



VOLUME 24

# AMMONIA PLANT

# SAFETY

(and related facilities)

Prepared by editors of  
CHEMICAL ENGINEERING PROGRESS



VOLUME 24

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Prepared by editors of  
CHEMICAL ENGINEERING PROGRESS

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## **INTRODUCTION**

This symposium is a forum where those involved with the manufacture, storage, and transportation of ammonia and related chemicals can share knowledge, experience, and new technology related to safe and efficient plant operation. It provides a means of improved communications among plant operators, contractors, and engineers to improve plant operability and safety.

We are grateful to the authors of the papers printed in these manuals and to their employers for the success of this symposium. Thanks for sharing this important information with the ammonia industry.

**AICHE Symposium—Denver, 1983**

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**John Livingstone, Program Vice-Chairman**

**Committee Membership—August, 1983**

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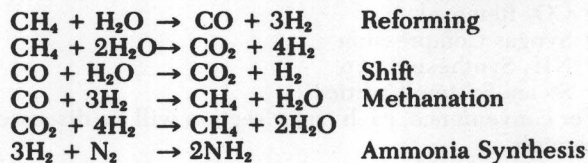
# Creative Ways of Revamping NH<sub>3</sub> Plants Can Improve Profitability

**A blow-by-blow case history account of the actual revamping of an existing ammonia plant.**

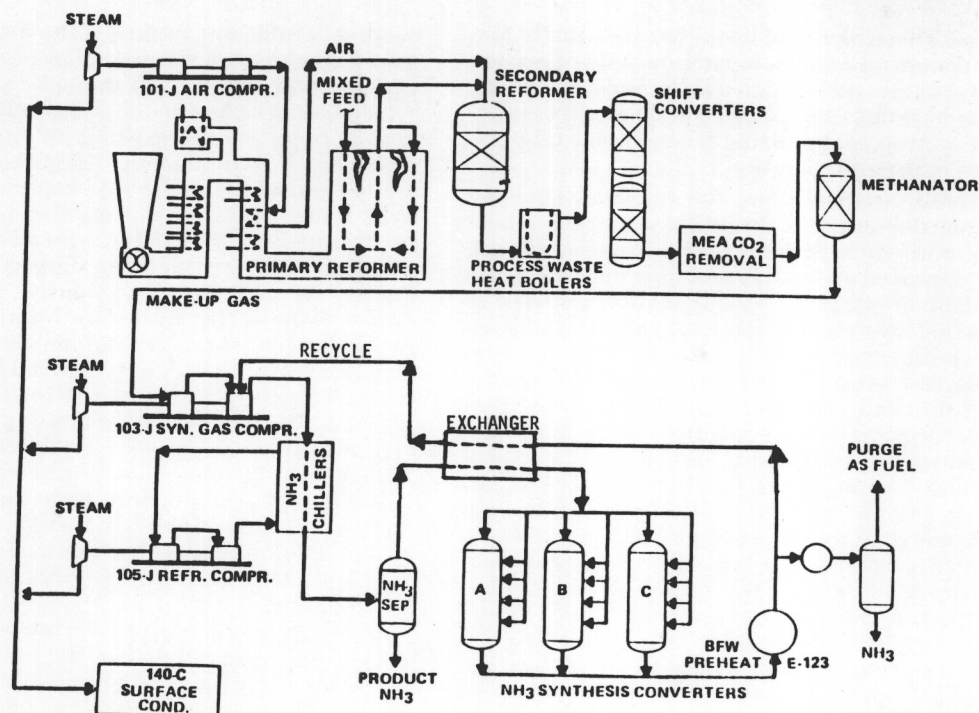
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About 70 to 80% of the world ammonia production is based on a well established conventional steam-methane or steam-naphtha reforming front end. Typical process steps consist of feedstock desulfurization, high-pressure steam introduction, preheating, and primary reforming with a nickel catalyst (Figure 1). Preheated air is then introduced into a secondary reformer to provide nitrogen for the later NH<sub>3</sub> synthesis stage and to provide oxygen to maintain the temperature of the reforming reaction. This gas is further processed in carbon monoxide shift converters and purified to form ammonia-synthesis gas which is then

converted to ammonia and removed as a liquid by refrigeration. The following are the major reactions which occur in reforming, shifting, methanation, and ammonia synthesis.



Existing ammonia plants typically require about 43 to



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Figure 1. Air Products Nola  $\text{NH}_3$  plant.

48 GJ/MT (gigajoules per metric ton) as energy input, while new plants can be designed to achieve energy levels close to 35 GJ/MT or as low as 32 GJ/MT [1]. High energy costs in the last decade have created the incentive to re-vamp existing  $\text{NH}_3$  plants in order to reduce the overall energy consumption and improve profitability. There are many ways to conserve energy and increase production in an existing ammonia plant. Numerous papers published by contractors and manufacturers detail a number of ideas on the subject matter [1, 2, 3, 4]. This paper presents typical limitations in each major section of the plant and discusses possible alternates to overcome these limitations to increase plant capacity and efficiency. In addition, this paper also discusses the Air Products and Chemicals, Inc. (APCI)  $\text{NH}_3$  plant bottlenecks and shares some successful experience in revamping to overcome bottlenecks associated with capacity expansion, and overall performance efficiency/reliability improvements.

#### TYPICAL $\text{NH}_3$ PLANT LIMITATIONS/SOLUTIONS

Typical  $\text{NH}_3$  plants consist of the following major sections as shown in Figure 1.

- Primary Reformer Furnace
- Secondary Reformer/Process Boilers
- Shift Conversion
- $\text{CO}_2$  Removal
- Syngas Compression
- $\text{NH}_3$  Synthesis Loop
- Steam System/Utilities

For convenience, each major section will be discussed separately.

##### Primary Reformer Furnace

The primary reformer furnace, consisting of critical high temperature process equipment, is the most important section of an  $\text{NH}_3$  plant. It is the biggest energy user since essentially all of the hydrocarbon feed and fuel are

consumed in the reforming section. Because of its nature, the primary reformer is often one of the major areas limiting plant capacity. Possible limitations can be either in draft, reformer tubes, or convection coils.

**Draft.** To increase production more firing duty is required. This requires more combustion air accompanied by higher flue gas generation. Therefore, due to increased flow rate (and thus higher pressure drop), the induced and/or forced-draft fan can easily become a plant bottleneck.

To overcome this limitation three possible approaches are available. The first approach is to increase the speed of the fans, if they are not driver limited, by modifying the rotor blade design (larger wheel diameter). This can provide an additional 5 to 10% capacity without requiring any major changes around the duct. The second approach is to add a small supplemental fan in parallel to the existing one. This can be done if the driver itself is limited, if space is available, and if duct modification is required. The third approach is to consider replacing the existing fan with a new larger unit if the first two approaches are not possible.

Furthermore, to conserve energy and improve the reformer furnace efficiency, one can consider addition of BFW preheat and/or steam generating coils in the convection section (depending on the overall steam balance and space available). APCI experience indicates that the additional pressure drop through these new coils can be compensated by lower horsepower required in the induced draft fan due to cooler stack gas.

**Reformer Tubes.** The reforming reaction involves a coupling of heat transfer and a strongly endothermic reaction. The balance between heat input through the reformer tubes and the heat consumption by the reaction (radiant efficiency) makes the tube the most critical item limiting  $\text{NH}_3$  production [5].

To increase production, more syngas is needed, and chances are that operating personnel have tried their best to increase production by utilizing different modes



of operation. These modes of operation are usually limited by maximum tube skin temperature and/or pressure drop of the reformer tubes. Increases in capacity can easily increase heat flux (maximum tube skin temperature) and pressure drop, jeopardizing the reformer tube life and overall reformer reliability.

Under these circumstances, the question becomes how does one debottleneck the primary reformer tubes? Obviously, one way to debottleneck is to use larger size tubes and upgraded material (depending on space available in radiant harps) if the existing set of tubes have already reached the end of their service life and are planned to be changed. With this change, the natural gas throughput can be increased while maintaining the same heat flux and expected tube life. On the other hand, if the tubes are in good condition, with no plans to replace them in the near future, one can consider one or both of the following options to increase natural gas throughput. The first option is to shift some of the primary reformer duty to the secondary reformer in conjunction with increased natural gas feed throughput and lower steam/carbon ratio. The second option is to replace the existing catalyst with new generation, high activity, and low pressure drop catalyst.

**Convection Section.** In a typical  $\text{NH}_3$  plant, the convection section heat recovery normally consists of mix feed preheat, secondary reformer air preheat, steam generation, and superheat, BFW preheat, N. G. feed and fuel preheat, etc. Out of these coils, the mix feed, secondary air, and steam superheat coils are usually severely deteriorated (exceeded design temperature) due to continuous long-term operation under extreme high temperature and refractory fouling. This operation can sometimes result in distorted coil supports and/or bowing of tubes. Poor degraded conditions of these coils result in lower preheat temperatures and high stack losses.

Any further increase in plant capacity requires more utilities (to drive compressors) and increases reformer firing rate. Increasing firing rate in an existing furnace lowers the radiant efficiency and sends hotter flue gas through the convection section coils. This results in overheating of these coils, lower preheat temperatures, and much more stack losses.

To overcome this problem, it is worthwhile to replace the mix feed, secondary air, and steam superheat coils with coils having more surface area, larger tube sizes, and upgraded metallurgy. This will also improve the overall plant reliability and efficiency. Increased mix feed temperature can save reformer firing fuel. Increased air preheat temperature improves secondary reformer performance, resulting in lower methane leakage. Increased steam superheat temperature, with lower pressure drop across the coil (utilizing larger diameter tubes), increases the available HP to the compressor turbines.

To further improve overall furnace efficiency by recovering excess stack waste heat, one can also consider preheating combustion air. Depending on the type of reformer furnace, the number of burners, and the extent of duct work required, the combustion air preheat may require considerable downtime. This change could also easily be an expensive project because modifications are needed to the forced draft fan as well as the burner.

#### Secondary Reformer/Process Boilers

In a conventional  $\text{NH}_3$  plant, the methane leakage from the primary reformer is usually in the range of 9 to 12%. To further reform this methane and bring in an equivalent amount of nitrogen for  $\text{NH}_3$  synthesis, secondary reforming is required. To ensure good performance of the secondary reformer, the following requirements are

needed: a sufficient air flow at the desired temperature, a good burner design for mixing (air plus reformed gas), and a proper flow distribution through the catalyst bed.

Depending on how much additional secondary reformer duty is required, the existing air compressor (or driver) might reach its limitation. Additionally, the convection section coil might not be able to preheat the air to the desired temperature, or the existing burner might be too small. Any one of the above mentioned limitations can contribute to higher methane slippage at the secondary reformer. As mentioned previously, shifting primary reformer duty to the secondary reformer compounds the problem even more. Desired methane slippage in a secondary reformer with a conventional  $\text{NH}_3$  plant ranges between 0.2 to 0.4%. Methane is an inert in the synthesis loop and must be purged out constantly. Therefore, any increase in methane slippage at the secondary outlet could increase the amount of the required purge flow and could lower the  $\text{NH}_3$  production. For example, an increase in methane slippage from 0.3 to 0.5% at the secondary reformer effluent lowers overall  $\text{NH}_3$  production by as much as 16 to 22 MT/SD (metric ton per standard day) in an 1100 MT/SD plant.

There are a number of possible ways to overcome the above discussed limitations in lowering methane slippage. These include: 1) adding a supplemental air compressor in parallel to debottleneck the available air capacity, 2) increase secondary air preheat capability (convection coil modification), 3) redesign the burner tip for good distribution, and 4) maintain a proper residence time for sufficient mixing of combustion products.

To avoid creating problems for downstream waste heat boilers, the secondary reformer outlet temperature should be maintained close to design temperature. However, with increased process flow, the heat flux and overall steam generation in these boilers increases. In that respect, the hydraulics of the downcomers and risers, circulation ratio, heat flux of the tubes, pump capacity (for forced circulation steam system), and steam drum retention time should be rechecked and modified accordingly if needed.

#### Shift Conversion

As a result of reduced steam to carbon ratio and increased secondary air injection, production of carbon monoxide increases in the reformers, requiring more shift capacity. Possible limitations for both the HTS and LTS can be the higher space velocity, resulting in higher CO slippage. The most economical way to overcome this limitation is to add more catalyst in the existing shift vessels if space is available or increase steam injection before each shifter. Other alternatives can be the installation of an LTS guard bed with quenching in between, or consideration of the SELECTOXO process [6] to convert excess low level CO to  $\text{CO}_2$ .

#### $\text{CO}_2$ Removal Area

The amount of  $\text{CO}_2$  required for removal is proportional to the plant capacity. Depending on the type of  $\text{CO}_2$  removal processes (MEA, hot carbonate, or Selexol), many debottlenecking ideas exist. Most plants using MEA or hot carbonate have added inhibitors to prevent corrosion and allow them to increase solution strength. One economical way of debottlenecking the  $\text{CO}_2$  removal capabilities for MEA and hot carbonate systems is by increasing the circulation rate and solution strength. In some cases, modifying column internals using high mass transfer packing instead of trays can also increase capacity and efficiency [7]. For debottlenecking with the Selexol process, one can consider increasing the circulation rate and/or adding a refrigeration exchanger to cool the lean solvent.

Other limitations in this area could be the MEA reboiler, overhead condenser, and/or other exchangers limiting stripping capabilities. These limitations can be overcome by installing supplemental parallel heat exchangers.

### Syngas Compressor

With increased syngas flowrate, the pressure drop across the front end (reformers, process boiler, shifters, absorber, methanator, and exchangers) increases, resulting in lower suction pressure at the syngas compressor. The combination of higher flow and lower suction pressure increases the duty requirement of this syngas compressor. Possible ways to increase the compression capacity in this machine depends on whether the limitation exists in the compressor itself or in the driver.

If the compressor is limited, modifying the wheel size and/or increasing its speed should be considered before replacing it with a larger unit. If the turbine is limited, the following three options can be considered. The first option is chilling the make-up gas at the suction and interstage to reduce horsepower requirements. The second option is installing a larger nozzle block and/or further increasing the steam superheat temperature/pressure. This increases the turbine power available to the compressor if the machines have not reached the maximum trip speed. The third option is modifying the synloop to reduce recycle flow and/or adding a separate circulator to compress the recycle stream.

If none of the above options are viable due to space limitations, a supplemental syngas compressor in parallel can be considered.

### Synthesis Loop

Producing more  $\text{NH}_3$  requires a higher conversion capacity and more refrigeration duty to condense the additional ammonia product. The question becomes how can this be done economically? One way to increase ammonia production in the synthesis loop is to install a purge gas recovery unit, such as cryogenic or selective permeation techniques to reject inerts and recover hydrogen that can be introduced back into the loop. Another possible alternative is to install additional  $\text{NH}_3$  condensing capabilities to lower the  $\text{NH}_3$  concentration entering the synthesis converter. This increases the overall converter yield (i.e.,  $\text{NH}_3$  produced per pass). A second alternative is to improve quench flow capabilities for quench type converters to achieve closer temperature control in the converter. This minimizes the converter temperature approach to equilibrium and increases yield per pass. A third alternative for certain types of synloops (Figure 1) is to install a molecular sieve dryer to remove water and  $\text{CO}_2$  from the fresh make-up gas. The synloop would also have to be repiped such that the make-up gas can be mixed with recycle gas and fed directly to the converter without having to first pass through a set of refrigeration exchangers. Thus, the refrigeration duty can be bottlenecked because the make-up gas does not need to be cooled and heated-up again before entering the converter. Additionally, higher yields are obtained due to a higher pressure and a lower  $\text{NH}_3$  concentration at the converter inlet.

If none of the above alternates are feasible, one can consider replacing the synthesis loop with a radial flow converter with steam generation capabilities.

### Steam System

To improve ammonia plant capacity and efficiency it is very important to initially have a good steam balance (showing all the different usages). In a conventional  $\text{NH}_3$  plant high-pressure superheated steam is utilized to drive

most of the compressors and pumps. Thus, steam consumption is the next major energy consumer after the primary reformer furnace.

As the plant capacity increases, more steam generation capabilities are required. Possible ways to lower steam consumption should be considered in cases where additional steam generation is limited and/or to conserve energy. The most economical and a very effective way to do this is to try to achieve better vacuum in the condensing turbines' surface condenser. Decreasing the vacuum by a few inches can significantly lower the amount of steam required to produce the same BHP in the turbines.

One alternate method is optimization of a different level pressure steam system. This can be accomplished by avoiding any high-pressure steam let down to lower level usage and by considering electric drives, if needed. Another alternate is utilizing low-pressure steam for process use wherever possible. An additional alternate is enlarging the steam turbine exhaust lines to reduce pressure drop and lower turbine water rates.

### Off-Site Utilities

The major off-site utilities for an ammonia plant include the cooling water system, make-up water, and electric power. When the plant capacity is increased, more cooling water for compressors and process coolers are needed. If the existing cooling tower is capacity limited, it can be expanded by the addition of a cell and pump. Additional make-up water and electric power may also be needed and their availability should be checked. Finally, one also needs to make sure that all the environmental permits are up to date with current plant production rates.

Depending on the process and equipment limitations of each individual plant, all these ideas may or may not be practical, especially for a plant 15 or more years old. Revamping an existing  $\text{NH}_3$  plant takes a lot of creative thinking, to plan appropriate designs which can provide maximum benefit and minimal capital expenditure. To achieve this, it is necessary to understand each plant's overall plant performance and individual equipment limitation.

### HOW WE REVAMPED

Over the years, APCI has been able to operate its New Orleans, La.,  $\text{NH}_3$  plant significantly above the design capacity without sacrificing the original design efficiency. To achieve this high rate, an additional  $\text{NH}_3$  synthesis converter and an auxiliary boiler generating 42.36 bar steam were added in the early years of operation. However, there were still several other major pieces of equipment limiting the plant capacity at different times of the year. The air compressor limited plant capacity throughout the year, but most predominantly during the warm months. During cold months, the primary reformer pressure drop limited the plant capacity by lowering the suction pressure of the syngas compressor. In March and April, the induced draft fan was the main plant constraint limiting the reformer furnace firing duty. In addition, our primary reformer tubes operated at an average heat flux of 118.3  $\text{kw/m}^2$  with a pressure drop of about 5.5 bar, compared to typical industrial operation at an average heat flux of 63.1  $\text{kw/m}^2$  and a pressure drop of 2.75 bar. Under these extreme constraints, it is very difficult to further expand the capacity and improve overall operating efficiency without major capital expenditure.

To make the revamp project attractive, creative and inexpensive ideas were needed. Here is how the  $\text{NH}_3$  plant was revamped to further increase the capacity and improve its overall efficiency.



## Plant Performance Test

To begin with, an extensive overall plant performance test was conducted to obtain actual plant operating data. This generated a meaningful heat and material balance for the entire  $\text{NH}_3$  plant (including reformer front end, synthesis loop,  $\text{NH}_3$  refrigeration circuit, steam system, etc.) at maximum production/efficiency mode of operation. Actual raw plant data were evaluated by computer simulation using our proprietary modular process simulator, cycle synthesis [9]. This program can perform complicated process calculations for the entire plant. Each portion of the process is modeled on the computer. This includes the following equipment: primary/secondary reformer, compressors, furnace, heat exchangers, high and low shift reactors,  $\text{CO}_2$  removal system, synthesis loop, etc. The various process units are put together into a process simulation flowsheet that models the entire plant including recycle streams.

After having established a detailed heat and material balance, including a steam/water balance, the performance of the major pieces of equipment were evaluated in order to identify plant "bottlenecks and inefficient areas."

From the plant performance test, the following plant bottlenecks were determined:

- Furnace induced draft fan
- Air compressor capacity
- Product  $\text{NH}_3$  condensing chillers
- Primary reformer
- Syngas compressor
- Refrigeration compressor

The following inefficient areas were also identified:

- Surface condenser
- Convection section
- Steam system

## Plant Modifications

Once the basis of the existing plant was fully established with its bottlenecks and inefficient areas identified, different ideas to revamp each limiting area were developed and carefully evaluated.

### Primary/Secondary Reformer

As mentioned previously, the primary reformer capacity is limited by pressure drop and heat flux. The best way to debottleneck it and increase capacity is to shift some of the duty to the downstream secondary reformer. In this respect, the steam to carbon ratio was lowered in the primary reformer from 3.85 to about 3.3. Consequently, the natural gas feed throughput was increased by an equivalent amount. With this modification, 8% more feed was introduced into the mass flow limiting primary reformer. The reformer furnace firing was held the same to maintain identical heat flux and expected reformer tube life. This increased the methane slippage at the primary reformer effluent from 10% to 13%.

An additional modification was made in the primary reformer. Here, space was available (in the radiant harps) at the area of the subassembly to subassembly welds, permitting the addition of 10 more tubes. This further debottlenecked the pressure drop limitation and increased the primary reformer throughput.

Higher methane slippage from the primary reformer generally results in higher methane slippage from the secondary reformer. The question becomes—How does one adjust the secondary reformer to achieve the same desirable methane leakage? One way is to introduce more secondary reformer air to reform the unreacted methane that slipped from the primary reformer. But, unfortunately, the secondary air compressor, as stated previously, was one of

the major plant bottlenecks. Therefore, the most economical way to debottleneck the air compression capacity was to install a new supplemental air compressor in parallel to the existing one. With a new 1491 kW electric driven centrifugal air compressor installed, sufficient air capacity was provided for the secondary reformer.

As the secondary reformer air flow increases, the duty requirement for the convection section air preheat coil also increases. But it was determined that all three coils (mix feed preheat, secondary air preheat, and steam superheat) in the hot portion of the convection section were severely deteriorated. This was due to continuous long-term operation under extreme high temperature services of refractory fouling. Poor degraded conditions of these coils resulted in lower preheat duties accompanied by higher stack losses. Therefore, to provide for adequate duty requirements and improve the reliability of these three coils, the coils were replaced with better designed (minor modifications) and upgraded material. For the steam/natural gas mixed feed preheat coil, no changes in the tube size or surface area were necessary to accommodate the new process condition.

To further increase the secondary reformer air preheat temperature from 480 to 540°C (to optimize secondary reformer performance), one more row of tubes was needed. The material of this coil was also upgraded accordingly. Increased air capacity, coupled with higher preheat temperature, improved secondary reformer performance, and achieved the desirable methane slippage of about 0.25%. The amount of primary reformer duty that could be shifted over to the secondary reformer (by lowering the steam to carbon ratio in the primary reformer and increasing the secondary reformer air flow) depends on individual plant operation, syngas H/N ratio, and overall process efficiency.

The steam superheat coil was completely redesigned with more heat transfer surface area and lower pressure drop through the coil. This increased the available HP to the topping turbine of the syngas compressor, which enhanced the syngas compression capability. During the revamp, a fix up procedure was followed to minimize down time requirements and also to accommodate the change out during the outage. This fix up procedure included cutting the top and one end wall out of the convection section, removing the old coils, and replacing them with a better design, making use of the existing steel structures.

### Shift/ $\text{CO}_2$ Removal

As a result of reduced steam to carbon ratio in the primary reformer and increased air flow in the secondary reformer, the carbon monoxide mole fraction in the reformers increased significantly. This required additional shift capacity. After carefully evaluating several options, the most economical and practical option for the plant was to install a low temperature shift (LTS) guard bed with intermediate quenching. Thus, a desirable  $\text{CO}$  leakage (0.2%) exiting the shift converters was achieved. Installing the LTS guard bed served a dual function. It reduced the  $\text{CO}$  leakage and prolonged the LTS life to as much as 5 years.

Debottlenecking of the  $\text{CO}_2$  removal section involved the following procedure. Since the  $\text{CO}_2$  loading is proportional to the  $\text{NH}_3$  production, the MEA strength and circulation rate were increased to handle the additional  $\text{CO}_2$  loading. The  $\text{CO}_2$  removal section was simulated, based on the new process conditions. All heat transfer equipment and circulation pumps were evaluated and found to be acceptable. A hydraulic test of the system at a higher MEA circulation rate was performed and there was no indication of foaming, flooding, or deficiency in  $\text{CO}_2$  removal capabilities.

## Syngas Compression

As discussed previously, the syngas compressor was the bottleneck. With a higher syngas flow rate and a lower suction pressure at the syngas compressor, the compression horsepower requirement of the syngas machine increased further. The problem then becomes how to overcome this. The work of isentropic compression can be calculated by the following equation [10]:

$$W = \frac{RT}{n} \left[ \left( \frac{p_2}{p_1} \right)^n - 1 \right]$$

where:

$W$  = work

$R$  = gas constant

$T$  = absolute temperature at suction, °K

$p_1$  = absolute pressures at suction

$p_2$  = absolute pressure at discharge

$n = \frac{\gamma-1}{\gamma} \left( \gamma = \frac{C_p}{C_v} \right)$

$C_p$  = specific heat at constant pressure

$C_v$  = specific heat at constant volume

As shown in this equation, the compression horsepower is proportional to the gas temperature (°R) at the suction of the compressor. In the plant situation, the most economic way to overcome this problem was to add a chiller to cool the make-up gas from 40°C to 5°C. As mentioned earlier, this machine always operated near maximum speed. Therefore, reducing the suction temperature compensated for the higher mass rate and lower suction pressure. To further increase the turbine driver horsepower availability, a larger steam turbine nozzle block was installed resulting in the turbine operating near maximum speed.

To further debottleneck the syngas compression capacity, an interstage refrigeration chiller was brought back into service. This chiller had been taken out of service several years back to allow some additional water to remain in the synthesis gas and be condensed with product  $\text{NH}_3$  (for  $\text{NH}_3$  loading safety reasons).

## $\text{NH}_3$ Synthesis Loop

One obvious limitation in the synthesis loop was the  $\text{NH}_3$  condensing heat exchanger. This heat exchanger became too small to condense the  $\text{NH}_3$  as designed due to lubricating oil fouling. This resulted in a higher  $\text{NH}_3$  concentration in the converter, which reduced the  $\text{NH}_3$  conversion (or yield). Since the compression was limited, a lower conversion rate meant a higher purge rate. Thus, capacity and efficiency were affected. This problem would be compounded with further increases in capacity. The logical solution was to install a supplemental  $\text{NH}_3$  condensing heat exchanger parallel to the existing one. This increased overall  $\text{NH}_3$  production capacity and efficiency.

Another idea for increasing capacity and efficiency in the synloop area is to recover hydrogen from the purge stream. Recovering this hydrogen as a chemical value rather than as boiler fuel improves efficiency. The make-up gas contains about 0.6% of  $\text{CH}_4$  and 0.4% of argon. These inerts build up in the synloop and require a continuous purge. The purge stream contains from 62 to 63% hydrogen. As mentioned previously, in order to debottleneck the primary/secondary reformer, more air (nitrogen) was introduced into the secondary reformer, lowering the H/N ratio to much less than 3.0. To recover the hydrogen, to reject inerts, and to balance the H/N ratio in the synloop, a hydrogen recovery unit was installed.

Increased  $\text{NH}_3$  production required more refrigeration duty to condense the additional  $\text{NH}_3$ . Thus, a 450 kW screw type supplemental compressor, integrated with the existing refrigeration system, was installed. This avoided a

refrigeration system imbalance in the synthesis loop and also provided the new make-up gas chiller refrigeration requirement.

## Steam System

The three main compressors (syngas, air, and refrigeration) in the ammonia plant are steam turbine driven. Each has a condensing type steam driver that exhausts steam into a common surface condenser. Vacuum losses of 130 to 150 mm Hg during the hot summer months have been observed as a result of increased plant capacity and plugging of some leaking tubes. During the plant performance test, this 140°C surface condenser was also identified as one of the major inefficient areas. A recommendation was made to upgrade its performance to save steam (a reduction of the amount of steam required to produce the same power).

Several alternates to upgrade the surface condenser were considered, including: adding a supplementary parallel unit, retubing the existing condenser with better material, or making a complete replacement with a larger unit. Because of the space limitation, the ease of operation, and the short down time requirements, complete replacement with a larger unit was selected.

The new condenser was designed to fit the existing mounting platform, with the same inlet and outlet nozzle locations. To accommodate the additional heat load the cooling tower water supply and return piping were modified accordingly.

Present operation of this surface condenser is much better than design, with slightly higher than 710 mm Hg vacuum being achieved. Another energy saving modification included installation of a larger nozzle block on the condensing turbine of the syngas compressor. This permitted more medium level steam to flow through the condensing turbine, requiring equivalently less high-pressure steam for the topping turbine.

## CONCLUSIONS

High energy costs have created a need to revamp existing ammonia plants to reduce the overall energy consumption per unit ton of production. Evaluating each limiting area of the ammonia plant takes a lot of creative thinking to plan appropriate designs which can provide maximum benefits with minimal capital expenditure. Understanding the overall plant performance and each individual equipment capability is a must for a successful, reliable revamp project to improve profitability of the  $\text{NH}_3$  plant. Hopefully, these general guidelines would assist the selection of possible alternates to effectively overcome the plant limitations, to increase capacity and efficiency.

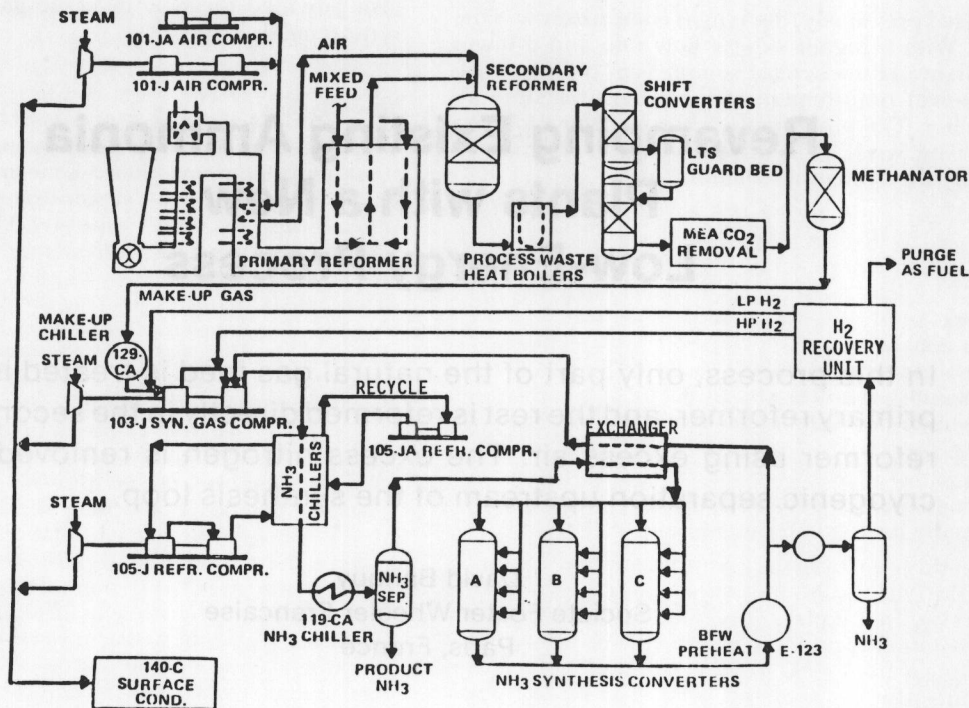
In revamping APCI's ammonia plant, an aggressive energy conservation/capacity expansion project was developed. The plant expansion included lowering the steam to carbon ratio in the primary reformer, radiant tubes addition, purge gas recovery, incremental secondary reformer air capacity, and additional ammonia condensing surface. The energy conservation included a new large-surface condenser for the major compressor turbine exhausts, a revamp and optimization of the hot convection section, and a primary/secondary reformer operating conditions adjustment (Figure 2).

Implementing the above mentioned projects increased the overall ammonia production by about 165 MT/SD and saved almost 35,170 kW in energy consumption.

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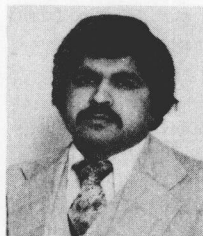
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Figure 2. Air Products Nola  $\text{NH}_3$  plant after revamp.

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# Revamping Existing Ammonia Plants with a New Low-Energy Process

In this process, only part of the natural gas feed is treated in the primary reformer, and the rest is reformed directly in the secondary reformer using excess air. The excess nitrogen is removed in a cryogenic separation upstream of the synthesis loop.

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## INTRODUCTION

Practically all the existing ammonia plants built in the last 20 years, based on a natural gas feedstock, use primary steam reforming followed by secondary reforming with air, the latter to achieve the stoichiometric amount of nitrogen required for the synthesis of ammonia. Most of these existing plants have an overall energy consumption higher than 30 MMBtu/ST (8.35 MMKCAL/MT) of ammonia and sometimes much higher. Due to the escalation of energy costs in recent years, this energy consumption has become a major factor in the overall production cost of ammonia, and needs to be thoroughly investigated to find improvements which would lead to possible reductions.

## CAUSES OF HIGH ENERGY USE IN EXISTING PLANTS

In these existing conventionally designed ammonia plants the following three major factors lead to high energy consumption :

- 1) The use of the stoichiometric amount of air in the secondary reformer limits the amount of reforming that can be done in this reformer. This increases the amount of reforming required in the primary which results in higher fuel consumption.

- 2) The need to obtain a low residual methane content from the secondary reformer (in the range of 0.2. to 0.5 Mol. per cent) in order to limit the amount of purge in the synthesis loop. Because of this requirement an appreciable amount of reforming must be carried out in the primary reformer, since the secondary is limited by the stoichiometric amount of air.
- 3) The need for a minimum amount of process steam in the primary reformer to assure a safe operation without carbon formation, and to achieve the desired methane content at the outlet of the secondary reformer.

The above three factors are the barriers which must be overcome in order to achieve a meaningful reduction in energy consumption for these existing plants.

## NEW LOW ENERGY PROCESS FOR REVAMPS

Many years ago C.F. BRAUN (1) introduced a process whereby some excess air is used in the secondary reformer, and the excess nitrogen is removed from the final synthesis gas, just upstream of the synthesis loop, by cryogenic separation.



In this process, due to the relatively small amount of excess nitrogen used in the secondary reformer, the refrigeration requirements for cryogenic separation are supplied by passing the main synthesis gas stream through a turbo-expander, incurring some pressure drop. This process was a some progress with respect to items 1 and 2 above.

In the new low energy process described in this paper we have gone much further on the excess air used in the secondary reformer, so that the amount of excess nitrogen removed in the cryogenic separation is sufficient to provide, by a Joule-Thomson expansion, most if not all of the refrigeration requirements for the separation. Furthermore, we have reduced substantially the amount of process steam required by treating in the primary reformer only a small fraction of the total feedstock, the rest going directly to the secondary reformer, after mixing with the effluent from the primary reformer.

#### BASIC FEATURES OF NEW LOW ENERGY PROCESS

Accordingly, the new low energy process which we have patented (2) in 1980, includes the following basic features which are represented schematically in Fig.1 and summarized in Table I :

- 1) The fraction of the feed that is treated in the primary reformer is in the range of 20 to 50% of the total feed.

Obviously, the lower this fraction, and the lower will be the process steam requirements. There is however a limit, because a minimum amount of hydrogen should be present in the mixture entering the secondary reformer, in order to prevent carbon formation in the gas phase reaction that occurs when mixing with air. This minimum amount of hydrogen depends itself on the general process parameters, and in particular on the amount of hydrocarbons heavier than methane.

On the other hand, if the feedstock contains a large fraction of heavier hydrocarbons, it is preferable to treat at least 50 per cent of the total feedstock in the primary reformer, in order to reduce the partial pressure of the heavier hydrocarbons in the mixture entering the secondary reformer.

- 2) The primary reformer operates at a very mild temperature, in the range of 680 to 760°C, and more preferably around 700°C. This represents a large savings of fuel for the existing plant.

The lower temperature range would be usually the case when a large fraction of the feed is treated in the primary reformer, and vice versa.

- 3) Due to this low temperature of the primary reformer, the pressure throughout the synthesis gas train may be much higher than used up to now; a pressure in the range of 40 to 70 atm is quite possible mechanically, using the same materials of construction as presently used in the industry but, in a revamp, of course, the existing pressure must be matched.

- 4) The excess process air in the secondary reformer is such that the molal ratio  $H_2/N_2$  at the inlet of the cryogenic separation is in the range of 1.0. to 1.8. This corresponds to an excess air of 65 to 200% over the stoichiometric amount.

- 5) Since all the methane in the raw synthesis gas is removed in the cryogenic separation, there is no need to have a very low  $CH_4$  content at the outlet of the secondary reformer : a content of 2 to 5% Mol is considered satisfactory, and possibly higher concentrations may be advantageous. In this regard, the limit would be when the potential heat liberation of the residual gas from the cryogenic separation is just equal to the heat liberation requirements of the primary reformer.

- 6) Since we have an appreciable amount of low level heat in the gas effluent from the low temperature shift converter, when using a physical solvent for  $CO_2$  removal, it is very desirable to use part of that low level heat to humidify the process air, as well as the process natural gas. A conventional closed loop of hot water circulation may be used for this purpose.

- 7) As the synthesis gas train can operate at high pressure, say above 40 atm, it is found very advantageous to use a physical solvent for  $CO_2$  removal, such as Selexol. In such a case, and if the  $CO_2$  is not recovered for urea production, the regeneration of the solvent may be carried with air stripping.



- 8) When using a physical solvent for CO<sub>2</sub> removal, there is enough heat in the effluent from the LTS converter to supply heat for an absorption refrigeration system, covering the needs of the synthesis loop, as well as, the needs of the CO<sub>2</sub> removal system.

To make the best use of the low level heat available after the LTS converter, whether for a new plant or for revamping a existing plant, it may be advantageous to have two adsorption refrigeration systems:

- an ammonia absorption system using the heat in the 100-130°C range and supplying refrigeration below 0°C.
  - a lithium bromide or ammonia system using the heat in the 70-100°C range and supplying refrigeration at about 7°C.
- 9) The cryogenic separation produces the final synthesis gas with a stoichiometric composition, and a residual gas containing the excess nitrogen, all the methane and a great part of the argon contained in the raw synthesis gas. This residual gas supplies most of the fuel requirements of the primary reformer.
- 10) In case a small amount of purge needs to be withdrawn from the synthesis loop, this purge may be recycled to the inlet of the cryogenic separation, thereby increasing the hydrogen recovery efficiency of the loop.
- 11) The process is steam balanced. The amount of steam produced and consumed in the process is much lower than in the existing conventional process, because of the much lower amount of heat liberated in the primary steam reformer. As a consequence, part of the heat contained in the effluent from the secondary reformer must be used to superheat the total steam produced in the process.

#### ADVANTAGES OF NEW LOW ENERGY PROCESS

All these features of the new low energy process concur to achieve the following advantages of the process, which are summarized on Table II.

- 1) The overall energy consumption of the new low energy process potentially can be as low as 25.1 MMBtu/ST (7.0 MMKCAL/MT). This includes the electric power consumption, which has been converted on the basis of 1 Kwh = 10.250 Btu/h. (2 588 Kcal/h).
- 2) The size of the steam reformer in the new low energy process is about four to five times smaller than in the conventional process, and the synthesis gas compressor would have one casing less than in the conventional process. On the other hand, the addition of cryogenic separation is a relatively small incremental cost.
- 3) As the operating temperature in the primary reformer (700 to 750°C) is much lower than in the conventional process, directionnally there are less problems in the maintenance of this heater, and a longer life may be expected for the reforming tubes.
- 4) The high operating pressure that can be used in the synthesis gas train makes it possible to have a single barrel synthesis gas compressor, and may be in the near future to delete the synthesis gas compressor entirely.
- 5) The ammonia synthesis catalyst can be expected to have a long life due to the high purity of the final synthesis gas.

The operation of the secondary reformer with a high residual methane content, in the range of 2 to 5 Mol %, combined with an extremely low overall steam to carbon ratio, in the range of 1.0 to 1.5, is the main reason for achieving very low energy consumption. Another reason is the very high excess air used in the secondary reforming, in which the exothermic reaction of partial combustion has a much higher efficiency than the conventional combustion in the burners of the primary reformer.

The cryogenic separation in essence consumes part of the energy introduced in the system by the air compressor, and it "kills two birds with one stone" by removing simultaneously the excess nitrogen and all the methane from the synthesis gas.

## REVAMPING AN EXISTING PLANT

It can be readily seen that this new low energy process is ideally suited for revamping existing ammonia plants, assuming the same pressure is kept for the synthesis gas train. This is so because the two major changes to be introduced, an extra air compressor and a cryogenic separation, can be added without changing the existing plant, and even without shutting it down for an appreciable length of time. In fact, to take advantage of the very low overall steam/carbon ratio used in the reforming, bearing in mind that the low level heat of the effluent gas from the shift conversion can be used advantageously to humidify the process air to the secondary reformer (and the process natural gas) and to supply heat to an absorption refrigeration system, it seems quite natural to consider seriously using a physical absorption system for removing the  $\text{CO}_2$  from the synthesis gas.

It will be also realized that in such a revamp, the synthesis loop would be fed with a very high purity dry synthesis gas, and there will be appreciable extra capacity in the synthesis loop; furthermore, the fresh synthesis gas can be introduced downstream rather than upstream of product ammonia separation leading to lower ammonia content in the converter feed. Therefore it seems quite natural to consider increasing the capacity of the ammonia plant while revamping for reduced energy consumption. This is further encouraged by the fact that the primary reformer will also have a tremendous overcapacity. By accepting a higher pressure drop, the secondary reformer as well as the HTS and LTS can accept a higher capacity without any change of equipment.

Since additional air compression capacity is required to reduce the energy consumption, the new air compression capacity can be tailored, at little cost, to match the new extra capacity. As to the synthesis gas compressor, it is also possible, at little cost, to increase its capacity by adding a booster at the suction, which also compensates for the higher pressure drop of the synthesis gas train.

In conjunction with a booster, it is also feasible to consider installing the cryogenic separation unit at a high pressure - say 60 atm - by withdrawing syngas at a suitable intermediate stage of the existing syngas compressor and routing it through the cryogenic unit and back into the next stage

of the compressor. This can gain the significant advantage of reducing the amount of excess nitrogen required hence reducing air compressor requirements.

In order to convert a conventional ammonia plant to the new low energy process, the modifications to be considered may be classified into two groups, each group representing a related package.

### First group :

- addition of a new air compressor
- addition of a cryogenic separation upstream of the synthesis loop
- revamping of the primary reformer
- if necessary addition of a booster at the suction of the synthesis gas compressor
- possibly humidifying the process air and the process natural gas, using part of the low level heat after the LTS.
- reducing appreciably the steam to carbon ratio by treating only part of the feedstock in the primary reformer.

### Second group :

- replacing the  $\text{CO}_2$  removal system by one requiring little or no heat for reboiling.
- replacing the compression refrigeration system by an absorption refrigeration system using the low level heat of the synthesis gas leaving the LTS reactor.

We will examine now briefly the main processing steps and see how they will be affected by the revamping operation.

As the existing primary reforming capacity is very large compared to the requirements of the new low energy process, the radiant section should be reduced to about half or one third its original size. The convection section should be modified to satisfy the preheat requirements of the process streams according to the new material and heat balances. Possibly, some supplementary firing may be added to the convection section to have some operating flexibility.

It is assumed that no modification will be required for the catalytic reactors such as the secondary reformer, the HTS and LTS converters, and the methanator.

The higher pressure drop in these reactors, due to the higher nitrogen content of the synthesis gas, and possibly due to the increased capacity of the plant, is compensated by a booster compressor and to a more