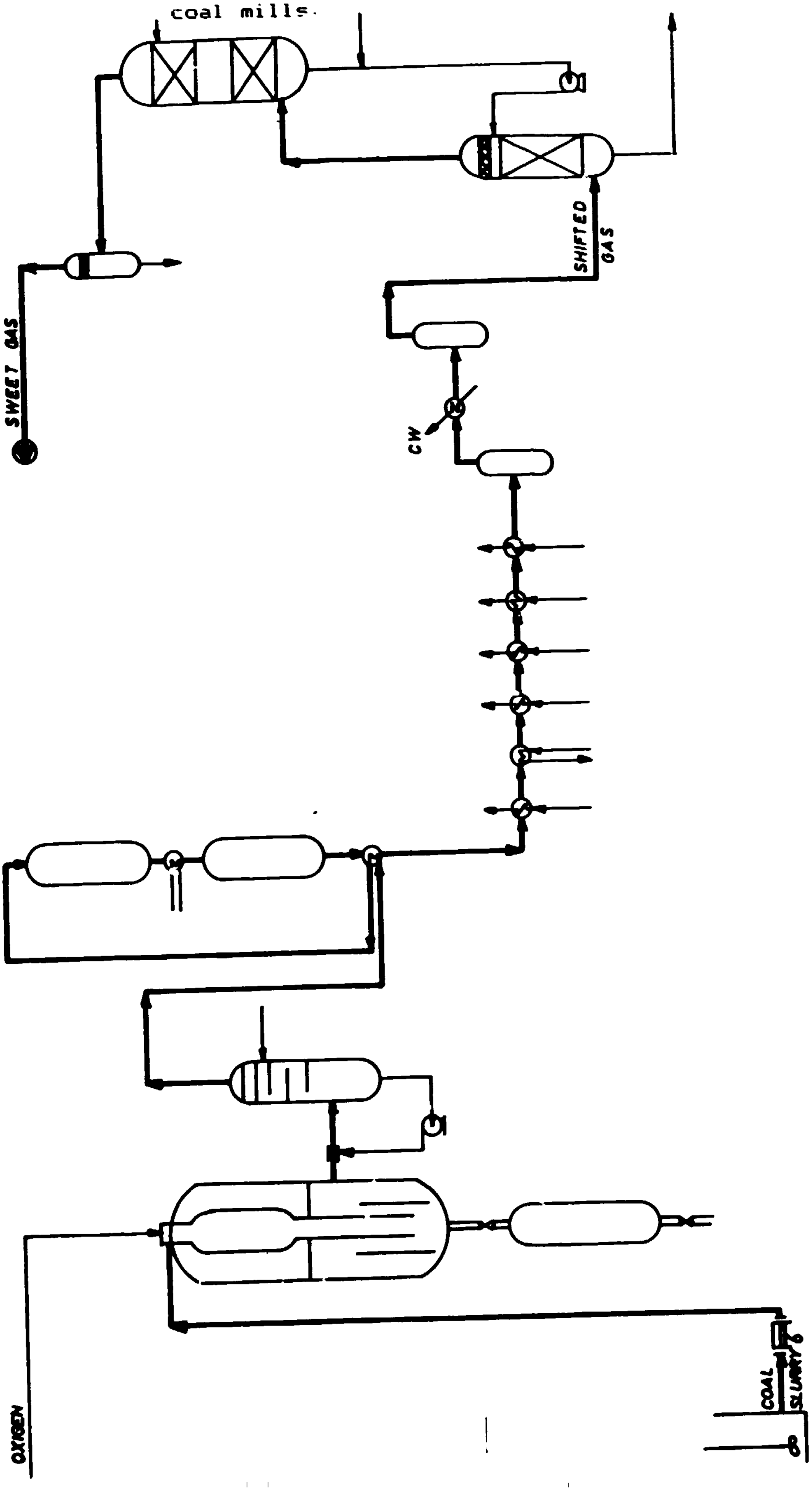


Fig. 7.a COAL BASED AMMONIA SYNTHESIS. GASIFICATION

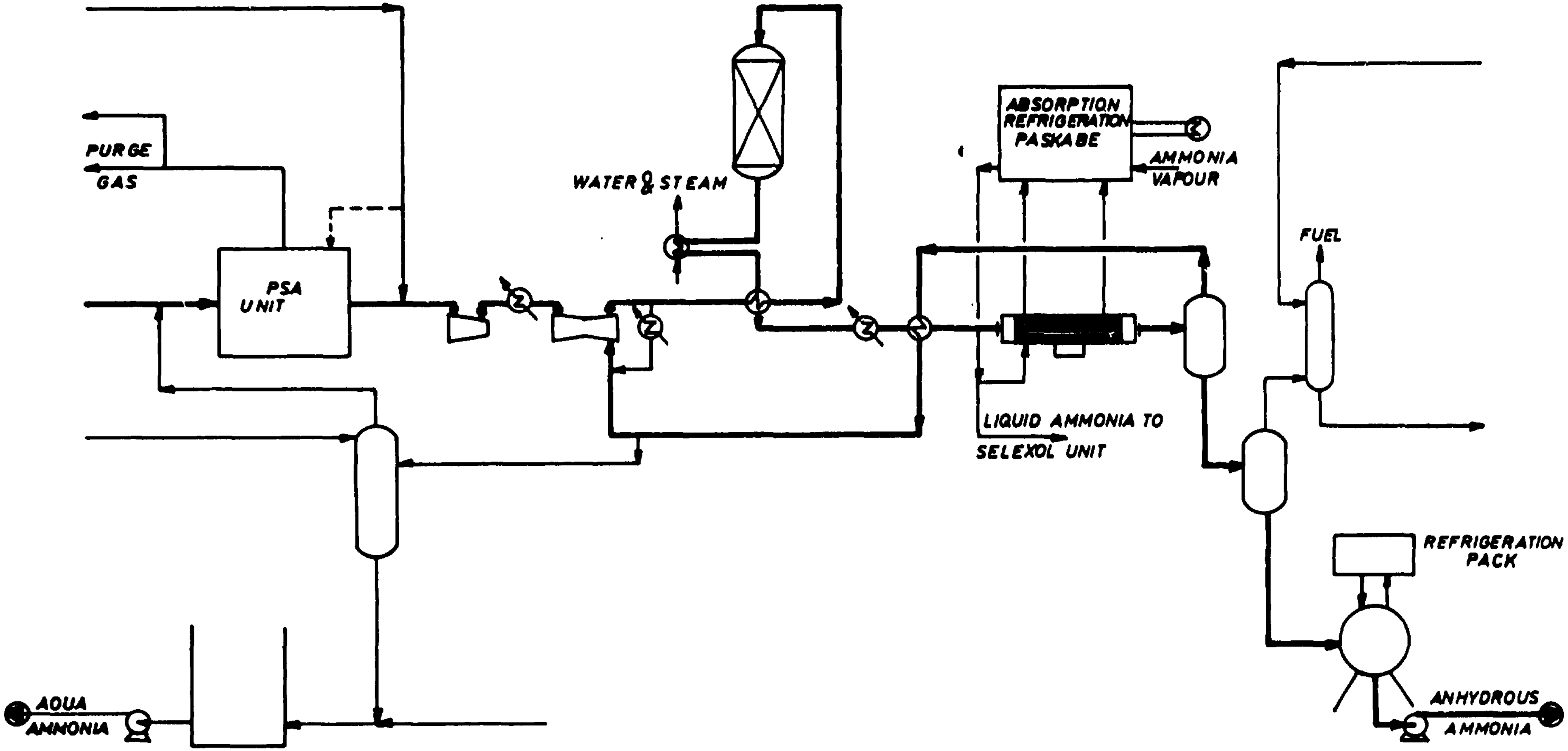


Fig. 7. b COAL BASED AMMONIA SYNTHESIS . SYNTHESIS

Water for milling is largely that used to wash the slag through the lock hopper plus a smaller quantity from the blackwater stripper overheads which will contain some suspended sulphur.

The milled coal goes to a hold tank from where it is pumped via a screening system to one of three checking tanks where the concentration is checked and adjusted as necessary. Certain special chemicals are added at the milling stage to enable a high solids concentration to be obtained.

Coal from Final Slurry Tanks is pumped by the reciprocating Slurry Pump into the burner of the Gasifier. Oxygen from the Oxygen Plant is compressed by reciprocating Oxygen Compressor into the same burner assembly in the gasifier. The coal slurry and the oxygen combust at 1350-1400 deg C to form a synthesis gas rich in carbon monoxide and molten ash and a small quantity of unburnt carbon. These are all quenched in the water bath at the base of the gasifier. The gas evaporates a very large quantity of water. The molten ash becomes round glassy granules and some of the unburnt carbon is trapped in the water.

The saturated gas leaves the gasifier above the water level and is scrubbed with more water in the venturi scrubber to remove all the suspended carbon. The water is knocked out in the Gas Scrubber which includes a clean water wash stage to remove all entrained droplets of dirty water. Water from the base is pumped into the venturi scrubber and the base of the gasifier.

Slag from the base of the gasifier is periodically discharged into the Slag Lock Hopper located just below the gasifier. The slag is then discharged through a screen into the Slag Settler. Most of the slag runs off the screen into a container for disposal. The final slag settles and is removed into the container. The slag-free liquid is then re-used to slurry the fresh coal at the mill.

Water containing suspended carbon known as 'blackwater' is continually let down from the gasifier through the Gasifier Water Exchanger to the Blackwater Flash Tank in the effluent treatment area.

After particulate removal the saturated gas is warmed up to 260 deg C by the effluent from the final CO Shift Reactor in the Shift Feed Heater. In the first CO Shift Reactor about 90% of the gas is converted using a cobalt-molybdenum oxide catalyst. The very hot gas is cooled by raising steam at 45 kg/cm2 in the Intershift Boiler. In the second reactor the CO is reduced to about 1%. After the second reactor are several heat exchangers which cool the gas to 40 deg C:

Shift Feed Heater
Shift BFW Heater
Process Condensate Heater

Absorption Refrigeration Reboiler
LP Boiler

Returned Steam Condensate Heater
Raw Water Heater
Shift Effluent Cooler

All apart from the last of these are making use of the heat for various essential duties.

The feed gas enters the Selexol plant at 40 deg C 32 Kg/cm2 and passes into the H2S Absorber. Here, H2S is preferentially absorbed using CO2-saturated solvent from the CO2 Absorber entering at the top of the column. As the quantity of H2S is small, the temperature change is due mainly to the cooling of the feed gas and resultant desorption of CO2 from the solvent. The gas, now containing less than 1 ppmv H2S, enters the base of the CO2 Absorber where it is contacted with cold lean Selexol at 0 deg C entering at the top. The gas leaving the top of the absorber contains about 6% of CO2.

The CO2-rich solvent is expanded through the Selexol Turbine, and flashed in the LP CO2 Selexol Flash Drum to produce the CO2 product gas. The solution is returned to the absorber via the Chiller.

The H2S-rich solvent passes to the LP H2S Selexol Flash Drum then passes through the Selexol Interchanger to the H2S Regenerator. The regenerator column is in two separate sections, the top section being a flash unit and the bottom section a stripping unit.

With about 6% CO2 in the gas the PSA unit passes about 89% of the hydrogen as a very pure gas. The gas is mixed with pure nitrogen from the oxygen plant. This nitrogen is compressed from 1.2 Kg/cm2.a to 30 Kg/cm2.a by reciprocating Nitrogen Compressors. The mixed gas is then compressed to 135 Kg/cm2 by a pair of reciprocating compressors.

The loop pressure has been selected to fit the available heat for the regeneration of the absorption refrigeration to the chiller duties for both the loop and the Selexol unit.

The synthesis section is practically identical with that of the two previous flowsheets.

The chillers are part of an absorption refrigeration system. This eliminates the need for another 1.1 MW compressor motor and makes the best use of a considerable quantity of heat available in the gas production system. It is comprised of simple pumps, columns and heat exchangers mainly constructed of carbon steel.

2.4 Coal gasification with air

Autothermal partial combustion was always performed with either pure oxygen (95-98 %) or enriched air (around 50 % O2), although the reaction could have been carried out with normal air also. In this case, however, the nitrogen introduced with the necessary quantity of air could be far in excess of the 3:1 hydrogen-nitrogen ratio needed in the final syngas. The new ideas described above (working with excess air in the secondary reformer and eliminate the surplus of nitrogen by partial condensation) could be applied also to the partial oxidation of coal, relieving this process from the necessity to build and operate an air separation plant. This process has not yet been tested, as a complet line, but the individual steps are commercially proven. The flowsheet proposed by Foster-Wheeler is very simple (Fig. 8.): entrained bed non-catalytic partial oxydation under pressure with preheated air and steam, heat recovery, removal, shift conversion, desulphurisation, CO2 removal (eventually in one step), drying (molecular sieve) and condensation of the excess nitrogen could deliver a pure syngas to the ammonia synthesis. The cold produced by the expansion of the condensed nitrogen to atmospheric pressure could cover the cooling energy needed for this condensation. The process could be applied for other feedstocks too.

The process essentially comprises the following steps:

Partial combustion of the feedstock (oil or coal) with preheated air and steam, to generate a raw synthesis gas containing chiefly hydrogen, carbon monoxide and nitrogen, typically at 1,300o C and 70 bar. Established partial oxidation and gasification processes may, according to the main licensors, be used in this way and in fact such processes have been operated successfully with air as the oxidant.

The partial combustion is advantageously followed by a waste heat boiler and a soot and ash removal, according to the systems of the gasifier licensors.

Carbon monoxide shift, using whichever of the commercially available catalysts is appropriate. The selection of catalysts would be made mainly according to the degree of conversion desired and the sulphur content of the gas, as in normal practice.

For ease of display, Fig.8. shows only one shift converter. Although the nitrogen from the gasification air will greatly increase the gas flow on a dry basis in comparison with an oxygen-based partial oxidation plant, the shift catalyst volume need not be significantly increased and the number of stages may sometimes be decreased. This is because:

* the nitrogen present acts as a thermal

reservoir, allowing the steam to the shift to be reduced to around the minimum the catalyst can accept; hence the total shift gas flow (including steam) is still reasonable; and

- with the provision for recovery of the calorific and pressure energy of the waste nitrogen steam as described below, there is apparently less economic advantage in a high degree of carbon monoxide shift than in normal practice.

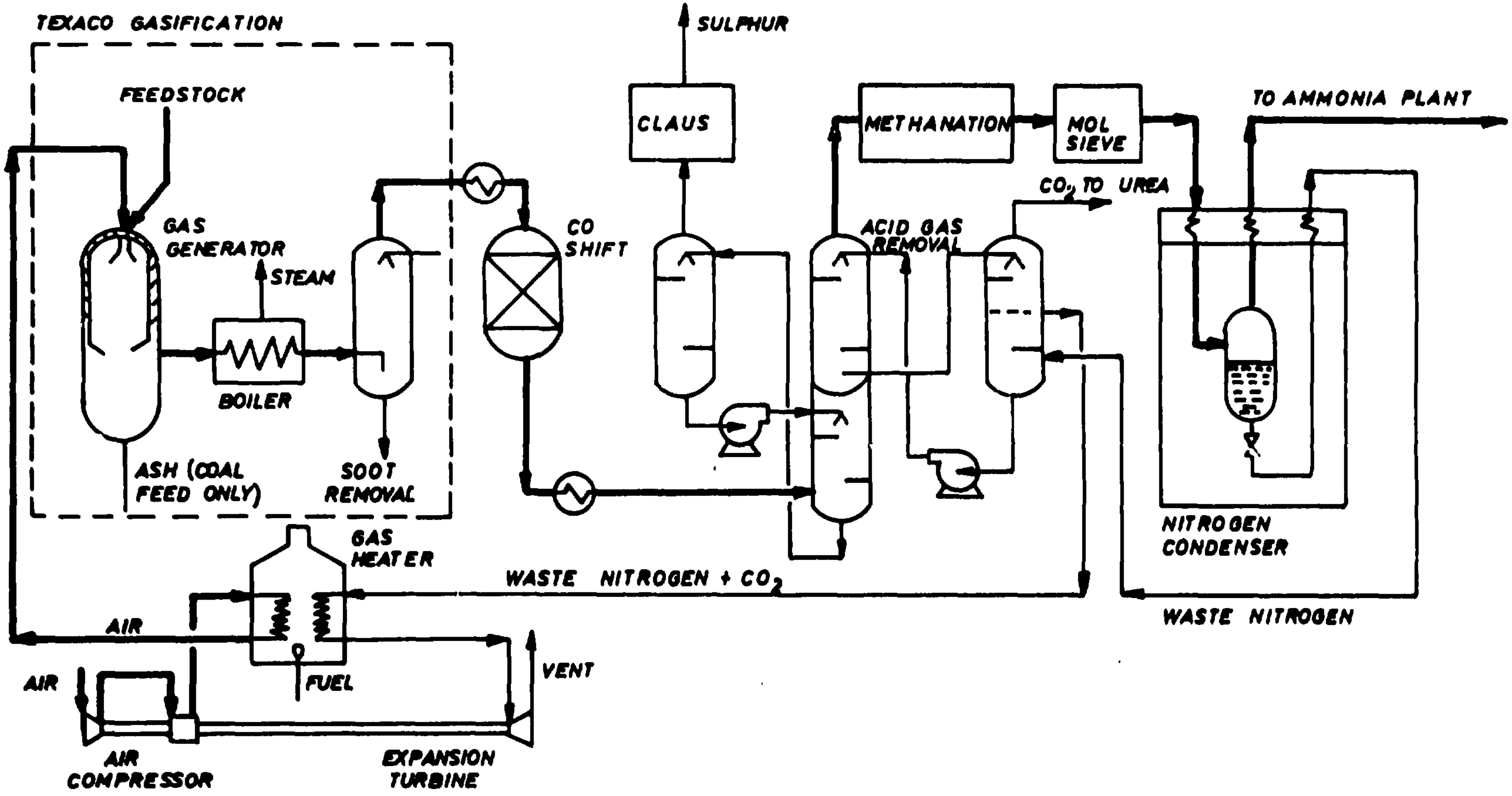


Fig.8. AMMONIA SYNTHESIS GAS PLANT BASED ON AIR GASIFICATION

Acid gas removal, by an established process, to remove the carbon dioxide present and the hydrogen sulphide (if any). Of course, hydrogen sulphide could be removed before the shift if preferred.

Cryogenic nitrogen condensation, to reduce the nitrogen content of the gas to 25 vol-% for ammonia synthesis.

This separation of nitrogen from hydrogen is very much easier than the separation of nitrogen from oxygen because of the much greater difference in boiling points of the two gases and because only 70-80 % of the nitrogen need be removed.

The separation can be made in a very simple cryogenic plant comprising only heat exchangers and separators. At the characteristic operating pressure (50-70 bar), this separation can be thermally sustained by the Joule-Thomson refrigerative effect alone, without need for mechanical expanders or external refrigeration cycles. Many similar, although smaller, 'cold boxes' are in use for the recovery of hydrogen from ammonia plant purge gases.

A molecular sieve cleaning stage is provided upstream of the cold box to remove traces of water vapour, carbon dioxide and other materials that would otherwise form a solid rim inside the cryogenic equipment.

When preceded by methanation to remove traces of carbon oxides, this cryogenic operation produces a completely dry, high-quality ammonia synthesis gas.

Alternatively, the final purification and nitrogen condensation can be performed in the 'cold box' by the incorporation of a nitrogen wash column, the wash nitrogen being generated from the synthesis gas itself. In this alternative, no methanation is needed.

Waste gas system. The waste nitrogen leaves the cold box at a pressure around 10 bar. It is heated and then expanded to atmospheric pressure in a turbine, thus providing a high proportion of the power needed for the air compressor.

2.5 Ammonia by water electrolysis (Fig.9.)

The synthesis of ammonia consists of making hydrogen and nitrogen react, in the ratio 3:1, under high pressure and temperature in the presence of a catalyst:

$$3H_2 + N_2 \longrightarrow 2NH_3$$

The production of 1 tonne of ammonia requires 1,970 Nm3 of hydrogen and 657 Nm3 of nitrogen. The nitrogen required is based upon air liquefaction and separation:

Standard plants with any capacity can be procured from several specialized firms for air separation. Water electrolysis plants are also of standard design in the capacity range required for mini plants. Fig.10. shows the flowsheet of an electrolyser plant.

Hydrogen from the water electrolysis plant and nitrogen from the air separation plant pass to separate gas holders, which provide a buffer capacity and stabilize the gas pressure. Compared with synthesis gas generated from a hydrocarbon feedstock, the water electrolysis gives an extremely pure gas, containing only a very small amount of oxygen (0.1-0.2%), which, however, has to be removed, as oxygen is a poison to the ammonia converter catalyst. The only purification needed is therefore oxygen removal, which is done by means of catalytic combustion. A small amount of the hydrogen reacts with the oxygen present and a corresponding amount of water is produced. The purification takes place immediately after the mixing of hydrogen and nitrogen and the purified mixed gas (make-up gas) passes to a gas holder serving as a buffer for the ammonia synthesis section.

The synthesis gas is next compressed to the pressure of the ammonia synthesis loop (normally between 100 and 280 bars) and ammonia is synthesized in a synthesis loop based on the same principles as are used in most other ammonia plants.

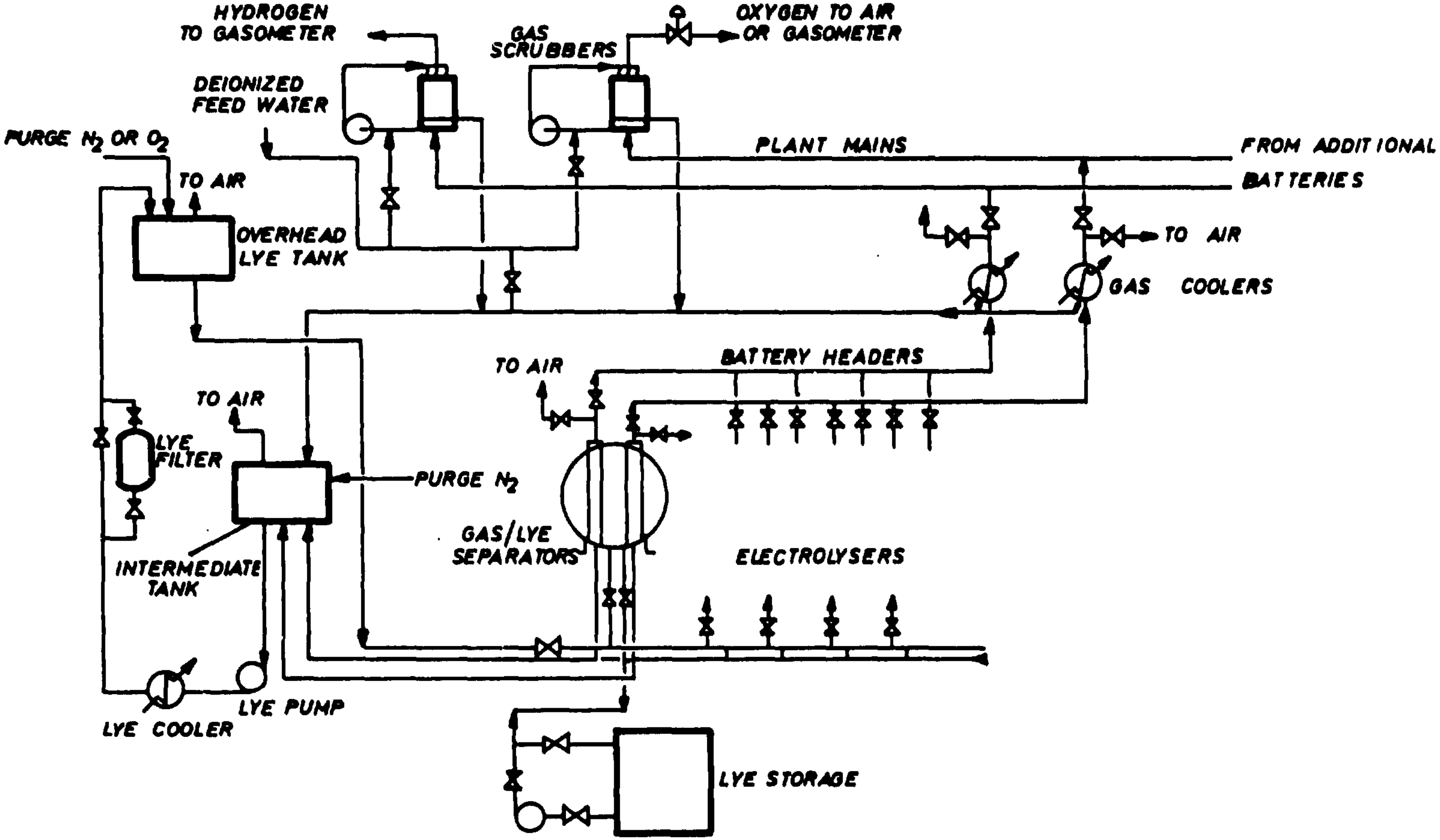


Fig. 9. FLOW SHEET OF ELECTROLYSER PLANT

The table below shows the main technical parameters of such a plant for three different capacities.

Technical Specifications for Three Sizes of Ammonia Plant Based on Water Electrolysis

NH_3 production	100 t/d (4.17t/h)	300 t/d (12.5t/h)	500 t/d (20.83t/h)
Electrolysis plant			
H_2 requirement	8200 Nm^3/h	24600 Nm^3/h	41000 Nm^3/h
O_2 production (by-product)	4110 Nm^3/h	12300 Nm^3/h	20550 Nm^3/h
Number of electrolysers (approx.)	28	72	140
Power requirements for H_2 (4.3 kWh/Nm^3)	36 MW(DC)	108 MW(DC)	180 MW(DC)
Air fractionation unit			
N_2 requirement (design)	3500 Nm^3/h	10000 Nm^3/h	17500 Nm^3/h
Power requirements for air (design)	0.5 MW	1.4 MW	2.35 MW
NH_3 synthesis section			
Power requirements (design) for compressor	3 MW	9 MW	15 MW

Total power consumption:
10 MWh per metric ton NH_3.

Comparing the production of ammonia from hydrocarbon feedstock this route is much simpler and does not contain any complicated process steps operating at high temperatures and pressures - apart from the ammonia synthesis unit, which is the same for all processes. Nevertheless, the investment costs are high (about 30-50% above those for a natural gas based plant). Recent development works on new electrolytic cell types (membrane) are promising: it seems reasonable to expect more competitive investment costs in a few years'time. On the other hand the very high energy consumption: 10 MWh/t NH3 will remain in the same region: more than 20-30% saving seems impossible to achieve even in the far future. This, calculated with a factor of 2700 kcal/kWh means 27 Gcal/t NH3, roughly four times more than the value of 7-7.5 Gcal/t NH3 usual for nowadays plants.

Electrolysis under pressure, another field of R and D activity would save the energy needed for the compression of hydrogen, but this is less than 10% of the total energy consumption.

In any case, this process will never be competitive if power is produced in a thermal power plant. Where hydroelectric power is available at low cost, due consideration should be given to this alternative. The calculations showed, that only if electric power is available for 3-14 dollar/MWh can this method be competitive.

2.6 Urea production

The industrial-scale manufacture of urea from ammonia and carbon dioxid involves two separate reactions. Initially the two reactants combine to form ammonium carbamate, from which a molecule of water is then eliminated to give urea.

$$CO_2 + 2NH_3 <\!\!> NH_2COONH_4$$

$$NH_2COONH_4 <\!\!> CO(NH_2)_2 + H_2O$$

The reaction is carried out in a reactor operated under pressure - at least 100 bar - and at an elevated temperature in excess of 160o C.

In a typical reactor only about 60-70 % of the stoichiometric mixture of ammonia and carbon dioxid will be converted to urea. It is necessary to separate product urea from unreacted carbamate in the solution leaving the reactor. This is done by decomposing ammonium carbamate to carbon dioxid and ammonia. The main differences between the different processes lies in the method used for this decomposition. The old once-through and partial recycling processes are practically not used any more, only different versions of the total recycle process and mainly the stripping methods can be considered for new plants.

In the total recycle stripping processes NH_3 or CO_2 or both can be used as stripping agent.

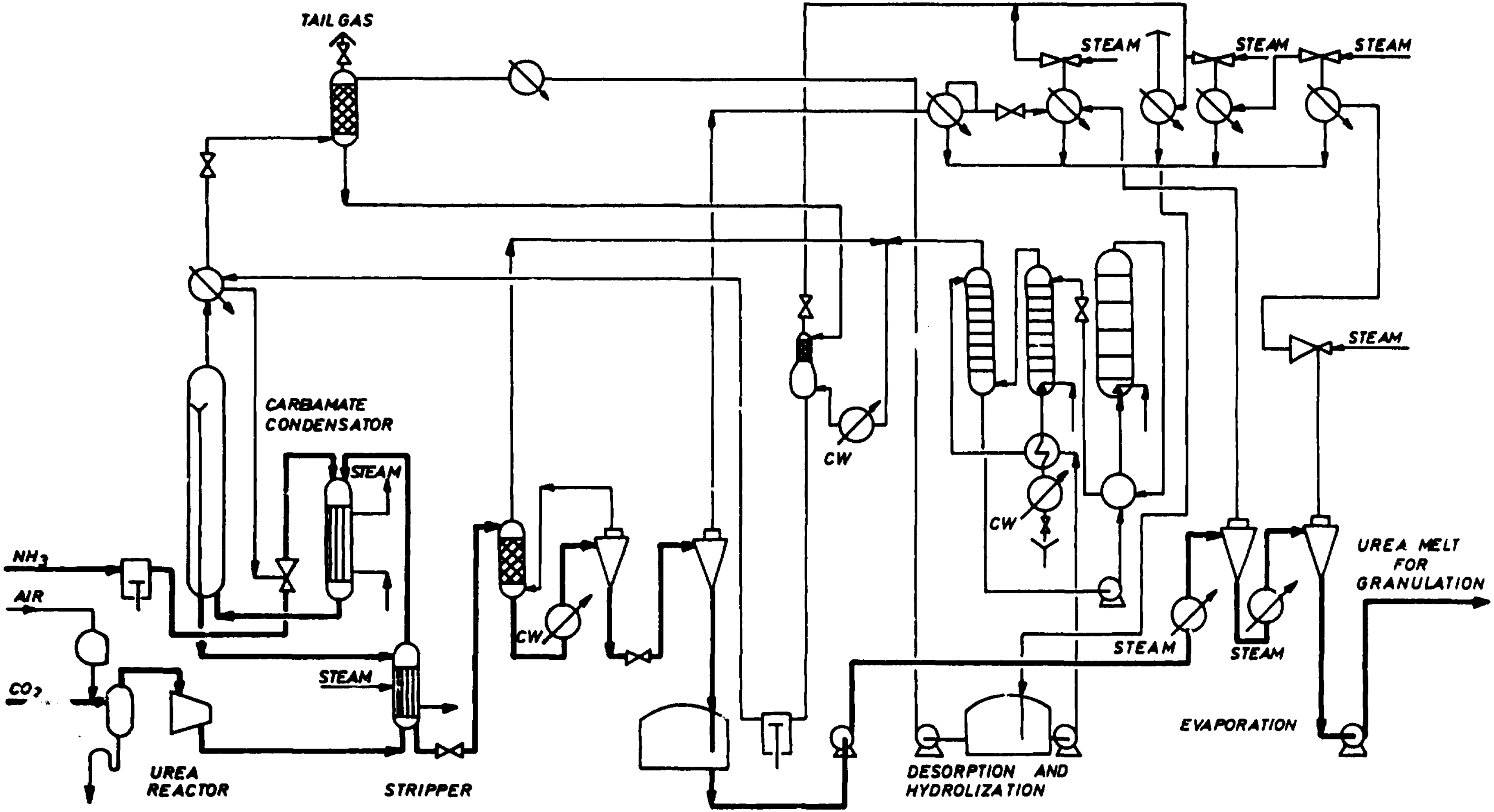


Fig. 10. UREA SYNTHESIS AND EVAPORATION

Stripping at syntheses pressure reduces considerably the onerous fluid pumping operations and greatly improves the heat recovery. Stripping with CO2 was first used, followed by ammonia and two step ammonia and CO2 stripping methods.

Practical plant capacity are between 300-1700 tpd and the technological processes are in this capacity range essentially the same.

There is practically no difference in the process whether the size is big or small. All the known processes (Stamicarbor, Snam Progetti, Toyo, Montecatini etc) can be used without any changes for mini plants. The stripping is today a general feature adopted in nearly all processes. Besides the stripping with CO2 shown in the flowsheet below, ammo a stripping and double stripping using both reactants is also used, but these features do not change substantially the characteristics of the process.

The Fig. 10. shows a typical total recycling stripping process. At the synthesis condition (T = 180o C, P : 150 bar) the carbamate producing reaction occurs rapidly and goes to completion. The urea reaction occurs slowly.

From the reactor the mixture flows to the steam heated stripper, where, as stripping medium CO2 is introduced to decompose the unreacted carbamate and the gases are fed to the carbamate condensers, while the solution flows to the rectification tower and a heater where at 3-4 bar pressure the remainder of the carbamate is decomposed. The gases condensed will be recycled, while the urea solution will be evaporated and the water free melt prilled in the prilling tower.

The urea plants have to be located at the same site as a correspondingly sized or larger ammonia plant since the ammonia plant supplies not only ammonia but also the high purity carbon dioxide.

Based on this new idea, complex ammonia-urea flowsheets were developed and tested in pilot plants where the CO2 removal is realised using an ammonia solution in water and the resulting liquor is introduced directly in the urea production. Since it does not seem advisable to propose for developing countries commercially unproven processes, this method was omitted.

2.7 Granulation, Bulk blending and Bagging

Granulation

As we have seen, the chemical processes used for the production of nitrogeneus fertilizers deliver a melt (sometimes with suspended solids) and therefore an adequate finishing step is necessary to arrive at a solid product needed in most agricultural uses. The phosphate fertilizers on the contrary can be produced in powder or in slurry form. The slurry must be processed to arrive at a solid product, which will be granuled and the powder can also be transformed to this form. For that reason, since the 1950's the solidification and cooling of the melts produced in the nitrogen industry was considered as an integral part of the processes involved and prilling gained nearly universal acceptance for this purpose.

The word granulation, was reserved to the phosphate field where drying was the basic operation involved, to eliminate the water content carried in the slurry or added in powder granulation.

Pan-, drum- and pug-mill type granulators were used for this purposes.

In the last few years drum and pan granulation technics gained more and more acceptance in the nitrogen industry both for urea and ammonium nitrate and several new methods were developed which are equally suitable for both type of products. NPK fertilizers -equally new products- can be made in similar equipment from the same phosphate, ammonia and nitric acid. All this seemed to justify a common treatment of all granulation technics in a single chapter.

Prilling

The wide use of prilling is due to the advantages of this system, in particular the great daily production capacity of the equipment, low labour and operating costs.

Prilling is the production of a granular solid by allowing molten droplets to fall through a gaseous cooling medium. Non-viscous homogeneous materials with well-defined melting points, such as pure ammonium nitrate or urea, are very easily prilled.

To obtain hard and non-porous prills, the water content must be reduced below 0.5% otherwise a porous, low-density product results which is troublesome in storage. Jets of free-falling molten materials are broken into droplets by the air. The droplets begin to solidify as they fall through the cooling medium. The crystallization starts at the surface and progresses gradually to the inside.

The prilling device and melt temperature must be carefully controlled. The retention time in the prilling tower is also an important factor. The proper design of the tower height and cooling air flow are essential to obtain completely hard prills at the bottom. The still hot prills arriving at the bottom must be collected and transported to the finishing, and cooling may also be required.

The prilling tower itself is a structure supporting the prilling equipment placed on the top, together with fans providing the necessary air stream. The main characteristics of the tower are the height determining the retention time, the cross-section fixing the capacity and the air stream.

The not completely satisfactory granulometric composition is the drawback of this system. Prills are relatively small; in practice most are around 1 mm and only a small proportion reach the 2 mm mark.

Prilling is very advantageous for big capacities. At the lower end specific investment costs begin to rise to such an extent that other granulation technics become more advantageous.

Pan Granulator

This principle, which had been widely used in the pharmaceutical industry was developed for superphosphate granulation and was extensively used in the phosphate industry.

Number of companies made considerable improvements, and the process lends itself for making granular ammonium nitrate and urea. Granulation is accomplished by spraying hot concentrated melt of fertilizer salts onto a cascading bed of recycle material in a pan granulator.

The granulated product is cooled and is sized in conventional equipment. The oversize fraction from the screen is crushed and returned with the undersize fraction for use as recycle material. The correct size product is treated with an appropriate conditioning agent and sent to the store.

Critical features of the pan granulator for best operation include slope, rotational speed, location of spray, concentration and the amount, particle size and temperature of the recycled material.

Drum Granulator

The classic drum granulator consists of a slightly inclined rotary cylinder with retaining rings at each end and with appropriate internal structures.

The basic materials must be well mixed before entering the drum, which serves only to form the granules; these are rounded at the bottom of the drum by their contact with each other. The speed of rotation of a drum granulator must be slow enough for the granules not to be carried around by centrifugal force since the principle of this system is that the granules should move relative to the drum.

Depending on the residence time required, drum granulators may be mounted with a downward slope up to 3 o.

Drying and cooling are both needed after granulation, so a typical granulator train consits of a drum with two sections, the first serving for mixing and chemical reaction, the second for granulation, followed by a drier and cooler, both of drum type.

A basic feature of the drum granulator system is the great amount of recycle material. Not only over- and undersize material, but a given part of the product must be recycled depending on the water content of the input materials for optimum granulation drying conditions.

This process can be combined with ammoniation of phosphates (SSP or TSP) NPK fertilizer manufacture etc., as well as for granulation of SSP or TSP powder.

Spherodizer

Hot Spherodizer Process

The process was conceived in an effort to simplify the processes in use for the granulation of complex fertilizers and to improve product quality. The major innovation of the spherodizer process consisted of combining granulation and drying into a single processing operation.

By this process, the conversion of liquid slurry to uniform granules is accomplished by spraying the slurry under pressure through nozzles onto a dense curtain of recycled material cascading from lifters in a rotating cylindrical drum. A steam of heated air flows through the drum co-currently with the solid recycle and the sprayed slurry, coming into intimate contact with the particles to be dried. As droplets of slurry hit the recycled granules, water is flashed off, resulting in new onion skin-like layers of material around each of the solid particles every time they are cascaded.

The remainder of the flowsheet is conventional, with screens to separate the product size, crushers to reduce oversize, and elevators and conveyors to transfer the solid materials.

This process was identified as the "Hot Spherodizer Process" to distinguish it from the one using only cooling air for the granulation of melts, which came to be known as the "Cold Spherodizer Process".

Cold Spherodizer Process

The cold spherodizer process is used in the granulation of ammonium nitrate and urea. A substantially anhydrous melt of either ammonium nitrate or urea is sprayed inside a rotating drum onto a rolling bed of solid particles. As the particles roll, they are repeatedly coated with thin layers of liquid melt, which solidify to give the granule an onion-skin structure.

Air flows through the granulation drum in countercurrent to the granules, removing part of the heat of crystallization of the melt, as well as the fine dust. From the drum it is drawn by an exhauster into a wet scrubber before being discharged to the atmosphere.

Pugmill

Mixing and granulation in the same equipment is achieved with a double-shaft granulating screw, called pugmill or blunger. The pugmill is followed by a drier, a cooler and screening.

The crushed oversize product is combined with the undersize product and the mixture is recirculated in a controlled ratio in the cold and dry state to the pugmill. The hot fresh slurry mixed with the recycle product gives soft balls with a moisture content of 3-6%, depending on the recycle ratio and the slurry.

Only partial crystallization and no moisture elimination takes place in the pugmill and therefore the soft balls must be dried in a rotary drum by hot air, cooled in a cooling drum and then screened.

Evaluation

For the mini-plant concept, due to the conditions prevailing in the regions favourable for the small size, the pan, respectively drum granulation seems more appropriate for nitrogenous fertilizers, due to the relatively high investment cost of the prilling power. For phosphate fertilizers it seems appropriate to start with non-granulated material (powder) and switch over only at a given degree of agricultural mechanisation to the granulated form.

Bulk Blending

Where granulated fertilizers should be blended, bulk blending is very advantageous. The process is usually of the batch type, with a minimal capacity of 1 to 2 tons per batch. The mixing time is 2 to 3 minutes, consequently as much as 10 to 20 tons per hour can be mixed easily. Depending on the working days of a year (approximately 100 days) the production will be about 10000 to 20000 tons per year.

The material commonly used in bulk blending are ammonium nitrate, ammonium sulphate, triple superphosphate, diammonium phosphate and potassium chloride. Other materials sometimes used are urea, ammonium phosphate nitrate (30-10-0), ammonium phosphate sulphate (16-20-0) and normal superphosphate.

The materials should be closely sized, dry enough to prevent caking in storage, and sufficiently strong to prevent fragmentation in handling.

Type of mixers and layout of storage, conveying and mixing facilities vary widely, so much that probably no two plants are alike. Since the plants are small and quite often built on a very limited budget, they tend to be homemade.

Mixers are mainly of rotating drum type, but various other types including ribbon mixers, mixing screws, gravity mixing towers and a volumetric metering device are used. The volumetric metering device is a continuous type in which materials are fed by gravity through adjustable gates onto a common belt. The materials mix as they flow into the receiving hopper and in the following screw conveyor.

Evaluation of Bulk Blending in the Distribution System

Until the emergence of bulk blending in the early 1960's, mainly in the USA, traditional distribution of fertilizer involved the movement of bagged fertilizers, from medium sized production plants producing between 25000 and 200000 tons per year of mixed fertilizers, to farm buyers, through general retail farm supply organizations.

However, in the early 1960's the advantage of bulk handling of fertilizers became apparent and the emergence of bulk blending developed quickly.

In bulk blending, a few basic high analysis materials containing single nutrients (or, in the case of ammonium phosphates, both nitrogen and phosphorus) are shipped in bulk form to retail bulk blending units. Here they are combinated physically in mixtures suited to the particular needs of individual farmers. Thus, at no point in the distribution chain are the materials handled in any but bulk form.

With the availability of a wide variety of blends of the three nutrients at the retailing station, the demand for basic products shifted from the chemically mixed fertilizer materials to major blending materials such as ammonium nitrate, triple superphosphate, diammonium phosphate, and muriate of potash.

As this mode of distribution developed, many of the major manufactorers developed their own organization of bulk blending stations, thus emerging as the direct seller to the farmer. A typical large fertilizer organization might develop a chain of 100 to 200 such bulk blending stations. These would typically handle from 1000 to 5000 tons of material per year and generally sell within a radius of 15 miles.

Advantages of bulk blending can be summarized in the following: (1) bulk blending shortens the marketing channel by combining the mixer and dealer functions; (2) handling and distribution costs are less for bulk material than for bagged product; (3) bulk blending reduces handling costs by eliminating the transfer from producer to dealer; (4) shipping distance of materials such as potash is shortened because the material goes directly from primarily producer to the mixer-dealer rather than detouring to a granulation plant; (5) a custom application service can be offered; and (6) the bulk blender, through his close contact with the farmer, can work with agricultural advisors in guiding the farmer's use of fertilizer. Assistance with soil testing and sampling is an important part of such a service.

Although this system is conceivable only as part of highly sophisticated national fertilizer supply and application chain and only countries with fairly developed agriculture can adopt it, it has its bearing on developing countries, too. The network initially developed in early stages of fertilizer production and distribution, composed of mini-plants and local dealers storage facilities can be easily transformed without any major change or investment.

Local stores can be provided with bulk blending facilities with very little cost and the new, or expanded phosphate fertilizer production units can produce granulated products, while the nitrogen is already manufactured in granulated form. The use of drum or pan granulation in the early mini-plants is advantageous, the particle size is particularly suitable for bulk blending.

Bagging

Although as much as possible of fertilizer output is dispatched as bulk, sometimes provision must be made for a significant amount of bagged process as well. Since bagging is a process with extremely high labour requirements, a fully automated line has been developed for this purpose.

Another question much debated is whether to use valved or cushion type bags. The latter are less expensive, prevent spillout, and protect the material much better: but weld sealing is difficult because of the fertilizer dust. The valve bags are more expensive and are not airtight, but they require no welding. Either type is available, according to the local conditions.

As compared to bulk goods, the handling of bagged fertilizers are much more labour-consuming operations.

Bagging in the mini-plant size range should be definitely discouraged for phosphates and limited even for nitrogen fertilizers. Anyhow, bulk storage and direct shipping to the plot is one of the most attractive features of mini-plants for agriculture.

3. ECONOMICS

The technical information summarized and analysed in the former part of the study served as basis for the economic evaluation of the miniplants. The investment figures and specific consumption data were taken from the latest published informations, from some bids and tenders made by leading contractors as well as from personel information sources. Table 3.1 and 3.2 summarize these data. The usual prices were introduced for the utilities. For the indirect costs, the usual factors were adopted.

The economic analysis and evaluation was prepared for three capacities: 150, 250 and 1000 ton/day ammonia production. The global investment cost for the whole complex, composed of the ammonia, urea and offsite units were calculated for three raw materials: natural gas, fuel oil and coal. For the 150 ton/day capacity two variants were taken into account: the first using the minimum amount of feedstock and importing the electric power for the drives; and the second, self contained, producing all the current needed at the expense of a higher feedstock consumption. This resulted in twelve cases, based on identical assumptions and calculation methods. The individual cost calculation sheets are presented in Tables 3.3 through 3.14.

Evaluation

First of all, it was obvious from the beginning, that with the prevailing very depressed world market prices no new project in this field can be profit-making, regardless of the capacity and feedstock. In 1986 all the leading fertilizer manufacturers realised heavy losses even with plants completely depreciated. Low feedstock cost helped neither: the plants based on associated gas worked also with losses. Faced with this situation, the ex-factory production costs were calculated for the mini and the big plants and compared.

Investment costs

In spite of the much simpler utility requirements and offsites, the miniplants need obviously higher specific investment costs as the big plants, calculated for the same conditions. So the 250 ton/day ammonia plant needs around 10% more investment per unit of production as the 1000 ton/day one. For the 150 ton/day plant 25% more specific investment is needed. This is much less, than that resulting from the usual relationship generally accepted between capacity and investment costs:

$$\frac{\text{investment cost A}}{\text{investment cost B}} = \left(\frac{\text{capacity A}}{\text{capacity B}}\right)^{n}$$

Accepting a relatively low value for n:

$$n = 0.6$$

the 250 ton plant would need 70% more specific investment. This comparison demonstrates clearly, that the small plants have no serious handicap in investment costs, especially when one takes into account the much lower absolute sums involved, facilitating greatly the credit procurement, accelerating the implementation and lowering the interest burden of the project.

Production costs

Natural gas based projects

The disadvantage in ex-factory costs for the 250 ton plant against the 1000 ton one is 25% or roughly 40$ in absolute figure. For the 150 ton case, it is 50%, respectively around 80$. Transport costs for ocean-going wessels with big tonnage amount easily to 40$/ton, while land transport costs in many developing countries exceed 80 $/tons for remote locations. On the other hand an option to produce in a big plant partly for export (to complete the smaller home market) would give certainly a certain advantage in the costs for the home market, but the losses on the exported quantity would largely offset these advantages. Supposing a 50 - 50% repartition between export and home market, a gain of nearly 7 million $ in the production for the home market would be accounted for against 7.5 million $ losses in the export at the prevailing prices.

So in all cases, when the transport costs exceed the relatively small difference in the costs at the disadvantage of the miniplants, which is the case in most developing countries for the remote areas, it is economically also justified to build miniplants, even when the other advantages exposed in other parts of this study are not considered.

The other feedstocks lead to substantially higher costs, calculated at world prices for the feedstock. In developing countries feedstocks can often be found with much lower cost level and thus economic projects can result from such conditions. The cost difference between miniplants and big plants is even smaller than for the natural gas based plants.

The general economic conclusion confirms that the minifertilizer plants can be competitive with the big plants in all the remote areas. The above rather general analysis justifies a detailed analysis based on the real actual conditions and prices in all cases, when the market is to small for a big plant and the transport costs are high.

Table 3.1.

Battery limit costs and consumption figures for ammonia production

(developed site)

Product: Ammonia

Nominal capacity		150 t/d			250 t/d			1000 t/d		
		nat. coal	fuel gas	oil	nat. coal	fuel gas	oil	nat. coal	fuel gas	oil
Battery limit cost	**M$**	28	32	45	38	43	60	140	160	230
Distribution of investment cost:										
- licence, know-how	%	2	2	2	2	2	2	2	2	2
- engineering	%	16	16	16	16	16	16	16	16	16
- equipment and machinery	%	53	53	53	53	53	53	53	53	53
- civil engineering and erection	%	29	29	29	29	29	29	29	29	29
Materials and Utilities Consumption/t	NH_3									
Feed and fuel	10^6 kcal	7.5	7.8	12.0	8.4	8.7	12.8	7.2	8.24	10.3
Power	kWh	885	900	1100	-	-	-	72	80	100
Cooling water	m3	515	550	800	300	320	345	120	160	171
BFW		1.8	2.0	3.5	1.2	2	3	1.2	2	3
Labour	manyear/year	40	50	75	40	50	75	40	50	100

Table 3.2.

Battery limit costs and consumption figures for urea production

(developed site)

Product: Urea

Nominal capacity		260 t/d	440 t/d	1700 t/d
Battery limit cost M$		15.0	23.0	68.0
Distribution of investment cost:				
- licence, know-how	%	23	23	23
- engineering	%	12	12	12
- equipment and machinery	%	52	52	52
- civil engineering and erection	%	33	33	33
Materials and Utilities Consumption/t urea:				
NH3	kg	578	578	578
CO2	kg	755	755	755
HP steam	kg	920	900	840
Power	kWh	150	140	130
Export steam LP	kg	-	-	500
Cooling water	m3	70	70	60
Labour	manyear/y (bagging + storage and loading included)	45	50	70

Table 3.3.

Estimated production cost for urea and ammonia

Feedstock: natural gas

Capacity:		
	150	ton/day ammonia
	49.5	Mton/year ammonia
	85.6	Mton/year urea
	41.3	Mton/year nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	28
Urea plant BL	15
Total BL	43
Offsites	11
Total fixed capital	54
Working capital	3.3

	Unit	Quant.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	7.5	3.5	1.29	26.25
El. power	kWh/ton	885.0	0.04	1.75	35.4
Cooling water	m3/ton	515.0	0.02	0.50	10.3
BFW	m3/ton	1.8	0.4	0.03	0.72
Total variable costs				3.59	72.67
Other direct costs					
Labour	manyear/y	40	20000	0.79	16.16
Maintenance(2% of fixed capital)				0.56	11.31
Total other direct costs				1.36	27.47
Total of all above costs for ammonia				4.95	100.14
Variable costs for urea production					
Ammonia	ton/ton	0.578	100.14	4.95	57.88
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.92	20	1.57	18.4
El. power	kWh/ton	150	0.04	0.51	6
Cooling water	m3/ton	70	0.02	0.11	1.4
LP steam credit	ton/ton	0	12	0	0
Total variable costs for urea				7.16	83.68
Other direct costs for urea					
Labour	manyear/y	45	20000	0.89	10.51
Maintenance (2% of fixed capital)				1.16	13.55
Total other direct costs for urea				2.06	24.06
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				0.67	7.94
General overhead (65% of labour)				1.10	12.90
Taxes, insurances (1.5% of fixed capital)				0.81	9.46
Interest (5% of working capital)				0.16	1.92
Depreciation (10% of fixed capital)				5.4	63.08
Total fixed costs for urea + ammonia				7.99	93.39
Total net costs for urea + ammonia				17.21	201.14
ROI (10%)				5.4	63.08
Ex factory costs				22.61	264.22

Table 3.4.

Estimated production cost for urea and ammonia

Feedstock: natural gas

Capacity:		
150	ton/day	ammonia
49.5	Mton/year	ammonia
85.6	Mton/year	urea
41.3	Mton/year	nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	28
Urea plant BL	15
Total BL	43
Offsites	11
Total fixed capital	54
Working capital	2.78

	Unit	Quant.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	8.6	3.5	1.48	30.1
El. power	kWh/ton	0	0.04	0.00	0.00
Cooling water	m3/ton	515.0	0.02	0.50	10.3
BFW	m3/ton	1.8	0.4	0.03	0.72
Total variable costs				2.03	41.12
Other direct costs					
Labour	manyear/y	40	20000	0.79	16.16
Maintenance(2% of fixed capital)				0.56	11.31
Total other direct costs				1.36	27.47
Total of all above costs for ammonia				3.39	68.59
Variable costs for urea production					
Ammonia	ton/ton	0.578	68.59	3.39	39.64
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.92	20	1.57	18.4
El. power	kWh/ton	150	0.04	0.51	6
Cooling water	m3/ton	70	0.02	0.11	1.4
LP steam credit	ton/ton	0	12	0	0
Total variable costs for urea				5.6	65.44
Other direct costs for urea					
Labour	manyear/y	45	20000	0.89	10.51
Maintenance (2% of fixed capital)				1.16	13.55
Total other direct costs for urea				2.06	24.06
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				0.67	7.94
General overhead (65% of labour)				1.10	12.90
Taxes, insurances (1.5% of fixed capital)				0.81	9.46
Interest (5% of working capital)				0.13	1.62
Depreciation (10% of fixed capital)				5.4	63.08
Total fixed costs for urea + ammonia				7.99	93.39
Total net costs for urea + ammonia				15.65	182.90
ROI (10%)				5.4	63.08
Ex factory costs				21.05	245.98

Table 3.5.

Estimated production cost for urea and ammonia

Feedstock: natural gas

Capacity:			
	250	ton/day	ammonia
	82.5	Mton/year	ammonia
	142.7	Mton/year	urea
	68.8	Mton/year	nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	38
Urea plant BL	23
Total BL	61
Offsites	18
Total fixed capital	79
Working capital	3.82

	Unit	Quant.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	8.4	3.5	2.42	29.4
El. power	kWh/ton	0	0.04	0	0
Cooling water	m3/ton	300	0.02	0.49	6
BFW	m3/ton	1.2	0.4	0.03	0.48
Total variable costs				2.96	35.88
Other direct costs					
Labour	manyear/y	40	20000	0.79	9.69
Maintenance(2% of fixed capital)				0.76	9.21
Total other direct costs				1.56	18.90
Total of all above costs for ammonia				4.52	54.78
Variable costs for urea production					
Ammonia	ton/ton	0.578	54.78	4.51	31.66
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.9	20	2.56	18
El. power	kWh/ton	140	0.04	0.79	5.6
Cooling water	m3/ton	70	0.02	0.19	1.4
LP steam credit	ton/ton	0	12	0	0
Total variable costs for urea				8.08	56.66
Other direct costs for urea					
Labour	manyear/y	50	20000	0.99	7.00
Maintenance (2% of fixed capital)				1.68	11.77
Total other direct costs for urea				2.68	18.78
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				0.72	5.04
General overhead (65% of labour)				1.17	8.19
Taxes, insurances (1.5% of fixed capital)				1.18	8.30
Interest (5% of working capital)				0.19	1.33
Depreciation (10% of fixed capital)				7.9	55.36
Total fixed costs for urea + ammonia				10.97	76.90
Total net costs for urea + ammonia				21.74	152.35
ROI (10%)				7.9	55.36
Ex factory costs				29.64	207.71

Table 3.6.

Estimated production cost for urea and ammonia

Feedstock: natural gas

Capacity: 1000 ton/day ammonia
330.5 Mton/year ammonia
570.9 Mton/year urea
275.6 Mton/year nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	140
Urea plant BL	68
Total BL	208
Offsites	83
Total fixed capital	291
Working capital	10.16

	Unit	Quant.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	7.2	3.5	8.31	25.2
El. power	kWh/ton	0	0.04	0	0
Cooling water	m3/ton	120	0.02	0.79	2.4
BFW	m3/ton	1.2	0.4	0.15	0.48
Total variable costs				9.26	28.08
Other direct costs					
Labour	manyear/y	40	20000	0.79	2.42
Maintenance(2% of fixed capital)				2.8	8.48
Total other direct costs				3.6	10.90
Total of all above costs for ammonia				12.86	38.98
Variable costs for urea production					
Ammonia	ton/ton	0.578	38.98	12.86	22.53
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.84	20	9.59	16.8
El. power	kWh/ton	130	0.04	2.96	5.2
Cooling water	m3/ton	60	0.02	0.68	1.2
LP steam credit	ton/ton	-0.5	12	-3.42	-6
Total variable costs for urea				22.68	39.73
Other direct costs for urea					
Labour	manyear/y	70	20000	1.4	2.45
Maintenance (2% of fixed capital)				5.52	9.66
Total other direct costs for urea				6.92	12.12
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				0.87	1.54
General overhead (65% of labour)				1.43	2.5
Taxes, insurances (1.5% of fixed capital)				4.36	7.64
Interest (5% of working capital)				0.50	0.88
Depreciation (10% of fixed capital)				29.1	50.97
Total fixed costs for urea + ammonia				35.77	62.66
Total net costs for urea + ammonia				65.37	114.51
ROI (10%)				29.1	50.97
Ex factory costs				94.47	165.48

Table 3.7.

Estimated production cost for urea and ammonia

Feedstock: fuel oil

Capacity:		
	150	ton/day ammonia
	49.5	Mton/year ammonia
	85.6	Mton/year urea
	41.3	Mton/year nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	32
Urea plant BL	15
Total BL	47
Offsites	17
Total fixed capital	64
Working capital	4.68

	Unit	Quant.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	7.8	12	4.63	93.6
El. power	kWh/ton	900	0.04	1.78	36
Cooling water	m3/ton	550	0.02	0.54	11
BFW	m3/ton	2	0.4	0.03	0.8
Total variable costs				6.99	141.4
Other direct costs					
Labour	manyear/y	50	20000	0.99	20.20
Maintenance(2% of fixed capital)				0.96	19.39
Total other direct costs				1.96	39.59
Total of all above costs for ammonia				8.95	180.99
Variable costs for urea production					
Ammonia	ton/ton	0.578	180.99	8.95	104.61
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.92	20	1.57	18.4
El. power	kWh/ton	150	0.04	0.51	6
Cooling water	m3/ton	70	0.02	0.11	1.4
LP steam credit	ton/ton	0	12	0	0
Total variable costs for urea				11.16	130.41
Other direct costs for urea					
Labour	manyear/y	45	20000	0.89	10.51
Maintenance (2% of fixed capital)				1.24	14.48
Total other direct costs for urea				2.14	25
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				0.76	8.87
General overhead (65% of labour)				1.23	14.42
Taxes, insurances (1.5% of fixed capital)				0.96	11.21
Interest (5% of working capital)				0.23	2.73
Depreciation (10% of fixed capital)				6.4	74.76
Total fixed costs for urea + ammonia				9.35	109.28
Total net costs for urea + ammonia				22.65	264.69
ROI (10%)				6.4	74.76
Ex factory costs				29.05	339.46

Table 3.8.

Estimated production cost for urea and ammonia

Feedstock: fuel oil

Capacity:			
	150	ton/day	ammonia
	49.5	Mton/year	ammonia
	85.6	Mton/year	urea
	41.3	Mton/year	nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	32
Urea plant BL	15
Total BL	47
Offsites	17
Total fixed capital	64
Working capital	4.48

	Unit	Quant.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	9.8	12	5.82	117.6
El. power	kWh/ton	0	0.04	0	0
Cooling water	m3/ton	550	0.02	0.54	11
BFW	m3/ton	2	0.4	0.03	0.8
Total variable costs				6.4	129.4
Other direct costs					
Labour	manyear/y	50	20000	0.99	20.20
Maintenance(2% of fixed capital)				0.96	19.39
Total other direct costs				1.96	39.59
Total of all above costs for ammonia				8.36	168.99
Variable costs for urea production					
Ammonia	ton/ton	0.578	168.99	8.36	97.67
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.92	20	1.57	18.4
El. power	kWh/ton	150	0.04	0.51	6
Cooling water	m3/ton	70	0.02	0.11	1.4
LP steam credit	ton/ton	0	12	0	0
Total variable costs for urea				10.56	123.47
Other direct costs for urea					
Labour	manyear/y	45	20000	0.89	10.51
Maintenance (2% of fixed capital)				1.24	14.48
Total other direct costs for urea				2.14	25
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				0.76	8.87
General overhead (65% of labour)				1.23	14.42
Taxes, insurances (1.5% of fixed capital)				0.96	11.21
Interest (5% of working capital)				0.22	2.61
Depreciation (10% of fixed capital)				6.4	74.76
Total fixed costs for urea + ammonia				9.35	109.28
Total net costs for urea + ammonia				22.06	57.75
ROI (10%)				6.4	74.76
Ex factory costs				28.46	332.52

Table 3.9.

Estimated production cost for urea and ammonia

Feedstock: fuel oil

Capacity:		
250	ton/day	ammonia
82.5	Mton/year	ammonia
142.7	Mton/year	urea
68.8	Mton/year	nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	43
Urea plant BL	23
Total BL	66
Offsites	26
Total fixed capital	92
Working capital	6.21

	Unit	Quant.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	8.7	12	8.61	104.4
El. power	kWh/ton	0	0.04	0	0
Cooling water	m3/ton	320	0.02	0.52	6.4
BFW	m3/ton	2	0.4	0.06	0.8
Total variable costs				9.2	111.6
Other direct costs					
Labour	manyear/y	50	20000	0.99	12.12
Maintenance(2% of fixed capital)				1.29	15.63
Total other direct costs				2.29	27.75
Total of all above costs for ammonia				11.49	139.35
Variable costs for urea production					
Ammonia	ton/ton	0.578	139.35	11.49	80.54
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.92	20	2.56	18
El. power	kWh/ton	140	0.04	0.79	5.6
Cooling water	m3/ton	70	0.02	0.19	1.4
LP steam credit	ton/ton	0	12	0	0
Total variable costs for urea				15.06	105.54
Other direct costs for urea					
Labour	manyear/y	50	20000	0.99	7
Maintenance (2% of fixed capital)				1.78	12.47
Total other direct costs for urea				2.78	19.48
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				0.8	5.6
General overhead (65% of labour)				1.3	9.11
Taxes, insurances (1.5% of fixed capital)				1.38	9.67
Interest (5% of working capital)				0.31	2.17
Depreciation (10% of fixed capital)				9.2	64.47
Total fixed costs for urea + ammonia				12.68	88.85
Total net costs for urea + ammonia				30.52	213.87
ROI (10%)				9.2	64.47
Ex factory costs				39.72	278.35

Table 3.10.

Estimated production cost for urea and ammonia

Feedstock: fuel oil

Capacity:			
	1000	ton/day	ammonia
	330	Mton/year	ammonia
	570.9	Mton/year	urea
	275.6	Mton/year	nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	160
Urea plant BL	68
Total BL	228
Offsites	135
Total fixed capital	363
Working capital	19.28

	Unit	Quant.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	8.24	12	32.63	98.88
El. power	kWh/ton	0	0.04	0	0
Cooling water	m3/ton	160	0.02	1.05	3.2
BFW	m3/ton	2	0.4	0.26	0.8
Total variable costs				33.95	102.88
Other direct costs					
Labour	manyear/y	50	20000	0.99	3.03
Maintenance(2% of fixed capital)				4.8	14.54
Total other direct costs				5.8	17.57
Total of all above costs for ammonia				39.75	120.45
Variable costs for urea production					
Ammonia	ton/ton	0.578	120.45	39.74	69.62
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.84	20	9.59	16.8
El. power	kWh/ton	130	0.04	2.96	5.2
Cooling water	m3/ton	60	0.02	0.68	1.2
LP steam credit	ton/ton	-0.5	12	-3.42	-6
Total variable costs for urea				49.56	86.82
Other direct costs for urea					
Labour	manyear/y	70	20000	1.4	2.45
Maintenance (2% of fixed capital)				5.92	10.36
Total other direct costs for urea				7.32	12.82
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				0.95	1.68
General overhead (65% of labour)				1.56	2.73
Taxes, insurances (1.5% of fixed capital)				5.44	9.53
Interest (5% of working capital)				0.96	1.68
Depreciation (10% of fixed capital)				36.3	63.58
Total fixed costs for urea + ammonia				44.26	77.53
Total net costs for urea + ammonia				101.15	177.17
ROI (10%)				36.3	63.58
Ex factory costs				137.45	240.76

Table 3.11.

Estimated production cost for urea and ammonia

Feedstock: coal

Capacity:			
	150	ton/day	ammonia
	49.5	Mton/year	ammonia
	85.6	Mton/year	urea
	41.3	Mton/year	nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	45
Urea plant BL	15
Total BL	60
Offsites	36
Total fixed capital	96
Working capital	5.55

	Unit	Quant.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	12	8	4.75	96
El. power	kWh/ton	1100	0.04	2.17	44
Cooling water	m3/ton	800	0.02	0.79	16
BFW	m3/ton	3.5	0.4	0.06	1.4
Total variable costs				7.79	157.4
Other direct costs					
Labour	manyear/y	75	20000	1.5	30.30
Maintenance(2% of fixed capital)				1.86	36.36
Total other direct costs				3.3	66.66
Total of all above costs for ammonia				11.09	224.06
Variable costs for urea production					
Ammonia	ton/ton	0.578	224.06	11.08	129.51
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.92	20	1.57	18.4
El. power	kWh/ton	150	0.04	0.51	6
Cooling water	m3/ton	70	0.02	0.11	1.4
LP steam credit	ton/ton	0	12	0	0
Total variable costs for urea				13.29	155.31
Other direct costs for urea					
Labour	manyear/y	45	20000	0.89	10.51
Maintenance (2% of fixed capital)				1.5	17.52
Total other direct costs for urea				2.4	28.03
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				0.95	11.21
General overhead (65% of labour)				1.56	18.22
Taxes, insurances (1.5% of fixed capital)				1.44	16.82
Interest (5% of working capital)				0.27	3.24
Depreciation (10% of fixed capital)				9.6	112.14
Total fixed costs for urea + ammonia				13.56	158.41
Total net costs for urea + ammonia				29.25	341.75
ROI (10%)				9.6	112.14
Ex factory costs				38.85	453.9

Table 3.12.

Estimated production cost for urea and ammonia

Feedstock: coal

Capacity:		
150	ton/day	ammonia
49.5	Mton/year	ammonia
85.6	Mton/year	urea
41.3	Mton/year	nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	45
Urea plant BL	15
Total BL	60
Offsites	36
Total fixed capital	96
Working capital	5.08

	Unit	Quant.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	14	8	5.54	112
El. power	kWh/ton	0	0.04	0	0
Cooling water	m3/ton	800	0.02	0.79	16
BFW	m3/ton	3.5	0.4	0.06	1.4
Total variable costs				6.4	129.4
Other direct costs					
Labour	manyear/y	75	20000	1.5	30.30
Maintenance(2% of fixed capital)				1.8	36.36
Total other direct costs				3.3	66.66
Total of all above costs for ammonia				9.7	196.06
Variable costs for urea production					
Ammonia	ton/ton	0.578	196.06	9.7	113.32
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.92	20	1.57	18.4
El. power	kWh/ton	150	0.04	0.51	6
Cooling water	m3/ton	70	0.02	0.11	1.4
LP steam credit	ton/ton	0	12	0	0
Total variable costs for urea				11.9	139.12
Other direct costs for urea					
Labour	manyear/y	45	20000	0.89	10.51
Maintenance (2% of fixed capital)				1.5	17.52
Total other direct costs for urea				2.4	28.03
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				0.95	11.21
General overhead (65% of labour)				1.56	18.22
Taxes, insurances (1.5% of fixed capital)				1.44	16.82
Interest (5% of working capital)				0.25	2.96
Depreciation (10% of fixed capital)				9.6	112.14
Total fixed costs for urea + ammonia				13.56	158.41
Total net costs for urea + ammonia				27.86	325.56
ROI (10%)				9.6	112.14
Ex factory costs				37.46	437.71

Table 3.13.

Estimated production cost for urea and ammonia

Feedstock: coal

Capacity:			
	250	ton/day	ammonia
	82.5	Mton/year	ammonia
	142.7	Mton/year	urea
	68.8	Mton/year	nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	60
Urea plant BL	23
Total BL	83
Offsites	58
Total fixed capital	141
Working capital	6.5

	Unit	Quant.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	10.8	8	7.12	86.4
El. power	kWh/ton	0	0.04	0	0
Cooling water	m3/ton	345	0.02	0.56	6.9
BFW	m3/ton	3	0.4	0.9	1.2
Total variable costs				7.79	94.5
Other direct costs					
Labour	manyear/y	75	20000	1.5	18.18
Maintenance(2% of fixed capital)				2.4	29.09
Total other direct costs				3.9	47.27
Total of all above costs for ammonia				11.69	141.77
Variable costs for urea production					
Ammonia	ton/ton	0.578	141.77	11.69	81.94
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.92	20	2.56	18
El. power	kWh/ton	140	0.04	0.79	5.6
Cooling water	m3/ton	70	0.02	0.19	1.4
LP steam credit	ton/ton	0	12	0	0
Total variable costs for urea				15.26	106.94
Other direct costs for urea					
Labour	manyear/y	55	20000	1.1	7.7
Maintenance (2% of fixed capital)				2.12	14.85
Total other direct costs for urea				3.22	22.56
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				1.04	7.28
General overhead (65% of labour)				1.69	11.84
Taxes, insurances (1.5% of fixed capital)				2.11	14.82
Interest (5% of working capital)				0.32	2.27
Depreciation (10% of fixed capital)				14.1	98.8
Total fixed costs for urea + ammonia				18.94	132.76
Total net costs for urea + ammonia				37.42	262.26
ROI (10%)				14.1	98.8
Ex factory costs				51.52	361.07

Table 3.14.

Estimated production cost for urea and ammonia

Feedstock: coal

Capacity:		
	1000 ton/day	ammonia
	330 Mton/year	ammonia
	570.9 Mton/year	urea
	275.6 Mton/year	nitrogen

Capital costs	MMUS$
Fixed capital	
Ammonia plant BL	230
Urea plant BL	68
Total BL	298
Offsites	200
Total fixed capital	498
Working capital	19.93

	Unit	Quan'.	Price US$/U	Ann.cost MM$	Unit cost US$/ton
Ammonia variable costs					
Feed and fuel	Gcal/ton	10.3	8	27.19	82.4
El. power	kWh/ton	0	0.04	0	0
Cooling water	m3/ton	170	0.02	1.12	3.4
BFW	m3/ton	3	0.4	0.39	1.2
Total variable costs				28.71	87
Other direct costs					
Labour	manyear/y	100	20000	2	6.06
Maintenance(2% of fixed capital)				9.2	27.87
Total other direct costs				11.2	33.93
Total of all above costs for ammonia				39.91	120.93
Variable costs for urea production					
Ammonia	ton/ton	0.578	120.93	39.9	69.9
CO2	ton/ton	0.755	0		
HP steam	ton/ton	0.84	20	9.59	16.8
El. power	kWh/ton	130	0.04	2.96	5.2
Cooling water	m3/ton	60	0.02	0.68	1.2
LP steam credit	ton/ton	-0.5	12	-3.42	-6
Total variable costs for urea				49.72	87.1
Other direct costs for urea					
Labour	manyear/y	70	20000	1.4	2.45
Maintenance (2% of fixed capital)				7.32	12.82
Total other direct costs for urea				8.72	15.27
Fixed costs for urea and ammonia					
Direct overhead (40% of labour)				1.36	2.38
General overhead (65% of labour)				2.21	3.87
Taxes, insurances (1.5% of fixed capital)				7.47	13.08
Interest (5% of working capital)				0.99	1.74
Depreciation (10% of fixed capital)				49.8	87.23
Total fixed costs for urea + ammonia				60.84	106.56
Total net costs for urea + ammonia				119.28	208.94
ROI (10%)				49.8	87.23
Ex factory costs				169.08	296.17

4. MINI FERTILIZER PLANT EXPERIENCE IN DEVELOPING COUNTRIES

There are few miniplants realised in developing countries. With the exception of two countries: China and Mexico, only single plants were built in the last fifteen years in some developing countries. These minifertilizer plants realized in developing world have a rather poor record. In most cases obsolete processes were used without modification and even the implementation was charged with severe burdens resulting from inadequate engineering and/or construction work. leading engineering and contracting companies did not invest in the big engineering work needed for a modern miniplant process (as explained before) so realisations which can serve as good examples are rather few. There are however two developing countries where substantial results were registered, China and Mexico.

4.1 China

Role and shape of miniplants

The ammonia industry is developing rapidly in China. In 1984, ammoria production reached 18,4 million tonnes. At present there are more than 1,000 small-scale and more than 30 medium-scale ammonia plants in China representing 57 % respectively 21 % of the total production. All small scale plants and most of the medium size plants are based on coal feedstock.

Table 4.1.1 shows how coal-based ammonia capacity in China has developed since 1970, while Table 4.1.2 shows the raw material base of the whole of China's ammonia capacity in 1984.

Table 4.1.1

Development of Ammonia Production in China

(thousand tonnes)

	Total	Large scale	Middle scale	Small scale
1970	2445	-	1445	1000
1975	6077	-	2533	3544
1977	8704	1245	2579	4880
1979	13481	2706	3518	7257
1980	14975	3127	3655	8194
1981	14883	3359	3667	7807
1982	15463	3448	3637	8379
1983	16771	3631	3683	9457
1984	18373	3928	3919	10526

Table 4.1.2

Ammonia Production in China from Various Raw Materials

(1984)

Raw material	Ammonia production (thousand tonnes)	Percentage of total production
Solid	12052	65,69
Anthracite	9821	53,45
Coke	1168	6,35
Coke *	979	5,33
Lignite	84	0,46
Liquid	2504	13,64
Fuel (crude) oil	1100	6,00
Naphtha	1404	7,64
Gaseous	3775	20,54
Natural gas	3475	18,91
Coke oven gas	192	1,04
Refinery gas	108	0,95
Other	42	0,23
Total	18373	100

* Coke made by local methods

Commercial coal gasification processes used in China for ammonia production

Fixed-bed gasification at atmospheric pressure

This is the oldest process for ammonia, but it is still popular in China because some of its characteristic features are especially suited to China. In particular:

It can operate on anthracite, which is abundant in China.

Semi-water gas (crude gas for ammoni with a calorific value of 2,000-2,100 kcal/m3) can be produced in this process using air; there is no need for an oxygen unit.

On account of the simplicity of the equipment and the consequent low investment cost, it is feasible to build up a large number of small- and middle-scale ammonia plant using the process.

At present the feedstock anthracite consumption of the most advanced medium-scale plants is about 1,150 kg/t NH3; the total energy consumption (fuel and feed) is around $14,5 \times 10^6$ kcal/t NH3.

Fluidized bed gasification

In the 1950s China imported two ammonia plants incorporating fluidized-bed gasifiers similar to the Winkler gasifier. The first plant was set up in Jilin and the second in Lanshou. The gasifier, with a diameter of 5,42 m, has a capacity of more than 20,000 m3/h.

The plants have their advantages and their disadvantages:

Advantages:

Cheap local lignite could be used as feedstock.

They operated smoothly.

Disadvantages:

The investment cost was much higher than that of the fixed-bed process for ammonia.

The ash has a high carbon content, and the carbon conversion rate is thus low (only 55-65 %).

With high feedstock and utility consumption, the overall energy consumption per tonne of ammonia was very high.

After operation for several years, the gasifiers were retrofitted to gasify fuel oil as feedstock in Jilin.

Fixed-bed pressure gasification

To make use of abundant and cheap local lignite available in Yunnan Province, four fixed-bed pressure gasifiers were installed in a medium-scale ammonia plant in Kaiyuan, Yunnan, in 1973. These gasifiers were similar to the original Lurgi gasifier and had been imported at the end of the 1950s. Each has a diameter of 2,6 m and a capacity of about 7,500 Nm3/h at a pressure of 22-25 kg/cm2.

Although the plant has operated for more than ten years there are still some problems. First, owing to the lower ash melting temperature of the lignite feedstock, the operating temperature in the gasifier has to be confined to a low level, so the gas effciency is not very high. Secondly, high levels of methane, tar and other impurities in the gas necessitate complicated gas purification procedures in the plant, and the resulting phenolic waste liquor and sulphur-containing waste gases have to be treated before discharge. Now a project to revamp the plant is under consideration. At the end of the 1970s an imported 1,000-t/d larga-scale ammonia plant based on coal was built in Shanxi. Four Lurgi gasifiers with a diameter of 3,8 m were installed in it. The downstream gas processing chain consists of sulphur-resistant shift, Rectisol washing, liquid nitrogen washing, methanation, ammonia synthesis and other sections. The total energy consumption is $12,6 \times 10^6$ kcal/t NH_3 (design value). It will be on stream in 1987.

Table 4.1.3
Commercial Coal Gasification Processes in Use in Ammonia Plants in China

	Fixed-bed		Fluidized	Fixed-bed
	Middle-scale	Small-scale	bed	at pressure
Feedstock type	Anthracite lump	Coal briquet	Lignite	Lignite
Size (mm)	25-100	644 x 38 x 26	1-10	10-40
Gasification agent	Air/ Steam	Air/ Steam	Oxigen/ Steam	Oxigen/ Steam
Gasifier:				
pressure (kg/cm2)	atm.	atm.	atm.	22-25
diameter (mm)	2740	2260	5420	2800
capacity (Nm3/h)	6000	3000	20000	7500
Gas composition (%)				
CO_2	8,5	11,5	26	26
CO	27,5	23	28	18
H_2	41	43	42	44
CH_4	1	1,5	2	10,5
N_2	21,5	20,5	2	1,5
O_2	0,5	0,5		
Consumption:				
Coal (kg/t NH_3)	1150 *	1200 *	3000 **	3500 **
Oxygen (m3/t NH_3)	-	-	620	580
Total Energy consumption (kcal/t NH_3)	$14,5 \times 10^6$	16×10^6		

* Standard anthracite ** Crude lignite

Coal gasification processes under demonstration or development

Entrained-bed gasification process

Research work on an entrained-bed gasification process started in the 1960s. At the time, a pilot gasifier (V=0.6 m3) was installed in Shanghai and the experiment to determine basic data on gasification of pulverized coal was carried out.

Subsequently, as the next phase in the development of the process a large pilot unit was installed in Lintong, Shanxi, in 1980. The gasifier (V=4 m3) has two burners and a capacity exceeding 1,800 m3/h.

The results of the experiments obtained from the pilot unit showed that two types of coal (long-flame coal and gas coal) could successfully be gasified in the pilot unit to produce high-quality crude gas suitable for producing ammonia and few by-products. The burner insulation materials and other equipment were shown to be acceptable.

The data and experience obtained in the pilot test were used as a guide in design, engineering and production work. To date, two entrained-bed gasification units have been built up as replancements for fixed-bed units in two separate small ammonia plants in Shangdong. Several feasibility studies on medium-scale ammonia plants are under way.

Vortex bed gasification process

In 1965, a vortex bed pilot unit for gasification of pulverized coal was built in Guangxi and experiments started. By the end of 1978 an industrial demonstration unit had been installed in a small ammonia plant in Guangxi too. The gasifier is basically a column with a diameter of 1,400 mm and a volume of 10,3 m3, fitted with two burners for pulverized coal/oxygen and two steam injection nozzles. The upper portion of the gasifier is the gasification reaction zone; the lower portion contains a slag bath. It has a production capacity of 3,300 m3/h (crude gas), which can meet the need of an 8,000/t/a NH3 (small) ammonia plant. The carbon conversion rate and slag removal are higher than in the entrained-bed process, but the time of continuous operation is shorter.

Pressure-gasification process for coal-water slurry

Since the 1960s a pressure-gasification process operating on a coal-water slurry has been under study. This process is know as second generation technolgy in the world. In addition to the wide range of suitable coal types, its outstanding advantages are its high carbon conversion, high production rate and low gas compression, as well as its lack of pollution.

To promote industrialization of this process, a bench-scale unit (20 kg/h coal at 20 kg/cm2) was installed in Lintong, Shanxi, in 1979. Up to the end of

1984, about a hundred runs had been made and various technical data were obtained in the unit, which provided the basis for further pilot unit design, engineering and operation.

Now a pilot coal-water slurry gasification unit (1,5 t/h coal at 35 kg/cm2) has been built in the same place. It consists of a wet grinder, slurry pump, gasifier, waste heat boiler and other sections. There are two grinding mills, one of which is a wet ball mill lined with rubber and the other is a horizontal colloid mill. The coal slurry concentration will be as high as 60-65 % and the carbon conversion rate will reach 99 % in future experiments.

It is expected that the pressure gasification of coal-water slurry will be widely applied in the ammonia industry in the near future.

Table IV
Coal Gasification Processes under Demonstration or Development in China

	Entrained bed	Vortex bed	Coal-water slurry Bench scale	 Pilot scale (design value)
Feedstock:				
type	long flame coal	lignite	long flame coal	long flame coal
size	80 % pass 200 mesh	60 % pass 130 mesh	coal-water slurry 60-68 %	 60 %
Gasifier:				
volume (m3)	4	10,3		
pressure (kg/cm2)	atm.	atm.	15-20	35
production	1800 m3/h	3300 m3/h	20 kg/h coal	1500 kg/h coal
Gas composition (%)				
CO2	15,8	16,7		
CO + H2	83	81,8	65	80
CH4	<0,1	0,1	<0,1	<0,1
O2	<0,1	0,1	<0,1	<0,1
N2	1	1,3		
Consumption (per 1000 m3 CO + H2)				
coal (kg)	730 (dry)	1880 *		
oxigen (m3)	400	400-440		450
Carbon conversion rate (%)	93	95	95-99	95-99
Cold gas efficiency (%)	70			

* Calorific value 3.542 kcal/kg

4.2 Mexico

The Mexican fertilizer industry has been growing by increasing the output of fertilizers until the end of 70's on the basis of erecting miniplants with the exception of the building of a triple super phosphate plant of capacity 270,000 tpa, one urea plant of 247,500 tpa in 1971 and the other of simple super phosphate of 300,000 tpa in 1978.

At present the total installed plant capacity is 4,800,000 tons of fertilizer of which 50.5 % is produced by miniplants and 49.5 % by maxiplants. On the other hand through the projects now under construction, the total installed plant capacity will increase to 6,960,000 tons which corresponds to 40 % installed plant capacity of miniplants and 60 % installed plant capacity by maxiplants. Thus the emphasis is now on the construction of maxiplants but it is important to point out that the strategy is not to displace all the miniplants. Both types of plants complement one another efficiently in meeting the increasing demand of fertilizer products and to optimise the use of raw materials.

The table gives an overview of the share of the miniplants in the fertilizer production capacity for both cases: actual operating plants (Table A) and taking into account the plants under construction too (Table B).

PARTICIPATION OF THE MINIPLANTS IN THE INSTALLED CAPACITY OF THE FERTILIZER INDUSTRY

(A) ACTUAL PLANTS

PRODUCT	Total (Thousand Mt)	Share of Total (%)	Miniplants (%)	Maxiplants (%)
Ammonium sulphate	1,673.7	34.79	69.31	30.69
Urea	1,753.0	36.44	10.58	89.42
Ammonium nitrate	168.0	3.49	100.00	-
Single superphosphate	482.5	10.03	37.82	62.18
Triple superphosphate	190.0	3.95	100.00	-
DAP-NPK	543.5	11.30	100.00	-
Total	4,810.7	100.00	50.50	49.50

(B) ACTUAL PLANTS AND UNDER CONSTRUCTION

PRODUCT	Total (Thousand Mt)	Share of Total (%)	Miniplants (%)	Maxiplants (%)
Ammonium sulphate	1,673.7	24.03	69.31	30.69
Urea	2,743.0	39.38	6.76	93.24
Ammonium nitrate	368.0	5.28	100.00	-
Single superphosphate	482.5	6.93	37.82	62.18
Triple superphosphate	340.0	4.88	100.00	-
DAP-NPK	1,357.5	19.5	40.04	59.96
Total	6,964.7	100.00	39.90	60.10

For the intermediate products, the installed plant capacity stands at 3,740.000 tons per year composed by the production of sulphuric, nitric and phosphoric acid, ammonia and ammonium nitrate solution. The details are given in table.

PARTICIPATION OF THE MINIPLANTS IN THE INSTALLED CAPACITY OF INTERMEDIATE PRODUCTS OF THE FERTILIZER INDUSTRY

(A) ACTUAL PLANTS

PRODUCT	TOTAL (Miles de t.)	TOTAL (%)
Sulphuric acid	2,949.6	78.84
Phosphoric acid	419.6	11.22
Nitric acid	155.0	4.14
Ammonium nitrate solution	195.0	5.21
Ammonia	22.0	0.59
TOTAL	3,741.2	100.00

(B) ACTUAL PLANTS AND UNDER CONSTRUCTION

PRODUCT	TOTAL (Miles de t.)	TOTAL (%)
Sulphuric acid	4,269.6	72.53
Phosphoric acid	815.6	13.85
Nitric acid	370.0	6.97
Ammonium nitrate solution	410.0	0.37
Ammonia	22.0	0.37
TOTAL	5,887.2	100.00

The plants utilize various technologies and different contractors involved in the projects. The ammonium sulphate plants have used Chemics knowhow with the exception of a crystalliser in Guadalajara where Struther Well process is used. The plants are mini in size and the contractor was Chemico.

For the last two plants of 200,000 tpa constructed in the complex of Queretaro, where FERTIMEX did the basic engineering for the project as the organization is interested in developing its own knowhow of the technologies needed and in this way to adapt processes for the conditions of the country.

For the production of urea, technologies developed by many organizations are is being used Lonza Lummus, Toyokoatsu, Stamicarbon and Snamprogetti and the contractors were Lummus, C&I Girdler, Foster Wheeler and Snamprogetti. The first two organizations built miniplants, whilst the Stamicarbon and Snamprogetti processes are maxiplants.

The technology and the contractor employed for the ammonium nitrate plants are as follows: Prilling Canada Development-Girdler Ltd.; Pechiney Saint Gobain-Saint Gobain and Stamicarbon-Krebs et Cie. The processes used for single superphosphate production were Sturtevant and Superflosoket and for triple superphosphate Dorr-Oliver, Saint Gobain and TVA.

The contractors involved in these projects were Chemico, Pedone, Girdler Ltd., Saint Gobain and Gulf Design. Lastly for DAP-NPK, the licensors and the contractors are as follows:

Dorr Oliver-Girdler Ltd., IMP, PEC-Girdler Ltd., TVA-Saint Gobain and TVA-Gulf Design.

5. CONCLUSION

Miniplants for most intermediates and end products of the fertilizer industry are available from reliable contractors with proven technologies. The processes used do not differ basically from those used for the big plants. Specific investment costs and operating costs are higher in the case of miniplants, but they can be nevertheless advantageous for the developing countries. Shorter implementation time, higher reliabilty, better utilisation, easier maintenance and less problems are the main advantages. The landed price at the farm gate in many cases will be not only competitive, but lower, than from the big plants, since in many developing countries transport costs and losses can add 30-100 % to the factory gate price, while the difference in the production cost is usually not more than 20-50 %.

Ammonia is a special case. The most recent developments introduced in the big plants need a scale down with corresponding simplifications and changes in the flowsheet and equipment design to render them applicable to the miniplant concept. No research or pilot plant experiments are necessary, but a considerable amount of engineering work must be spent on the design. No new flowsheet or process will result, but a simplified and more transparent plant using the same process steps and equipment. The engineering work will however represent a relatively high financial burden, which cannot be charged to the first implementation without affecting severily its viability, but should be distributed among several new plants to be realised.