Alternate Ammonia Feedstocks

Most of the ammonia in North America will come from natural gas, with $1 per million B.t.u. being the breakaway point where heavy oil and coal would be considered.

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The importance of fertilizer is well recognized, and today this importance is perhaps more sharply spotlighted than ever before. The current world fertilizer situation is characterized by constraints on the supply side, and in some instances the agricultural demand for nitrogen is not being met. There are many causes for this supply problem, among which is the constraint imposed by the worldwide energy imbalance.

The various industrial organizations, cooperatives, and government agencies involved in nitrogen production are, of course, taking steps to alleviate the supply shortage, but this task has been made all the more difficult by our energy problems.

This article summarizes some of the technical and economic aspects of ammonia production in light of the recent sharp changes in the availability and price of feedstocks. Naturally, the economic aspect is the most difficult one to come to grips with in a period in which the costs of everything are changing. However, we have developed some guidelines and principles we believe will be of general interest and application.

A recent paper (1) suggested the concept of giant plants located in areas at which natural gas is available in abundance, and subsequent transport of relatively low cost nitrogen product over long distances to the ultimate user. This approach has merit, but is complicated by the various factors involved in international commerce. Because of the risk associated with such international transactions, producers are showing considerable interest in alternate feedstocks, such as heavy fuel oil, or coal. The partial oxidation process can be applied to units with the flexibility to handle a wide range of liquid hydrocarbon feedstock, and there is merit in paying the premium for this flexibility in terms of the benefit of greater feedstock security. In some areas, notably in South Africa and India, new coal based ammonia plants have been announced.

A look at the various processes

Before discussing some of the economic and operating factors associated with the use of alternate feedstocks, it is advisable to review the various major processes available for ammonia manufacture. The bulk of worldwide ammonia production is obtained via steam hydrocarbon reforming. It is roughly estimated that ammonia from steam reforming operations totals 75%–80% of world production, and approximately 60–65% of this utilizes natural gas feed. It is expected that these percentages will more or less be maintained in the near future, but the prospects are that over the long range, utilization of coal and heavy oil feeds may increase significantly. In fact, many low capacity plants based on coal feeds (2) have been in operation for many years, and the background of experience with this type of plant is reasonably extensive. Also, announcements have recently been made that several large plants of the order of 1,000 short tons/day are in various stages of design and construction (3) based on coal feeds. Plants embodying use of the fuel oil partial oxidation process have been in operation for over 20 years, and the indications are that more may be built in the near future, particularly those having capacities of 1,000 short tons/day and greater.

Based on raw material availability, the choice of a process design is between catalytic steam reforming of light hydrocarbons and partial oxidation of heavy hydrocarbons. In certain exceptional cases where a source of hydrogen rich gas is available, the design involves cryogenic processing or some alternate process for purifying hydrogen. These designs, of course, must be provided with a source of nitrogen which is available as a waste stream from an air separation plant.

Figure 1. Typical ammonia process based on catalytic steam hydrocarbon reforming.
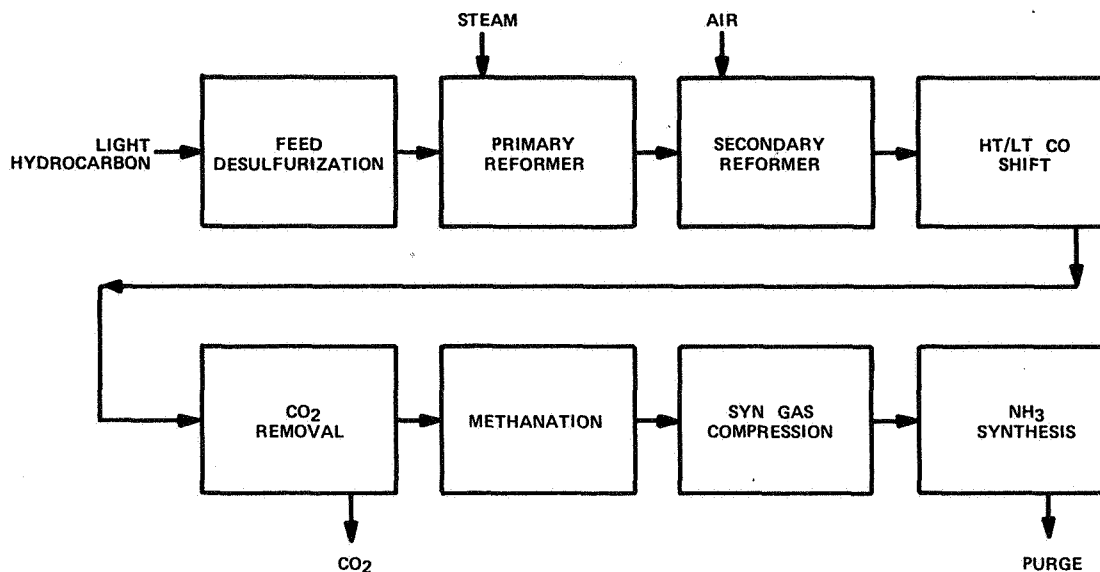
Catalytic steam reforming

Figure 1 shows a typical schematic flow sequence for a plant based on steam reforming of light hydrocarbon feeds. While the process sequence is essentially the same for raw gas generation, purification, and synthesis, there are variations with respect to feed pretreatment which largely depends on the type and quantity of contaminants in the feed. Briefly, the process consists of desulfurization, primary and secondary reforming for raw gas generation during which preheated air is introduced into the system, high and low temperature CO shift conversion, CO_2 removal, methanation for converting residual carbon oxides to methane which is purged in the synthesis loop, and finally conversion to ammonia. As a rule, synthesis is carried out at pressures ranging from 150 to 300 atm., depending on capacity and individual preferences. Whether low or high pressure is selected, all compression services for the large plant are based on use of centrifugal machinery using steam turbine drives in an efficient energy cycle embodying high pressure steam generation.

A large number of plants based on steam reforming have been constructed over the past two decades. Considering only the past nine year period, it is estimated that approximately 150 large plants (600 short tons/day and greater) are in operation or in various phases of design and construction. Most of these are based on high pressure reforming usually in the range of 450–500 lb./sq. in. gauge and synthesis pressures of 150 to 240 atm. Several large units having plant capacities of 1,500 short tons/day have been based on synthesis at 330 atm. It is estimated that by 1980 there will be about 170 large plants in operation on a worldwide basis.
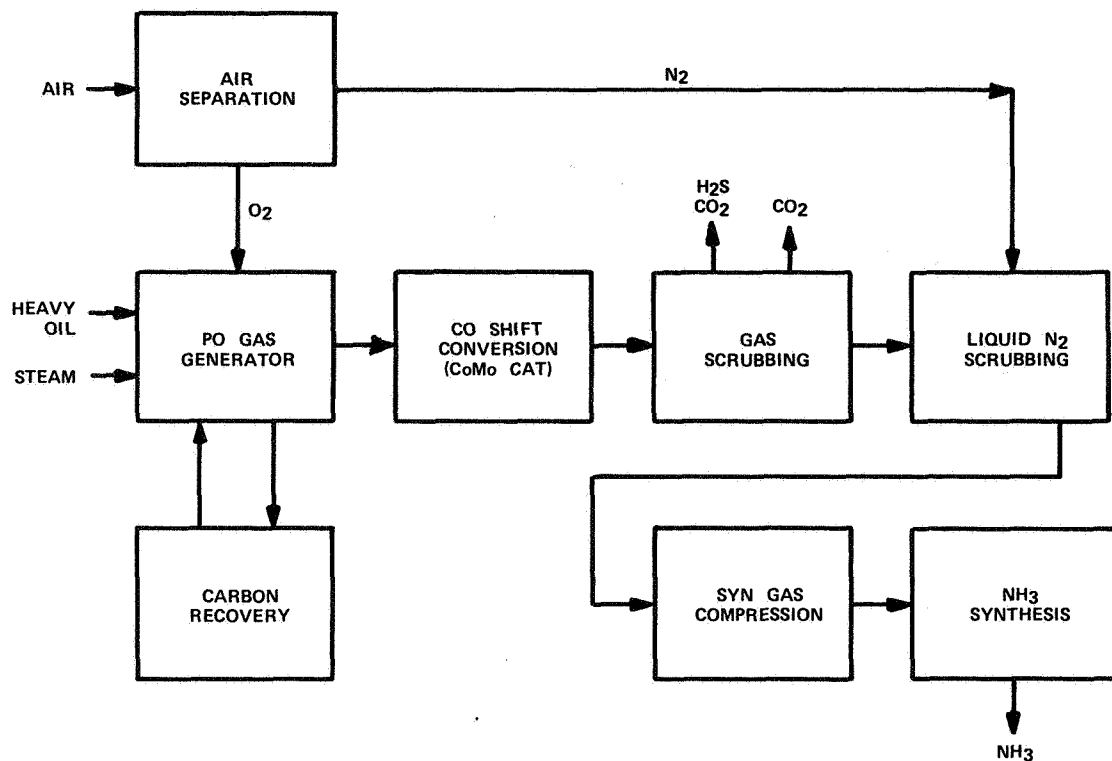
Partial oxidation of heavy hydrocarbons

Production of ammonia by the steam reforming route is straight-forward and requires a lesser number of equipment items. Reforming however, requires use of catalysts for the raw gas generation step. With the partial oxidation processes, generation of raw synthesis gas is carried out non-catalytically at relatively high temperature and pressure in conjunction with use of high purity oxygen in the combustion step. Moreover, additional equipment is required for treating raw gas, particularly for removing unreacted carbon which is ultimately recycled to the gas generation facility. Two major processes are available: one from Shell Development Co., and the other from Texaco Development Corp. Operating conditions for the gas generation step vary from 2,200–2,500°F. and 450–1,200 lb./sq. in. gauge. It should be mentioned that even higher pressures can be specified although uncertainty exists as to whether these can be economically justified. As already stated, the partial oxidation process can be used for both gaseous and liquid feedstocks. In fact, pilot study is under way on use of coal as a possible feedstock with these processes.

For locations where an abundant cheap source of natural gas or light liquid hydrocarbon exists, the economics show that the route of partial oxidation cannot be justified. For situations where light feeds are expensive or unavailable, the obvious process route is the partial oxidation one.

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Figure 2. Ammonia process based on partial oxidation of heavy hydrocarbons (Alternate 1).

AIR → AIR SEPARATION → N2

O2

HEAVY OIL →
STEAM →
PO GAS GENERATOR → CO SHIFT CONVERSION (CoMo CAT) → GAS SCRUBBING (H2S CO2, CO2) → LIQUID N2 SCRUBBING

CARBON RECOVERY

SYN GAS COMPRESSION → NH3 SYNTHESIS → NH3

Several raw gas treating techniques can be employed with the partial oxidation process and each depends on a number of factors including gas generation pressure, the energy balance, and whether cryogenic or chemical treatment is used for the gas purification step. All require an air separation unit, the design of which is related to the technique selected for raw gas treatment and the partial oxidation pressure. A brief description will be given for three alternate schemes which, for the most part, can be adapted to both the Shell and Texaco processes.

1. Maximum cryogenic treatment

Alternate 1, Figure 2, involves maximum cryogenic treatment of high pressure partial oxidation effluent gas which incidentally consists largely of H_2 and CO in almost equal parts, as well as small amounts of CO_2, CH_4, N_2, COS, and H_2S. Note that following gas generation, the gases are shifted for converting CO and small amounts of COS to CO_2 producing corresponding amounts of H_2 and H_2S. The shift catalyst specified for this service is cobalt-molybdenum, which must be maintained in the sulfided state for maximum activity. This is no problem since heavy oil feeds usually contain large amounts of sulfur compounds. Steam required for the shift conversion is obtained by quenching the hot partial oxidation gases using recycle and make-up water as the quench medium. Following the shift step and accompanying heat recovery, the gases are treated by absorption with a refrigerated physical solvent. Regeneration of solvent is carried out in two stages, the first of which enables recovery of H_2S as a concentrated stream suitable for sulfur recovery; and the other involves production of pure CO_2 which may be used as urea synthesis feed. This operation is followed by liquid nitrogen scrubbing which removes residual CO, CH_4 and argon. Nitrogen required for this operation as well as that for the ammonia synthesis step is provided by the air separation unit which also feeds oxygen directly to the partial oxidation generator without any compression. Synthesis gas leaving the nitrogen scrubbing facility is in an extremely high state of purity thus simplifying the design of the synthesis loop which incidentally does not require an inert gas purge. The minute amount of inerts in the synthesis feed gas are dissolved in the liquid ammonia product. The Alternate 1 scheme has been specified for partial oxidation systems operating at 80–85 atm. although it is applicable for operations involving lower pressures. The basic features of the Alternate 1 flowsheet therefore are: high pressure partial oxidation and associated water quench; elimination of oxygen compression; and a minimum number of catalytic services.

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Figure 3. Ammonia process based on partial oxidation of heavy hydrocarbons (Alternate 2).
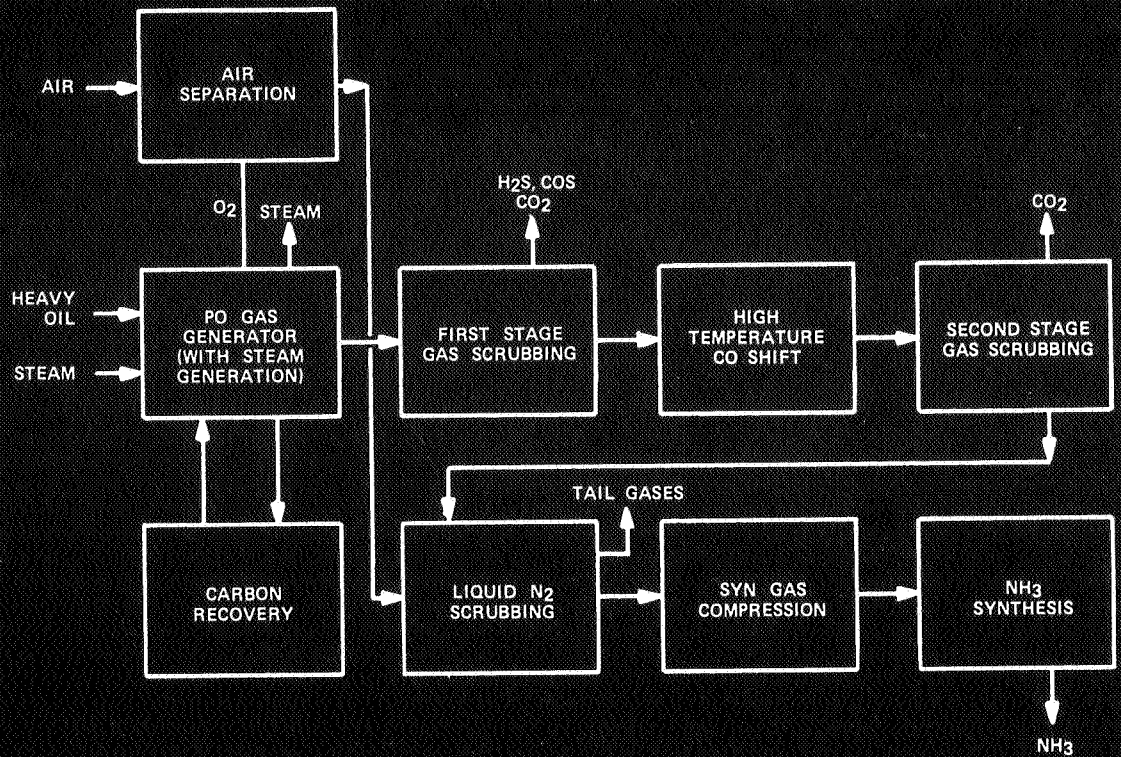
2. Gas generation at lower pressure

Alternate 2, Figure 3, embodies some of the features of Alternate 1 except that gas generation is carried out at lower pressure (usually 60 atm. and lower) and includes a waste heat boiler for recovering the heat of the effluent gases in the form of high pressure steam which is used in a high efficiency energy cycle. Following waste heat recovery, the gases are treated as follows: 1) gas scrubbing for removal of H_2S, COS and CO_2, 2) CO shift conversion using conventional catalyst, 3) gas scrubbing for removal of CO_2, 4) nitrogen scrubbing, and 5) compression and ammonia synthesis. The amount of steam used in the CO shift conversion step varies depending on the level of residual CO. As a rule, residual CO levels of 3–4% are selected which can easily be handled by nitrogen scrubbing with no penalty in investment. The basic differences between Alternates 1 and 2, therefore, are: variations in gas generation pressure, method of waste heat recovery, the system specified for CO shift conversion, and the acid gas scrubbing operations are carried out in two stages.

3. Using proven catalytic services

The Alternate 3 design, Figure 4, involves use of a large number of well proven catalytic services for treating partial oxidation gases. This scheme can be specified for a wide range of pressures up to 75–80 atm. Since the flow scheme involves use of low temperature shift catalyst, the partial oxidation pressure is limited because of water dew point considerations. This problem can be eliminated if the technique of double shift and methanation is selected. The double shift approach reduces the quantity of steam required for the LT shift step and hence enables an increase in the partial oxidation pressure. Briefly, Alternate 3 involves the following sequence: 1) recovery of waste heat via steam generation, 2) removal of H_2S and COS by solvent scrubbing, 3) high and low temperature shift conversion in conjunction with either the single or double shift technique, 4) CO_2 removal either as a single stage or double stage system depending on CO shift selection, 5) methanation, and 6) synthesis. Since this design does not include nitrogen scrubbing, inerts must be purged from the synthesis loop and hence hydrocarbon feed rate is somewhat higher. However, this is not a critical factor since purge gas serves as a source of fuel. Moreover, there is no significant horsepower penalty because this process can be used with low pressure synthesis which has been demonstrated in many large scale ammonia units.

There are many design approaches that can be

taken with respect to the partial oxidation processes and it is obvious that the sequence will be dictated by individual preferences and project situation. The question of which design offers the best overall economics would require a detailed engineering and estimating study which is not within the scope of this article.

Worldwide, there are approximately 100 Shell gasification units in use today in 19 countries for manufacturing raw hydrogen gas in conjunction with ammonia, methanol, hydrogen, and other products. An equal number of units are in operation embodying the Texaco process. It is estimated that the partial oxidation processes account for about 10% of total synthesis gas generation capacity.

Coal based processes

Several plants based on use of coal are in operation. One major Kellogg unit which was built in South Africa (SASOL) utilizes Lurgi gasification and Kellogg Fischer-Tropsch synthesis. This plant was designed for a large gas generation capacity and a relatively high gasification pressure. Currently several ammonia units embodying coal gasification are in various stages of design and construction.

Prior to World War II, practically all ammonia units were based on use of coal as a source of synthesis gas. Coal based plants slowly gave way to hydrocarbon based units, notably natural gas and liquid hydrocarbons, primarily because of the lower investment and the favorable price structure on these feeds. With the current energy supply problem and the prospects of rising costs for both natural gas and light hydrocarbons, the use of coal has re-entered the picture and is again being given serious consideration. It should be mentioned, however, that even with a coal based plant the economics will be highly dependent on the cost of coal which lately has also increased. Quality of coal also will be a significant factor in determining whether such a plant should be constructed. Nevertheless, in countries where there are abundant supplies of cheap coal, an absence of hydrocarbon feedstocks, and an assured domestic fertilizer market, coal gasification is a possible route to pursue.

Solid fuels ranging from coke, anthracite, bituminous, lignite, and peat can be gasified including caking grades of coal. Whether on-stream efficiency for this type of plant can approach that for the steam reformer remains to be seen. One critical economic factor, and probably the most important, is that to be competitive, a coal based design must be based on high ammonia capacity, using a high efficiency energy cycle in conjunction with centrifugal compression.

Lurgi gasification process

The scheme shown in Figure 5 represents the design described by Lurgi (4) for ammonia manufacture. Coal is gasified at about 20–30 atm. using oxygen and steam in a fixed bed system. Coal is charged through specially designed automatic lock hoppers with ash removal handled continuously through a rotating grate into a semi-automatic ash lock. Gasification temperature ranges from 1,050–1,150°F. depending on feed characteristics. Since the gasification temperature is in the intermediate range and operating pressure is high, the content of methane and CO_2 in the crude gas is considerably greater than that for conventional reforming and partial oxidation processes. A typical gas composition for a Lurgi gasifier would show levels of approximately 10–11% for methane and about 28% for CO_2. Methane, which is present at a relatively high concentration, is removed by liquid nitrogen scrubbing in a downstream processing operation and subsequently reprocessed for additional production of H_2 and CO. One of the advantages of the high pressure gasification is that it reduces the horsepower of the synthesis gas compressor. High pressure, low temperature, gasifier operation results in production of other components in the crude gas such as tars, naphtha, phenols, etc., components which must be removed from the gas stream and which can possibly be recovered as saleable product. One advantage of low gasification temperature is that it reduces the consumption of high cost oxygen and thus reduces compression horsepower for the air separation unit.

Crude gas from the Lurgi gasifiers undergoes several processing operations including recovery of waste heat; removal of tars, phenols and other byproduct components; Rectisol treatment for removal of sulfur compounds and CO_2; liquid nitrogen scrubbing to produce a highly purified synthesis gas; compression and, finally ammonia synthesis. The liquid nitrogen scrubbing step removes CO and methane which is recycled, after refrigeration recovery, to a sidestream conventional reforming and shift conversion facility for generation of additional H_2 and CO which are ultimately returned to the inlet of the Rectisol system.

There are several low capacity coal gasification units in operation based on the Lurgi process. Several high capacity plants are under design in the U.S. to produce substitute natural gas based on this process.

Koppers-Totzek process

The K-T coal dust gas process is illustrated schematically in Figure 6. Unlike the Lurgi process, gasification is based on low pressure and much

Figure 4. Ammonia process based on partial oxidation of heavy hydrocarbons (Alternate 3).
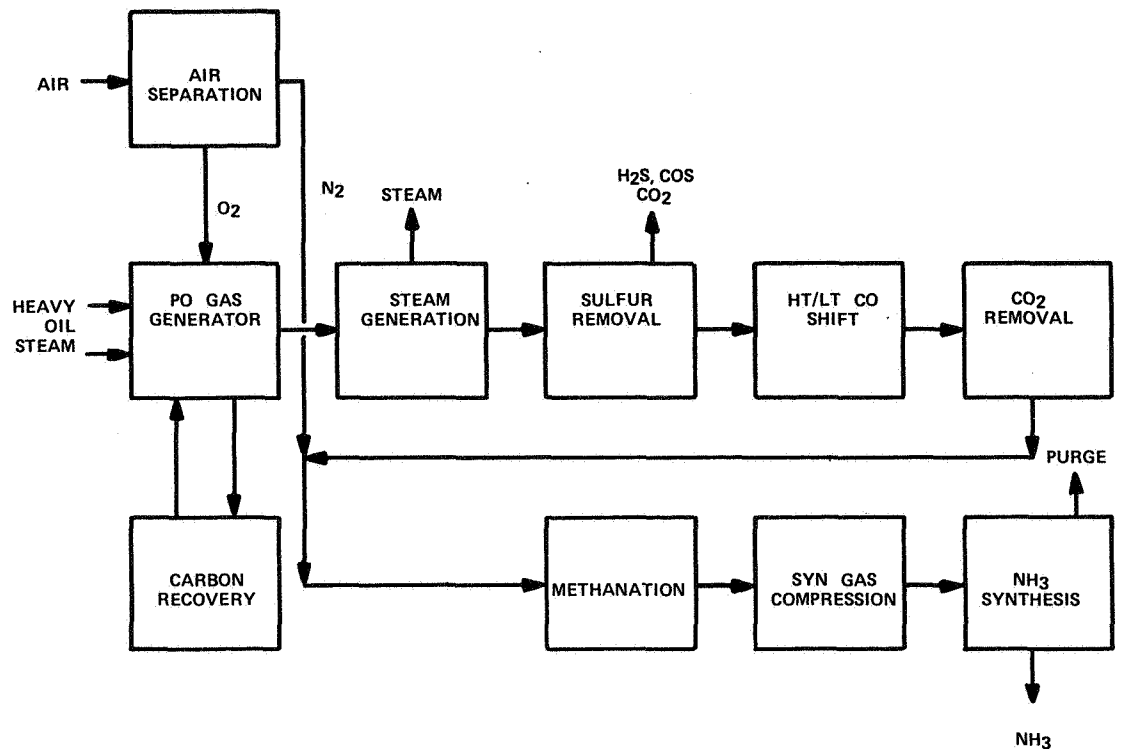
higher temperatures which insures complete conversion of higher hydrocarbons and thus eliminates additional crude gas conversion. Compression horsepower, however, is greater for the synthesis gas service. Depending on the reactivity of coal, up to 99% of the carbon is gasified using oxygen in the gasification step. The bulk of the gasified stream consists of H_2 and CO. Temperature of gasification is in the order of 2,750°F. and thus residual CH_4 levels of 0.1% and less can be expected. Koppers report that practically any type of coal can be gasified, notwithstanding the caking quality of the coal. Feed to the gasifier consists of a mixture of coal dust and oxygen and, if necessary, steam can be injected into the process through special burners. Following gasification and heat recovery, the crude gas undergoes a series of operations comprising steam generation, washing, compression, sulfur removal, additional compression, shift conversion, acid gas removal, nitrogen scrubbing, and synthesis.

Several Koppers ammonia plants in the range of 100 to 405 metric tons/day have been built in about eight countries. It is further reported that several large units are under construction in India and South Africa.

Other processes and feeds

Ammonia processes involving other schemes and feedstocks are also available. Most of these are variations of the processes described previously. As an example, light hydrocarbon feeds or refinery streams rich in hydrogen can be processed either catalytically or by cryogenic means to produce hydrogen, after which nitrogen is injected into the system in an amount required to satisfy the synthesis feed requirements. These plants are basically hydrogen units and are based on receiving low cost nitrogen from a neighboring air separation facility. Such a design is illustrated in Figure 7.

Coke oven gas was used as a feed for many ammonia units prior to World War II, and has been adopted more recently for several plants, including one large facility in the U.S. The principal drawbacks to coke oven gas are that it is available at low pressure, requires extensive pretreatment, and above all, the theoretical H_2 derived from each cubic foot of feed is relatively low. Compression horsepower, therefore, is high with this type of feed. A similar situation exists for any gas stream containing CO (or H_2) at atmospheric pressure. Use of coke oven gas for large, ammonia plant operation would obviously require close association with a large metallurgical facility, and would require an abundant source of nitrogen.

Economic considerations

With the present world situation on petroleum and the prospects for further increases in the cost

Figure 5. Ammonia process based on Lurgi coal gasification.
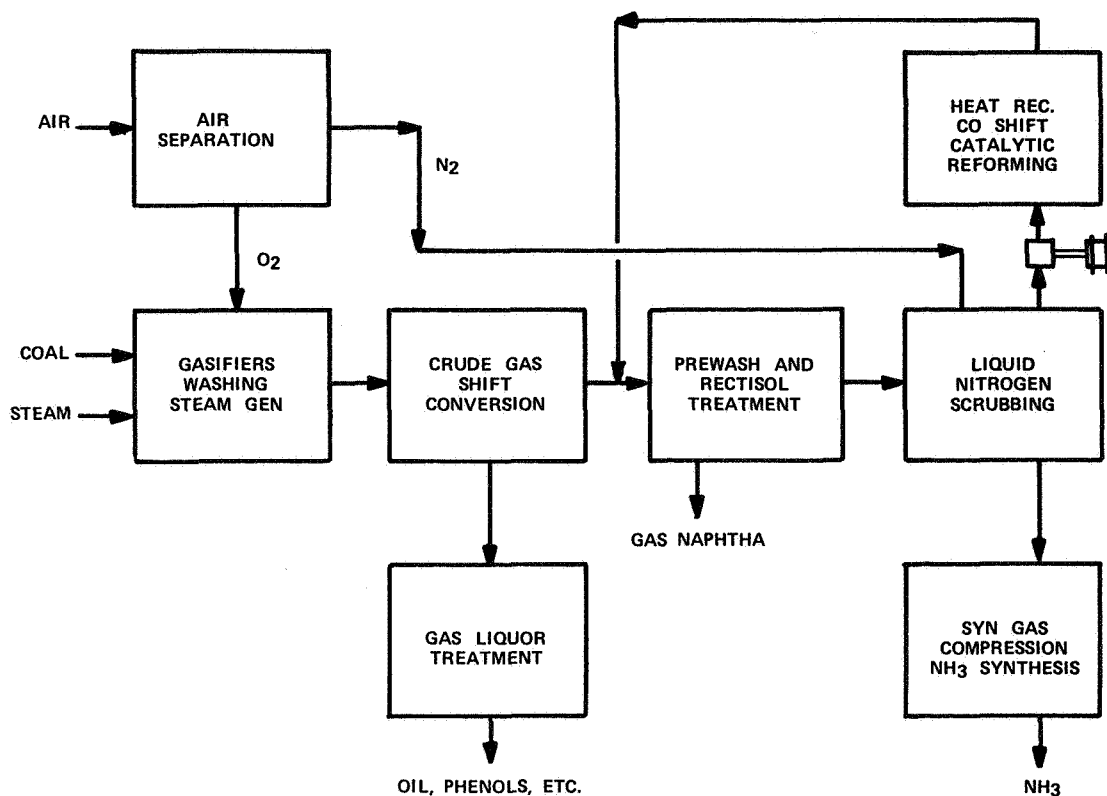
of feedstocks, it is difficult to make any precise predictions as to which feed (and fuel) to consider on a long range basis for ammonia production. There are a large number of factors playing major roles in determining which feedstock to use, among which are the feed price structure, plant size and location, transportation and shipping costs, contractual considerations, cost of capital, ultimate use of the product, that is, whether for export or captive use, and international situations. One therefore can only speculate as to how a few of these factors will affect the choice of feedstock which will briefly be discussed with emphasis on the unit price of feed.

Economic studies (1) reveal that the potential savings derived per ton of ammonia with increasing plant capacity begin to diminish as capacities exceed 2,000 short tons/day. In fact, it is uncertain whether there is economic incentive in designing giant plants. The optimum operating costs appear to fall in the range of 1,000–1,500 short tons/day, which represents the capacity selected for most large plants designed over the past nine years. For economic discussions, therefore, a capacity of 1,000 short tons/day will be selected which can be considered to be a representative size as far as the present and near future are concerned.

For a natural gas feed, manufacturing costs have been evaluated on a broad price range which will be divided into three categories: 1) $0.10/MM B.t.u., which can be considered for a location where there is an abundant supply of cheap gas such as the Middle East and North Africa; 2) $0.50/MM B.t.u., which is representative of many U.S. and European locations; and 3) a price of $1.00/MM B.t.u., which has been projected for many U.S. locations in the next few years. Table 1 was prepared to show the effect of feed price on ammonia manufacturing costs. Several assumptions indicated below were made in these calculations which incidentally reflect a battery limits plant including cooling tower and boiler feed water treatment. In view of the current situation with respect to equipment costs, available shop space and escalation, the investment figures must be considered very approximate and definitely not applicable to all locations.

1. A relatively low efficiency energy cycle was assumed for a plant design based on $0.10 gas, including a medium pressure steam system. For this type of plant, fuel economy is not considered a critical factor and thus thermal efficiency is low.

2. A high efficiency operation was assumed for the high feed cost area including use of a

1,500 lb./sq. in. gauge steam system. In addition, this design includes many process features for increasing thermal efficiency including combustion air preheat for the reformer, an "Amine Guard" system for CO_2 removal, a hydraulic turbine for power recovery in the CO_2 removal system, additional shift catalyst to reduce residual CO and synthesis loop purge, an additional stage of refrigeration in the loop to reduce compressor horsepower, and efficient steam turbines coupled with much higher steam superheat temperatures. While all these provisions permit significant reductions in utility consumption, the investment of course increases.
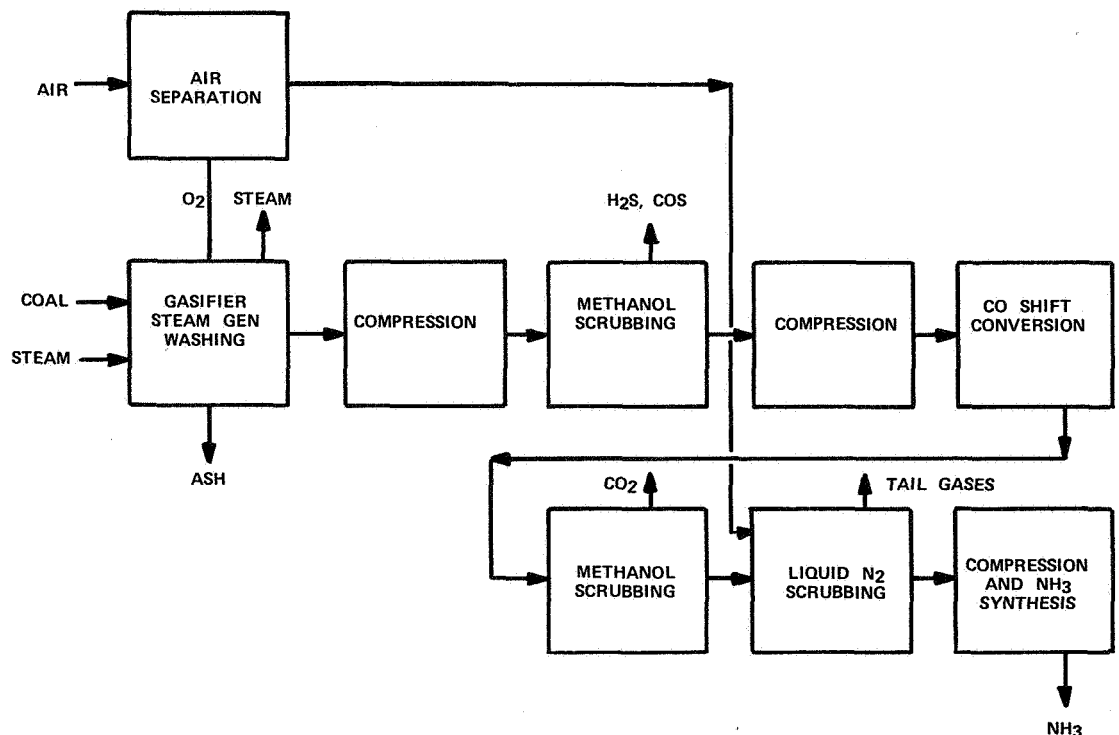
The economic data in Table 1 also includes an allowance of 17.5% for indirect costs and a 20% pre-tax return on investment. It will be noted that the cost of producing a ton of ammonia is roughly $62 for feed priced at $0.10/MM B.t.u.; $78 for $0.50 gas; and $93 for $1.00 natural gas. Note further that for the low feed cost location, feed and fuel represents about 6% of the total operating cost as compared to 33% for a design based on $1.00 gas. Apparent savings of approximately $31/ short ton exist for a unit feed price differential of $0.90/MM B.t.u. ($1.00-$0.10). If it is assumed that ammonia produced in a low feed cost location will serve solely for export purposes, the $31/short ton cost advantage permits shipment over long distances, the magnitude of which is dependent on the

shipping cost. If an additional allowance of approximately 30% of the investment is made for spare parts and offsite charges, the incremental operating cost for the low feed cost location would be about $16/short ton and thus would increase operating cost to about $78/short ton. This would then reduce the production cost differential from $31 to $15/short ton. Assuming that this price differential is genuine it would still be possible to ship low cost ammonia over reasonably long routes the distance of which is dependent on the cost per ton-mile and the quantity to be shipped on a regularly scheduled basis.

To ship or not to ship

If the price differential of natural gas feed is only $0.40/MM B.t.u. (say $0.50-$0.10/MM) between a high and low cost feed location, and again assuming the same 30% additional investment allowance for spare parts and offsite charges, the incentive to ship bulk ammonia obviously no longer exists. On the other hand, if natural gas were priced at much higher rates than indicated in Table 1, such as $1.50-$2.00/MM B.t.u., interest in production of ammonia in low feed cost locations will increase enormously in view of the extremely favorable production cost advantage. At present, it is difficult to pinpoint break-even rates on shipping distances in view of the lack of precise data on the cost of shipping bulk quantities of refrigerated ammonia over long routes. The gist of this eco-

Figure 6. Ammonia process based on use of Koppers-Totzek coal gasification.

Figure 7. Ammonia process based on use of outside source of nitrogen.

STEAM → CATALYTIC REFORMING

CO2 ↑ (from CO2 REMOVAL)

LIGHT HYDROCARBON → CATALYTIC REFORMING → HT/LT CO SHIFT → CO2 REMOVAL → H2 GAS TREATMENT

N_2 (FROM PIPELING FEEDER SYSTEM) → SYN GAS COMPRESSION → NH3 SYNTHESIS → NH3
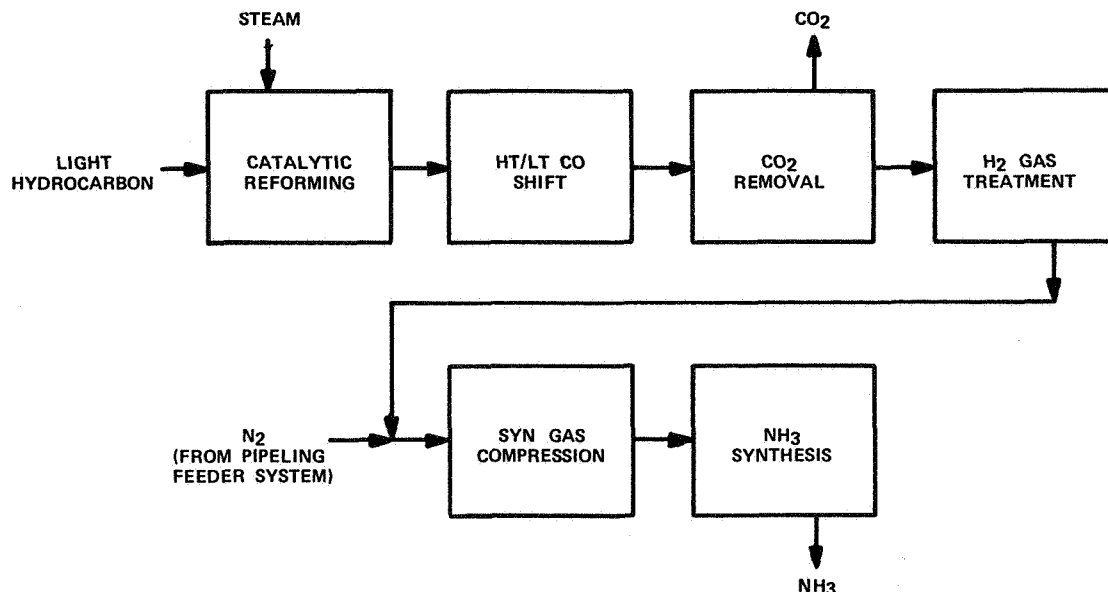
nomic analysis is that if an exceedingly high feed cost differential exists, the probability is that a sizable proportion of ammonia capacity in the future could shift to low cost natural gas locations. It should be mentioned, however, that while these paper analyses tend to favor ammonia manufacture in a low feed cost area, one cannot lose sight of the fact that a plant located in a 'low cost area may be faced with lack of maintenance facilities and other support facilities which may affect plant operations thus reducing stream efficiency with resultant increase in operating cost.

In recent years, considerable study has been devoted to use of cheap low grade fuels for firing a steam hydrocarbon reformer and its associated auxiliary boiler. Firing low grade fuel introduces many problems such as reformer tube attack due to presence of vanadium, sodium and ash; use of special burners which must be serviced frequently; and corrosion at the cold end of the furnace due to local condensation of acid (SO_2) on tube surfaces. If it is assumed that low grade fuels can be used, it is possible to effect slight reductions in the operating cost. Referring again to Table 1, it will be noted that if a lower cost fuel is used operating cost can be reduced by about $5/short ton for the fuel cost assumption indicated for the $1.00/MM B.t.u. natural gas feed case. While this approach effects some economy in operating costs, one would have to recognize that maintenance costs would be expected to increase. Moreover, more expensive materials of construction in the reformer add to investment, with corresponding increases in indi-
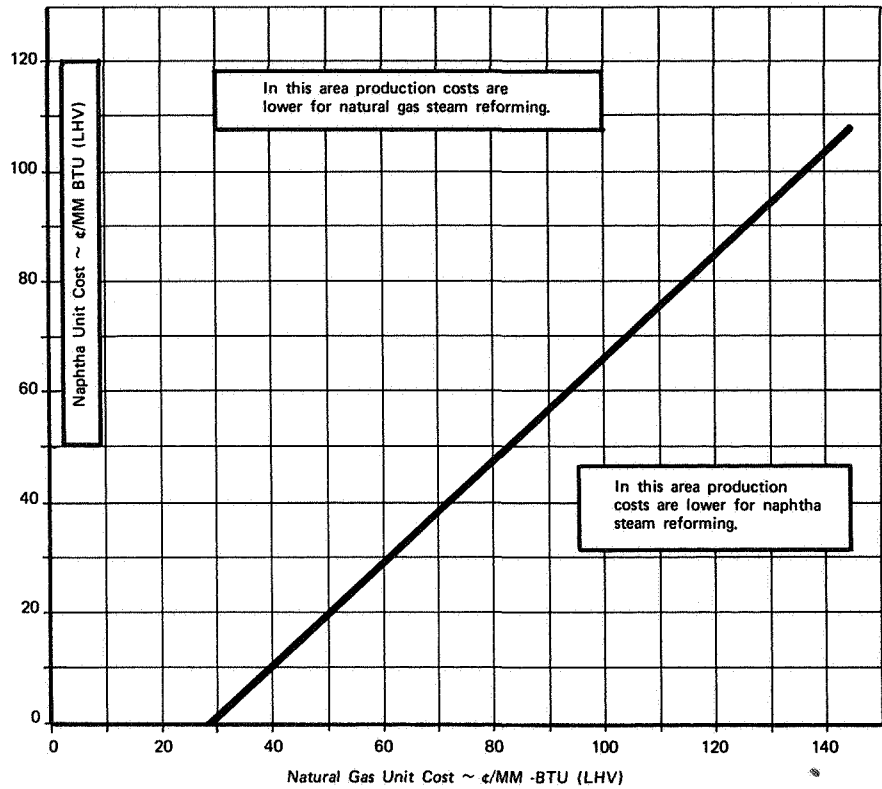
rect operating costs. The $5/short ton savings shown in the calculation, therefore, may not be fully realized.

Naphtha feeds not for the U.S.

A naphtha based plant presently costs about 15% more than a natural gas based ammonia plant. Assuming 1,000 short ton/day capacity, the incremental investment would run in the order of $7 million. Furthermore, naphtha feeds require more utility in terms of CO_2 removal, reforming steam, and feed pretreatment. Also, the unit cost of feed is usually much greater than natural gas on a calorific basis. Table 2 presents economic data on the same basis as that prepared for the natural gas case. Depending on price structure the operating cost can vary over a wide range. Note that for a feed price ranging from $0.50/MM B.t.u. to $2.00/ MM B.t.u. ($18.90–$75.60/short ton) the cost of producing a ton of ammonia ranges from $87 to $138. If lower grade fuels can be used for firing the reformer, substantial reductions in operating cost can be realized bearing in mind that the risk factor increases with use of lower grade fuels. The potential cost savings with use of lower grade fuels are also indicated in Table 2.

The comments made on natural gas with respect to operating cost and shipping considerations apply equally as well to naphtha feeds except that the operating cost differential between a plant based on naphtha and one based on low cost natural gas is even greater. Economic justification of a naptha based plant obviously is highly dependent on feed

Figure 8. Equivalent process of natural gas and naphtha feeds (steam reforming). Basis: see Table 1 and 2.

In this area production costs are lower for natural gas steam reforming.

In this area production costs are lower for naphtha steam reforming.

Naphtha Unit Cost ~ ¢/MM BTU (LHV)

Natural Gas Unit Cost ~ ¢/MM -BTU (LHV)

cost as well as several important factors such as whether the ammonia plant is part of a refinery-petrochemical complex and also on certain import/export restrictions on naphtha within a given country. It is highly unlikely that naphtha would be given serious consideration as a feedstock in the U.S. Use of naphtha for ammonia production would appear to be feasible only under special project situations existing at a given location.

Figure 8 shows relative differences between natural gas and naphtha processing from the standpoint of unit feed costs using the data developed in Tables 1 and 2 as a frame of reference.

Heavy oil feeds

Because of design and process simplicity, a steam-hydrocarbon reformer based plant costs less than a partial oxidation unit primarily because it requires less equipment. A partial oxidation unit requires an air separation plant, additional desulfurization, carbon removal and recycle, and other facilities. Nevertheless, if the cost of fuel oil is sufficiently low one could justify a design selection based on partial oxidation. In view of the uncertainty of equipment cost information and the continual need to escalate investment with current marketing conditions, it is extremely difficult to

pinpoint investment differences between a steam reformer plant and a partial oxidation unit. Estimates are that the investment difference varies over a wide range. Banquy (5) reports this to be as low as approximately 20%, but Mondo (6) states the difference to be closer to 40%. Some recent studies reveal that the difference may be 50% and possibly higher. Obviously, this is an area of great uncertainty. We believe it to be fruitless to present specific economic differences between the processes in this article largely because of constantly changing cost structures on various feeds and the difficulty of obtaining a sound economic relationship between both processes. In view of these considerations, we will evaluate the economics on the basis of the unit cost of feed for each case and indicate the degree of potential direct operating savings that can be expected for various price differentials on feed. This can be carried out on the basis of use of the economic information given in Tables 1 and 2 and assuming that the comparison is based on hydrocarbon (natural gas or naphtha) priced at $1.00/MM B.t.u. which will serve as a projected future cost. Direct operating data for the partial oxidation process were taken from Banquy's (5) article for comparable items. These are shown in Table 3 for four price structures for heavy fuel oil which will be the feed for the partial

oxidation process. The Banquy data have been converted to a basis comparable with that given in Tables 1 and 2 with respect to cooling water make-up and price structure.

Referring to Table 3, note that direct operating costs with the partial oxidation process range from $12.54 to $36.19 for fuel oil costs ranging from $10 to $34.50/short ton ($0.29 to 1.00/MM B.t.u.). Using these data along with those shown in Tables 1 and 2 enables development of the economic comparison given in Table 4, which refers only to direct operating cost items. Table 4 is divided into two categories: 1) direct operating cost difference between natural gas and fuel oil, and 2) the same cost difference between naphtha and fuel oil. If it is assumed that fuel oil is priced at $10/short ton ($0.29/MM B.t.u.), the data reveal that a savings of approximately $21 million over a three year period can be realized with the partial oxidation process if the natural gas cost is $1.00/MM B.t.u. Additional figures on potential direct cost savings are shown for other fuel oil cost structures. It will be noted that the economic incentive decreases sizably as the cost of fuel oil increases.

Similar data are given for naphtha in the Table 4 comparison. Again assuming that fuel oil is priced at $10/short ton, the data reveal a potential direct cost savings of about $24 million over a three year operating period with the partial oxidation process. Obviously, because of the unusual worldwide price/demand patterns of various feeds, it is difficult to explore the economics of producing ammonia on the basis of light hydrocarbon reforming and heavy oil partial oxidation. Sufficient data are given in the various tables, however, to enable the reader to insert other price structures on each feed to obtain potential direct cost savings with the partial oxidation process and thus determine whether the partial oxidation process is an economically viable proposition for a given project situation. It is worth mentioning that in years past the price differential between natural gas and heavy oil (or naphtha and heavy oil) was relatively low, and thus there did not appear to be sufficient justification to design on the basis of the partial oxidation technique. This is borne out by the fact that total ammonia produced from heavy oil is estimated to be about 10% of world capacity. How-

Table 1. Economics of ammonia manufacture, 1,000 short ton/day capacity.

Process Sequence: Natural Gas desulfurization, reforming, CO shift conversion, CO_2 removal, methanation, compression, synthesis

Investment	$47,000,000		$50,000,000		$50,000,000
Design Features					
Steam Pressure, lb./sq. in. gauge	600	....	1,500	....	1,500
CO_2 Removal System	MEA	....	Amine Guard	....	Amine Guard
Combustion Air Preheat	No	....	Yes	....	Yes
Natural Gas Price	Low—$0.10/MM B.t.u.		High—$0.50/MM B.t.u.		High—$1.00/MM B.t.u.
Power Cost	$0.006/kwh		$0.01/Kwh		$0.01/Kwh

Raw Material	Unit Rate	$/day	Unit Rate	$/day	Unit Rate	$/day
Natural Gas Feed	850 MM B.t.u./hr	2,040	850 MM B.t.u./hr	$10,200	850 MM B.t.u./hr	20,400
Utilities						
Fuel	670 MM B.t.u./hr	1,608	440 MM B.t.u./hr	5,280	440 MM B.t.u./hr	10,560...
Power	640 KW	92	640 KW	155	640 KW	155
CW (Make-up)	1290 gal./min.	465	1290 gal./min.	465	1290 gal./min.	465
BFW (Make-up)	390 gal./min.	560	390 gal./min.	560	390 gal./min.	560
Steam	In Balance	—	In Balance	—	In Balance	—
Other						
Catalysts & Chemicals		700		700		700
Labor	5 Men/Shift	480		700		480
Labor/Plant Overhead	100% Labor	480		480		480
Indirect Charge & Return on Investment		55,728		59,285		59,285
Total Cost—Per day		$62,153		$77,605		$93,085
per short ton		$ 62.15		$ 77.60		$ 93.09 ($87.80)(7)

Notes:
1. Investment covers battery limits unit including cooling tower and boiler feedwater treating. No inclusion is made of spare parts.
2. Utilities for high natural gas location ($0.50 and $1.00/MM B.t.u.) relates to high efficiency operation including combustion air preheat, high efficiency pumps and turbines, high steam superheat temperature. Cost of CW circulation unit not included in costs.
3. Utility Cost: Cooling water makeup @ $0.25/1000 gal., BFW make-up @$1.00/1,000 gal. Make-up water relates to net requirements for BFW and cooling water.
4. Catalyst and chemicals—$0.70/short ton.
5. Labor refers to battery limits unit only. Labor—$4/MH. Costs include allowance for supervision and fringe benefits at 100% of labor.
6. Indirect charges include return on investment.
7. Figure in bracket is estimated cost per ton if a relatively lower cost fuel ($0.50/MM B.t.u.) is used for firing reformer and auxiliary boiler.

Table 2. Economics of ammonia manufacture, 1,000 short ton/day capacity.

Process Sequence: naphtha desulfurization, reforming, CO shift conversion, CO_2 removal, methanation, compression, synthesis

Investment.. $57,000,000

Design Features
    Steam System Pressure, lb./sq. in. gauge 1,500
    Desulfurization .... Hydrodesulfurization—H_2S Stripping—Catalytic

Naphtha Cost Structure

$/MM B.t.u. (LHV) ....$	0.50 .... $	1.00 .... $	2.00
($/short ton)	(18.90)	(37.80)	(75.60)

Raw Material	Unit Rate	Cost/day, U.S. Dollars		
Naphtha Feed .... 873 MM/B.t.u./hr.		10,476	20,952	41,904
Utilities				
Net Fuel.................... 533 MM/B.t.u./hr.		6,396	12,792	25,584
Power .... 900 KW		216	216	216
CW (Make-up)..............1,130 gal./min.		407	407	407
BFW (Make-up)............. 465 gal./min.		670	670	670
Other				
Catalysts and Chemicals $0.80/ton......................		800	800	800
Labor, 5 men/shift...		480	480	480
Labor/Plant Overhead, 100% Labor......................		480	480	480
Indirect Charges &				
Return on Investment....................................		67,585	67,585	67,585
Total Cost—per day		$87,510	$104,382	$138,126
per short ton....................................		$ 87.51	$ 104.38	$ 138.13
			($ 97.99)(4)	($ 118.93)(4)

_____

Notes:
1. Utility cost structure: Power @ $0.01/KwH, cooling make-up water @ $0.25/1,000 gal., BFW makeup @ $1.00/1,000 gal.
2. Indirects same basis as Table 1 conditions.
3. Naphtha LHV 18900 B.t.u./lb. (10,500 K-Cal./KG.).
4. Figure in bracket is estimated cost per ton if a relatively lower cost fuel ($0.50/MM B.t.u.) is used for firing reformer and auxiliary boiler.

ever, recent figures show that naphtha cost is increasing at a much greater rate than heavy fuel oil thus tending to eliminate naphtha as a feedstock contender, and leaving the competition to natural gas and heavy oil.

Coal feeds cost 1.75 times gas

As in the case of steam reforming and partial oxidation, a plant based on use of coal as feedstock is expected to cost more than one based on partial oxidation because of additional equipment requirements, particularly with regard to coal handling operations, gasification, and raw gas treatment. Despite these cost differences coal based plants can be justified if the cost of coal is sufficiently low, that is, relative to light or heavy hydrocarbons. The approximate investment (7) for a coal based plant appears to be at least 1.75 times that for a natural gas plant, and possibly even more. The investment is at least 1.3 times that for a fuel oil plant.

Many factors enter the economic picture as far as selection of a coal based plant is concerned such as feedstock availability, and whether the price differential between the two feeds under consideration exists over a long operating period. If the economics are considered strictly on feed and fuel consumption, which is a reasonable criterion for a preliminary economic investigation, the following are typical requirements for each of the major processes considered in this article.

Approx. Feed and Fuel Requirement

Process	Ton/Ton	MM B.t.u./ ST(LHV)
Natural Gas Reforming .... —		31
Naphtha Reforming.............0.9		34
Fuel Oil Partial Oxidation 1.0		35
Coal Gasification* 2.0		45

*After credit for value of by-products.

As discussed, in energy deficient locations the price of natural gas can range from $0.50 to $1.00/MM B.t.u. Naphtha has recently been priced from $1.00 to $2.00/MM B.t.u. ($37.80–$75.60/short ton). Fuel oil cost varies considerably depending on location, but has been reported to range from $11 to as high as $30/ton, although in petroleum deficient locations prices approaching $40 to $60/short ton have been quoted. In view of this wide range of prices, it is extremely difficult to derive any generalized conclusion on the relative merits of one feedstock over the other particularly when the price of coal also shows equal variations, but at somewhat lower cost levels. Some estimates show the cost of coal being as low as $5/short ton in South Africa compared with figures of $15–$20/short ton in other locations. If the economics are deliberately considered on the basis of a high price projection for natural gas (say $1.00/MM B.t.u.) and the lowest price for coal (say $5/short ton), the potential savings per ton is of the order $21 for the coal based plant. Assuming a simple payout period of three years and 330 operating days per year, the overall savings would be about $21 million. Based on the projected investment for a coal based plant of 1.75 times for the natural gas plant, the economics indicate that the utilization of coal for ammonia production cannot be justified. If, however, no other feedstocks are available or a longer payout period is considered, then a coal based plant has to be given serious attention. This analysis is shown in Table 5 along with data covering other unit price structures.

Similar analyses can be carried out with other feeds using the same approach employed for the natural gas versus coal comparison. Comparing coal gasification with naphtha (refer to Table 5) would follow the same general line of reasoning. But if the investment for the naphtha based plant is about 15% higher than the natural gas plant, then a payout period of slightly over three years would tend to favor the coal based facility assuming that naphtha is priced at $1.00/MM B.t.u. and coal at $5/short ton.

If it is assumed that natural gas and/or naphtha is available at $2.00/MM B.t.u., a coal based operation would definitely be economically attractive even for payout periods of less than three years.

Comparing partial oxidation of fuel oil with coal gasification differs somewhat from steam reforming. Both fuel oil and coal processing involve the partial oxidation technique and similar downstream gas treating schemes. Furthermore, both involve use of an air separation plant and sulfur recovery facilities. Differential investment for the

Table 3. Partial oxidation of fuel oil, feed, utility, catalysts, and chemicals summary.

Capacity: 1,000 short ton/day

Design Features

Feed ....Heavy Fuel Oil
Process ....Partial Oxidation
Gas Generation Pressure ....80 Kg./Cm.^{2}
Process Sequence ....See Figure 2

Fuel Oil Cost Structure

	10	20	30	34.50
$/short ton	10	20	30	34.50
($/MM B.t.u)	(0.29)	(0.58)	(0.87)	(1.00)

Fuel Oil and Utilities

Item	Rate/short ton	Cost/short ton, Dollars			
Fuel Oil	0.965 ST	9.65	19.30	28.95	33.30
Power	30 KWH	0.30	0.30	0.30	0.30
CW (Make-up)	3070 gal.	0.75	0.75	0.75	0.75
BFW (Make-up)	320 gal.	0.32	0.32	0.32	0.32

Other

Catalysts & Chemicals	$0.30	0.30	0.30	0.30	0.30
Labor @ $4/MH,	7 Men/Shift	0.61	0.61	0.61	0.61
Labor & Plant Overhead,	100% Labor	0.61	0.61	0.61	0.61
Total Cost—per short ton		12.54	22.19	31.84	36.19

Notes:
1. Utility cost structure as follows:
   Power @ $0.01/KWH
   Water make-up @ $0.25/1,000 gal. for cooling water
   Water make-up @ $1.00/1,000 gal. for BFW
2. LHV of fuel oil 17,300 B.T.U./lb. (9610 K-Cal./Kg.).

two designs is not as great as that reported for the reforming vs. coal gasification comparison, and thus a more careful economic appraisal would be required which would have to include such factors as fuel oil availability over a lengthy period, labor costs covering coal mining operations, price stability, transportation costs for both product and feed, and plant location.

How the conclusions shape up

Production capacity for synthetic ammonia will continue to increase throughout the world as a result of accelerating demand for nitrogen fertilizers and nitrogen based chemicals. One would expect that as population growth continues, world needs for food, and hence for fertilizer, will follow this trend. While there are many areas where ammonia shortages exist, on the whole one would have to state that a good balance has been maintained thanks to the enormous increase in ammonia production over the past 10 years which witnessed a major breakthrough in ammonia technology. In 1967 world production of nitrogen was approximately 38 million metric tons (8). Projections for 1977, including plants presently under construction, indicate that capacity will exceed 77 million metric tons representing more than a two-fold increase over a 10 year period. This sizable increase in capacity is being brought about largely by the construction of plants in the range of 600–1,700 short tons/day, the bulk of which are based on use of light hydrocarbon feedstock in conjunction with

catalytic steam reforming. Most new construction centers on plants of 1,000–1,500 short tons/day capacity, and the indications are that this capacity range will apply in the near future for all feedstocks including heavy fuel oil and coal, primarily because it represents the bulk of commercial experience and appears to be an economic size. There will, of course, be some interest in giant plants having capacities of 2,000 short tons/day and greater, but these will be isolated cases and certainly will be associated with cheap natural gas feed.

Several conclusions can be drawn with respect to the use of the various feedstocks considered in this article.

1. Most, if not all, of the ammonia production in North America and Europe will be based on natural gas feed. If the cost of natural gas approaches $1.00/MM B.t.u., other feedstock may be given serious consideration, including heavy oil feed and coal.
2. Shipment of ammonia from areas where natural gas is abundant appears to be highly attractive. Ammonia or ammonia based products can be economically exported from the Middle East, North Africa, and other low cost feed locations to high cost areas. Barring international complications, shipment of nitrogen to the U.S., Europe, and other locations will play an increasingly important role in worldwide distribution. Shipment of bulk ammonia will be dependent on carrier limita-

Table 4. Steam reforming vs. fuel oil partial oxidation.

Capacity: 1,000 short ton/day of ammonia

Comparison with Natural Gas @ $1.00/MM B.t.u.

Fuel Oil Cost $/MM B.t.u.	($/short ton)	Price Difference $/MM B.t.u.	Total Operating Natural Gas[2]	Costs $/short ton[1] Fuel Oil[2]	Savings with Fuel Oil $/short ton	Simple Payout Savings with Partial Oxidation over 3 years[3]
0.29	(10)	0.71	33.80	12.54	21.26	$21,000,000
0.58	(20)	0.42	33.80	22.19	11.61	11,500,000
0.87	(30)	0.13	33.80	31.84	1.96	2,000,000
1.00	(34.50)	0	33.80	36.19	(−2.39)	—

Comparison with Naphtha @ $1.00/MM B.t.u.

			Naphtha[2]			
0.29	(10)	0.71	36.80	12.54	24.26	24,000,000
0.58	(20)	0.42	36.80	22.19	14.61	14,500,000
0.87	(30)	0.13	36.80	31.84	4.96	5,900,000
1.00	(34.50)	0	36.80	36.19	0.61	600,000

Notes:
1. Includes feed, utility, catalysts, chemicals, and operating labor.
2. Refer to Tables 1, 2, and 3 for details on economic calculations.
3. Savings with partial oxidation process calculated on basis of 330 operating days per year over three year simple payout period.

Table 5. Steam reforming vs. coal gasification.

Capacity: 1,000 short ton/day

Comparison with Natural Gas @ $1.00/MM B.t.u.

Coal Cost $/ton	Feed & Fuel Costs, $/short ton Natural Gas[1]	Coal	Savings with Coal, $/short ton	Simple Payout Savings with Coal over 3 years[2]
5	31	10	21	$20,790,000
10	31	20	11	$10,890,000
15	31	30	1	$ 990,000

Comparison with Natural Gas @ $2.00/MM B.t.u.

5	62	10	52	$51,480,000
10	62	20	42	$41,580,000
15	62	30	32	$31,680,000

Comparison with Naphtha @ $1.00/MM B.t.u.

Naphtha (1)

5	34	10	24	$23,760,000
10	34	20	14	$13,860,000
15	34	30	4	$ 3,960,000

Comparison with Naphtha @ $2.00/MM B.t.u.

Naphtha (1)

5	68	10	58	$57,420,000
10	68	20	48	$47,520,000
15	68	30	38	$37,620,000

Notes:
1. Refer to Tables 1 and 2 for details.
2. Savings with coal gasification on the basis of 330 operating days per year over a three year simple payout period.

tions and possibly on associated safety hazards.

3. Use of fuel oil and other heavy hydrocarbon feeds with the partial oxidation process will increase to a greater degree than previously, primarily because of the increasingly favorable price structure on these feeds. If the natural gas price in any given location exceeds the fuel oil price by a significant degree, fuel oil will be adopted as ammonia plant feedstock.

4. Greater emphasis will be placed on use of liquid fuels for firing both the reformer and auxiliary boilers in steam reformer units.

5. Coal based operations can be justified in locations where the unit cost of coal is low, where gas does not exist, and where the alternative is expensive imported oil, as in South Africa and India. Moreover, use of such feed will be governed to a great degree by whether an attractive price differential between solid and hydrocarbon feedstocks can be maintained on a long term basis. However, the high costs of mining and transportation charges will of course reduce any potential economic advantage inherent in use of solid feeds.
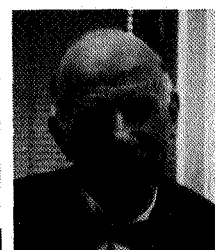
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