Ammonia Plant Capacity Increase by Autothermal Reforming and Dual Pressure Synthesis

This paper investigates the technical and economical feasibility of several concepts for a 30 % capacity increase of an old ammonia plant. It shows an interesting way to overcome the limitations in the two most critical plant units: Reforming capacity is increased by a newly added autothermal reformer, while capacity is added to the ammonia synthesis by the Uhde Dual Pressure Process. Using experience from reference projects, this process concept is compared to other technical options and is discussed on the basis of investment and operating cost. One of the factors making this concept competitive is the fact that by installation of parallel equipment with few tie-ins only, the shutdown time for its implementation is very short. Another interesting feature is that the ATR concept offers more CO_2 as a pure stream to be used in a urea plant compared to the other concepts. This provides the possibility to easily combine it with a larger urea plant.

Klaus Noelker
ThyssenKrupp Uhde GmbH, Dortmund, Germany

Introduction

A capacity increase of an ammonia plant can be a successful way to increase its economic viability. Also, a revamp involves considerably less risk than the erection of a new plant since the overall investment is moderate and project implementation takes less time.

A capacity enlargement up to about 10 to 15 % can usually be realized with moderate modifications by mobilizing the reserves which are already present in the majority of the process units. Only some equipment items are acting as bottlenecks and require modifications or replacement.

Larger capacity increases tend to require more substantial measures and bigger changes in the process. As this makes the capacity increase considerably more expensive, it is of key importance to select the most cost effective solution.

Scope of the Study

Basis for the investigation is an existing ammonia plant in Russia. Its actual capacity at the time of preparing the study was 1680 MTPD (1852 STPD). An expansion target of 30 % extra capacity was chosen. The plant had already undergone several capacity enlargements and modifications. Its current production capacity is considerably larger than its original nameplate capacity. Therefore, no significant potential for additional capacity is available in the existing plant equipment.

Three different revamp concepts are presented, with their main difference being within the reforming section. To compare their economic viability, capital and operating cost have been determined and evaluated for each revamp concept [1], [2].

In order to obtain the base data both for capital and operating data, material and energy balances have been prepared for each revamp concept using AspenPlus. These simulations include the process itself as well as its waste heat utilization including steam production and consumption.

Basics of Capacity Increase

Extending the capacity of a chemical process plant in general requires a larger throughput of the feedstock to be treated, i.e. larger flowrates in almost every part of the plant. The measures to be taken in a capacity revamp have to provide the following conditions for the larger flowrate:

Process Flow

If no measures are taken to increase cross sectional areas along the flowpath, the larger flowrates inevitably increase pressure losses. Higher pressure drop means higher loads on compressors and drivers. A 30 % increase in flowrates associated with the envisaged capacity expansion would result in an almost 70 % higher pressure drop. Hence, it is worthwhile to look at options to mitigate the effect of higher flowrates on pressure drop.

Higher natural gas inlet pressure could compensate the additional pressure drop. However, the potential for pressure increase is rather limited as the equipment design pressures are usually only a little above their original operating pressure. With respect to pressure drop, the arrangement of additional equipment in parallel to existing equipment is obviously preferable compared to arrangement in series. However, parallel arrangements might require more elaborate piping arrangements.

Heat Transfer

The larger heat duties which have to be transferred to or extracted from the process gas can be met via increasing the area of heat transfer surfaces, improved heat transfer coefficients or larger temperature differences. In certain areas a change in outlet temperatures caused by larger temperature differences can be tolerated, e.g. if the associated effects are only moderate drops in plant energy consumption.

Chemical Reactions

Every catalyst loses activity with time. Larger flowrates speed up this process. If no further measures are taken, reductions in service life of the catalyst charges are expected after the revamp implementation. On the other hand, a plant dating back from a time some decades ago has been designed for the catalysts available at that time. Since some catalysts have undergone significant improvements, the higher activity available today can compensate for the effect of higher throughput. In other cases, larger catalyst volume or smaller grain size might be the choice, both leading to higher pressure drop.

Separation of Undesired Species

In several process steps of an ammonia plant separation of undesired species from the process gas takes place. The most critical area is the CO_2 removal. However, there are several other unit operations such as the desulphurization as well as the removal of condensate from process gas and finally the separation of the produced ammonia from the recycle gas in the ammonia synthesis.

All these process steps have been designed for certain volume flowrates. Once their capacities are exceeded, they usually respond with fast drops in performance, i.e. with effects like sharp rises in concentration of the species to be removed or with entrainment of liquid into the process gas.

Compared Process Concepts

As shown in the block flow diagram of the existing ammonia plant in Figure 1, its process is fairly conventional.

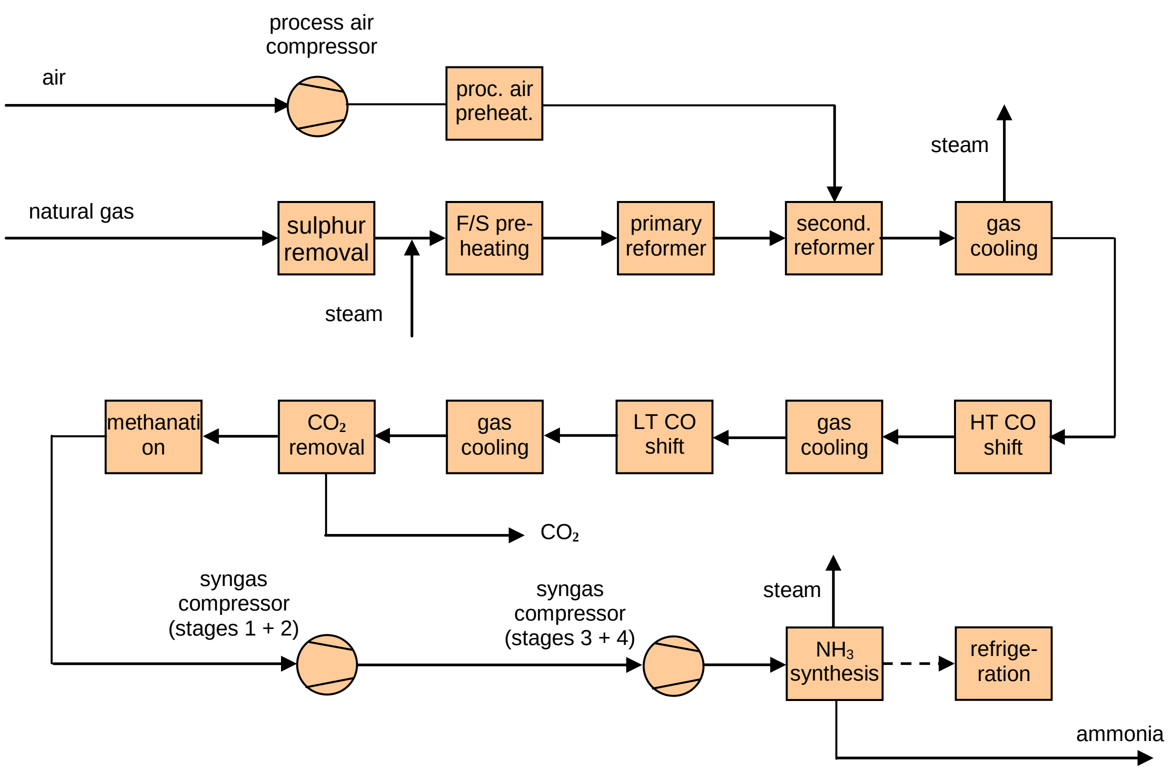

Figure 1: Block flow diagram of the existing ammonia plant.

Desulphurization

The desulphurization has to cope with the higher gas throughput. The 30 % higher flow rate will reduce the lifetime of the zinc oxide bed by 30 %. Whether this is acceptable or not depends on the actually achieved lifetime and the turnaround strategy of the plant. If more catalyst volume is required, the installation of a second parallel vessel which allows catalyst replacement while the plant is kept in operation is fairly simple.

Reforming

The most critical and cost intensive part of the synthesis gas generation is the reforming section. The following three process alternatives have been selected for this comparison:

- Concept 1: Enlargement of existing primary / secondary reforming section
- Concept 2: Secondary reformer operation with oxygen-enriched air
- Concept 3: Autothermal Reformer (ATR) parallel to existing reforming section

The main features of the concepts are discussed in detail below.

Concept 1: Enlargement of Existing Reforming Section

Figure 2 presents a block flow diagram of the plant's synthesis gas generation section for this revamp concept. For every process unit of the reference plant it contains the modifications required for the envisaged capacity enlargement. The dashed lines indicate the unit operations which have to be modified or need an upgrade by additional equipment.

In the existing plant preheating of the feed for desulphurization and of the primary reformer feed (feed / steam mixture) is done in the waste heat section of the primary reformer. Modifications of the respective heat exchanger coils are required to achieve the additional heat transfer.

Within the reforming section, the split of reforming duties between primary and secondary reformer remains unchanged. Hence, both reformers have to transfer approximately 30 % larger heat duties to the process gas. In principle, the following options are available to increase the overall heat duty of the steam methane reformer (SMR):

- increased average heat flux
- larger heat transfer area by:
  - larger reformer tube diameter
  - longer reformer tubes
  - additional reformer tubes

For the existing primary reformer of the reference plant a significant rise of the average heat flux has been ruled out. Also, new reformer tubes with a larger diameter are not an option since the reformer is already equipped with 5" tubes. Longer reformer tubes are in principle a viable path to increase the heat exchange surface of a new reformer. However, for an existing reformer this option is not practical, as it would entail a complete rework of the furnace box including a relocation of the entire inlet manifold arrangement.

For these reasons, the addition of a suitable number of tubes was considered the only feasible option. The additional reformer tubes can either be placed in a separate oven box or integrated into the existing primary reformer. A separate box would have advantages compared to integration of the new tubes into the existing box both from the process side (lower pressure losses, lower flow velocities on the flue gas side) as well as from the construction side (easier tie-in of the revamp part, less shutdown time of the plant for tie-in). However, a completely separate new furnace box plus the associated waste heat section would be an expensive solution and render this revamp concept uncompetitive.

Enlargement of the existing oven box and integration of the additional reformer tubes is a fairly complicated exercise. Nevertheless, it has been successfully carried out in the past.

To cope with the larger process gas flowrate and the associated larger heat duty, the secondary reformer has to be replaced. A new compressor must be installed parallel to the existing process air compressor to provide the additional amount of process air. An electric motor is selected as driver. The process air preheating coils in the waste heat section of the primary reformer have to be modified for the additional heat transfer. All heat exchangers in the process gas cooling train downstream of the secondary reformer have to be backed by additional parallel heat exchangers or replaced by single larger units.

Concept 2: Secondary Reformer Operation with Oxygen-enriched Air

Figure 3 is a block flow diagram of this process alternative. The basic approach of Concept 2 is to provide the additional heat duty for reforming the larger process gas flow via the secondary reformer (SR). Using ambient air to supply the required larger oxygen flow to the reactor would introduce a considerable amount of excess nitrogen into the process gas. This can be avoided via operation with oxygen-enriched air.

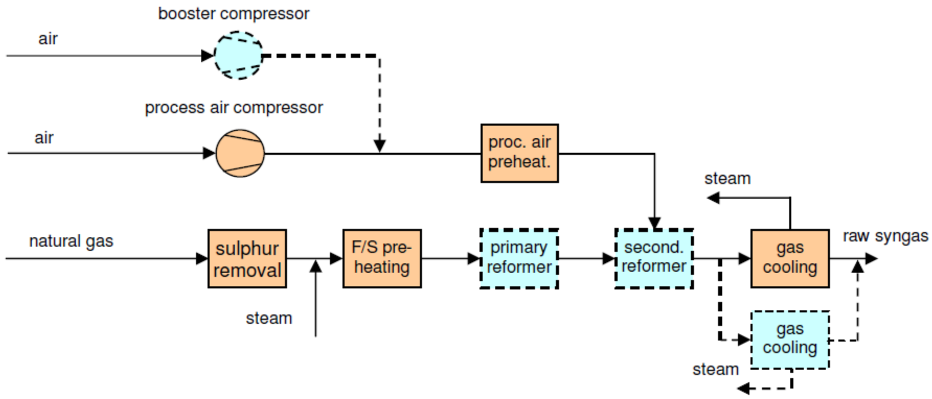

Figure 2: Block flow diagram of reforming section of Concept 1: Enlargement of existing reforming section. Modifications marked in dashed / blue.

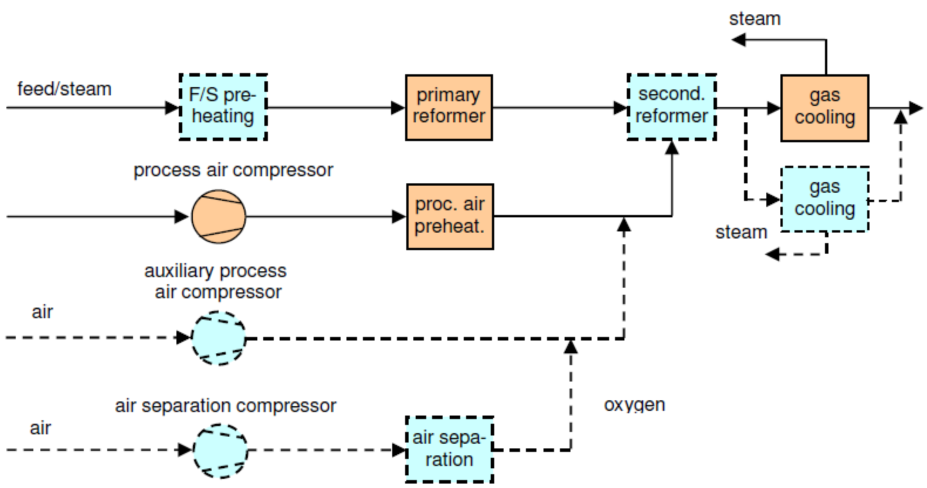

Figure 3: Block flow diagram of reforming section of Concept 2: Secondary reformer with enriched air. Modifications marked in dashed / blue.

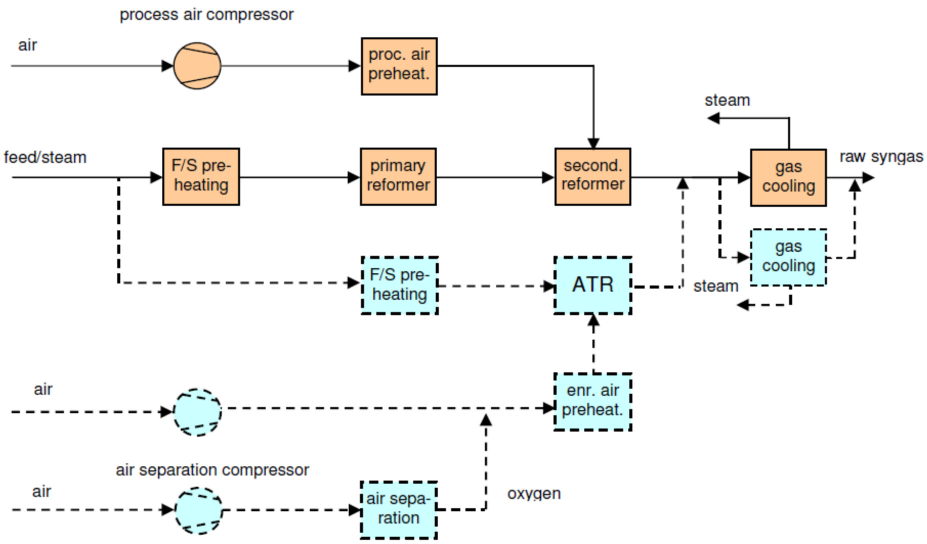

Figure 4: Block flow diagram of reforming section of revamp Concept 3: Parallel ATR. Modifications marked in dashed / blue.

The oxygen required for air enrichment must either be imported from outside or provided by an air separation plant within battery limits. In both cases, its operating and capital cost may not be neglected in the economic analysis. Generating the oxygen within battery limits is associated with significant additional capital investment. However, it has the advantage that the purity of the oxygen can be adjusted to the requirements of the process and in general does not have to be very high. Also, integration of the power requirements of the air separation with the plant's steam system is possible. Hence, this alternative has been assumed for the cost comparison.

The primary reformer remains mostly unchanged. Also, the other modifications required in the reforming section are fairly similar to Concept 1.

The process gas of Concept 2 contains more CO and CO_2 than the gas of Concept 1 due to the autothermal supply of the extra reforming duty. Hence, the loads for the CO shift converters and for the CO_2 removal unit are somewhat higher compared to Concept 1.

Concept 3: ATR parallel to Existing Reforming Section

The basic approach of this alternative is an autothermal reformer in parallel to the existing reforming section. Figure 4 contains the block flow diagram of this process concept. In the same way as in Concept 2 it is operated by oxygen-enriched air in order not to exceed a reasonable amount of nitrogen in the process gas.

The autothermal reformer is a brick-lined vessel in which the two inlet streams of feed / steam mixture and oxygen are brought to reaction. It consists of a first reaction zone (or combustion zone) at the top, at the inlet of the two streams, and of a second catalyst-filled reaction zone in the bottom. Its principle design is shown in Figure 5.

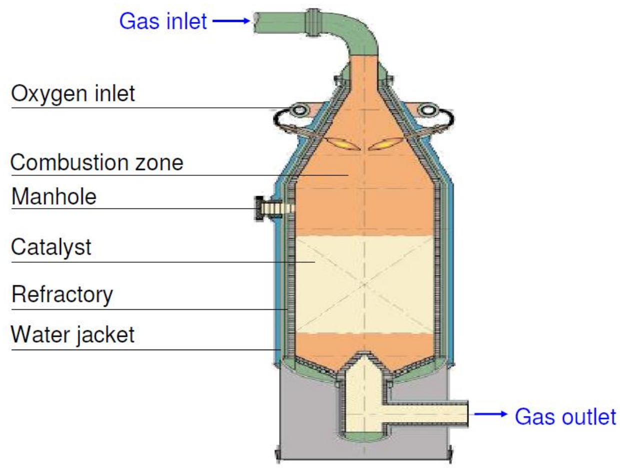

Figure 5: Principle sketch of the autothermal reformer.

The existing reforming section remains essentially unchanged. Since the additional reforming is done entirely through autothermal reforming, this concept requires more oxygen than Concept 2. Thus, the CO_2 content in the process gas is even higher than in Concept 2, having an impact on the duty of the CO_2 removal unit.

CO Shift and Methanation

The three different reforming concepts of course have influence on the conditions and capacity of the other process units, but in principle for all concepts, identical solutions for the other process units are selected for all three reforming options. The difference in equipment size has been taken into account for in the cost evaluation.

As the CO shift reactors are sufficiently sized in the reference plant, no change is made with them and a slightly higher CO content is tolerated. Also the methanation reactor does not require changes.

The final gas cooling upstream of the CO_2 removal requires an additional cooler.

CO_2 Removal

As there are many different processes available for CO_2 removal, there are also many different options for a capacity increase. Options for capacity increase include:

- Change of packing material in the absorber, allowing for higher gas and liquid loads
- Installation of an additional flash step in the desorption section of the solvent cycle for improving the solvent regeneration
- Change of the activator in the absorption solution to a more effective one
- Complete change of solvent type (e.g. from potassium carbonate to amine-based), involving also significant modifications at equipment

Typically, the absorber is the bottleneck. Changes in the desorption section are easier to implement as they involve low-pressure equipment. If needed, additional regeneration heat can be provided by low pressure stream.

Syngas Compression and Ammonia Synthesis

Upgrading all items in the synthesis loop would be a task which would involve many modifications at the existing high-pressure piping and equipment. This would cause a high amount of modifications which would have to be executed in a small area in a short time while the plant is not in operation.

Therefore, an approach is selected where a whole new unit can be installed while the plant is in operation, and only a few tie-in points have to be connected during a shutdown of the plant. This is done by selecting the Uhde Dual-Pressure Synthesis [3]. It consists of a once-through (OT) ammonia synthesis which is added to the plant on an intermediate pressure level between synthesis gas generation and the synthesis loop as shown in Figure 6.

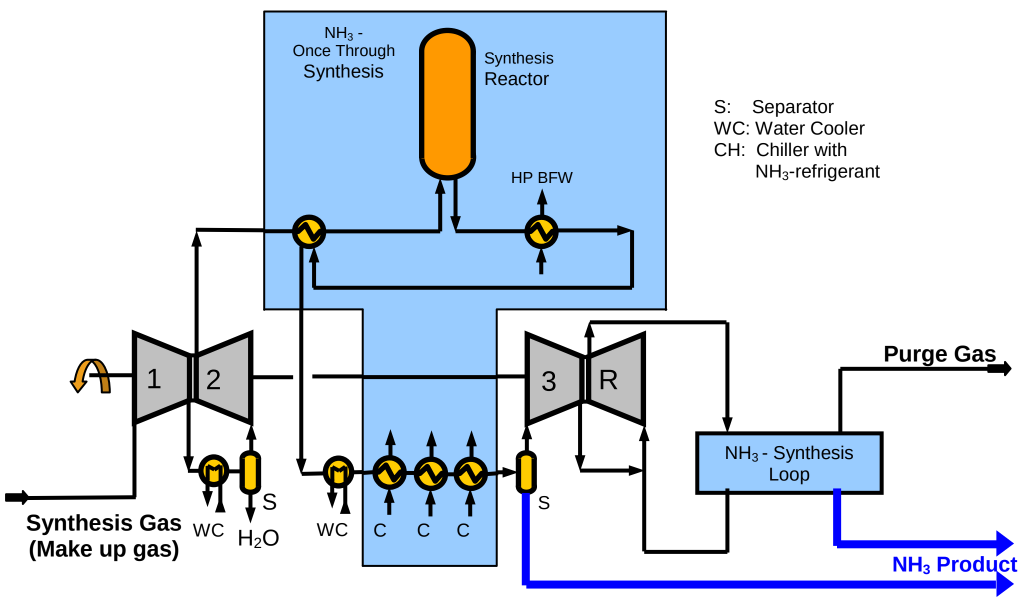

Figure 6: Principle sketch of the Uhde Dual Pressure Ammonia Process. Added once-through synthesis highlighted in blue.

For the envisaged relatively large capacity expansion, it can be assumed that the synthesis gas compressor is not able to cope with the significantly larger flowrate. Hence, an auxiliary compressor parallel to the first and second stage of the existing syngas compressor has been selected (not shown in Figure 6), essentially taking the additional gas up to the intermediate pressure level.

The additional syngas is then mixed with the gas coming from the second stage of the existing syngas compressor and the combined gas is then passed through the OT synthesis. This process unit comprises of a gas / gas heat exchanger to provide the elevated converter inlet temperature, the actual OT synthesis converter, a steam generator / boiler feed water preheater and a cooling train. The latter consists of a water cooler and a series of chillers, bringing the process gas temperature down to a level for separation of most of the generated ammonia by condensation.

Subsequently, the remaining process gas is passed on to the third stage of the synthesis gas compressor for further compression up to the pressure level of the ammonia synthesis loop. Essentially, the flowrate and composition of the gas fed to the synthesis loop is the same as in the original plant.

The Uhde Dual-Pressure Process has been successfully installed already for a revamp in a plant in Slovakia [4] and for two new plants [5], being the two largest single-train ammonia plants in the world.

One item particular to this plant is that the purge gas from the loop is not sent to a hydrogen recovery unit because it is already used for other purposes in the existing complex. This feature is maintained also for the revamp calculations. Since it is done like that for all investigated concepts, it does not affect the results of the study. Technically, there would be no difficulty to add a unit separating the hydrogen from the purge gas and returning it to the synthesis loop.

Steam System

For each process concept, the steam system is adjusted to match steam production from waste heat with steam consumption of process and turbines. Same as in the existing plant, also after the revamp, some MP steam has to be imported from outside battery limits.

Whole plant

Table 1 gives an overview of which main equipment items have to be replaced or significantly modified for the three concepts. This list forms the basis for the capital cost assessment.

Results of the Economical Comparison

Operating Cost (OPEX)

The following streams entering or leaving the plant (see also Figure 1) are associated with dedicated energy contents and respective cost data:

- Import:
  - Feed gas
  - Fuel gas
  - MP steam
  - Electric energy
- Export
  - Purge gas stream from ammonia synthesis

Table 2 contains a representation of these streams for each revamp concept in terms of their energy content:

- Feed, fuel and purge gas export are represented by their lower heating value (LHV).

- MP steam is represented by the fuel energy required for its production in a boiler.
- The imported electrical energy is shown as the fuel energy required for generating it via a steam cycle with an overall efficiency of 30 % (1 kWh electrical energy corresponds to 12000 kJ natural gas).

The overall consumption figures are shown in the bottom line of the table. Of course, the air separation unit with its motor-driven compressor is always included in the figures.

Equipment item	**Process Concept**		
	1	**2**	**3**
	Enlarged SMR	**SR with enriched air**	**parallel ATR**
New parallel process air compressor	x	x	x
Air separation unit		x	x
Steam reformer furnace box expansion	x		
Replacement of reformer combustion air fan	x	x	
Replacement of reformer flue gas fan	x		
Replacement / modification of secondary reformer	x	o	
New autothermal reformer			x
New fired heater for process gas preheating			x
Process air preheating coil	o	o	
Combustion air preheating coil	o	o	
Feed / steam preheating coil	o	o	
Natural gas preheating coil	o	o	o
Waste heat boiler with steam drum	x	x	x
CO_2 absorber	x	x	x
New once-through ammonia synthesis	x	x	x
New parallel synthesis gas compressor	x	x	x

Table 1: Main equipment items which need to be replaced or modified – basis for the investment cost determination. x: new equipment; o: modified equipment.

Item	**Unit**	**Process Concept**		
		1	**2**	**3**
		Enlarged SMR	**SR with enriched air**	**parallel ATR**
Feed gas	GJ/t NH_3	22.83	24.28	24.15
Fuel gas	GJ/t NH_3	13.28	11.94	11.78
MP steam import	GJ/t NH_3	2.07	1.72	1.58
El. power	GJ/t NH_3	0.98	1.16	1.15
Purge gas export	GJ/t NH_3	-1.73	-1.79	-1.74
Overall consumption figure	GJ/t NH_3	37.43	37.31	36.91

Table 2: Specific energy consumption of the revamp concepts, LHV Basis (37.43 GJ/t = 32.17 MMBTU/ST)

The consumption figures are relatively high compared to the values achieved by newly built ammonia plants. It is worth to mention that the study deals with an old plant and that the focus of the study is on capacity increase, not energy optimization. Certainly it would be possible to additionally improve the energy efficiency of the plant, but that would lead to additional cost. Since the changes would be about the same for all three concepts, they would not add value to the target of comparing the revamp concepts and thus are left out of consideration.

As the revamp parts are fully integrated into the existing plant, it is not possible to give individual figures for the energy consumption of the revamp parts alone.

Capital Cost (CAPEX)

The capital cost for the revamp is estimated by first estimating the cost of the main equipment and then reflecting other costs for the implementation by the factor method.

That means, first the equipment cost is determined for the all the three process concepts using the process data determined in the simulations. Then the cost of all other contributions (piping, instrumentation, electrical, civil, engineering, procurement, erection and commissioning) is added, assuming that these can always be expressed by multiplication of the "pure" equipment cost by a certain factor. These factors are known from experience.

Concepts 2 and 3 require oxygen-enriched air for operation. It shall be noted that for the sake of a fair comparison the oxygen stream is not treated as a readily available utility but that the investment and operating cost of the air separation unit is included in the data.

Although the cost determined by the factor method includes cost for construction, it does not include the cost of lost production due to downtime for the revamp implementation. This is an important part of the real cost of a capacity expansion.

Obviously, erection time for the revamp extends over several months. During this time, the existing plant can maintain in operation. A complete shutdown of production is needed only to carry out the final tie-ins and for commissioning of the new sections.

However, an analysis of the revamp concepts reveals significant differences between them with respect to the activities for their final implementation. As the concepts vary mainly in the reforming section, the main differences are related to this equipment:

- Concept 3 requires only tie-ins at relatively cold and therefore non-critical piping.
- Concept 2 requires the installation and tie-in of a new secondary reformer. Hence, it is assumed that it demands at least one additional week for this work.
- Concept 1 requires difficult structural work to enlarge the box of the existing steam reformer. Even with a considerable amount of preassembling, it seems likely that this work would prolong the scheduled shutdown by four weeks compared to Concept 3.

In the ammonia synthesis, the installation of the once-through section is proposed. This offers the same advantage of short time requirement for the tie-ins which Concept 3 offers for the reforming as described above

The additional shutdown periods for Concepts 1 and 2 have been turned into capital costs via the assumption of an ammonia sales price of 400 USD/t and energy cost of 4.0 USD/MMBTU. This leads to the individual implementation cost for each revamp concept listed in Table 3.

Item	Unit	Process Concept		
		1	2	3
		Enlarged SMR	SR with enriched air	parallel ATR
Synthesis gas generation	million USD	80.4	98.4	93.8
OT synthesis	million USD	71.2	71.6	69.1
steam system	million USD	2.1	2.2	2.1
reformer waste heat section	million USD	3.9	2.9	3.1
Total erection cost	million USD	157.5	175.1	168.0
Additional shutdown time	weeks	4	1	0
Lost profit by shutdown time	million USD	15.7	3.9	0.0
Overall capital cost	million USD	173.3	179.1	168.0

Table 3: Capital cost of the revamp concepts: Total erection cost is determined by equipment cost and factors for other cost contributions. Overall capital cost (bottom line) is total erection cost plus lost profit by additional shutdown time. Assumptions for calculation of lost profit: refer to text.

Table 3 shows that Concept 1 is the one with the lowest erection cost. However, due to its complicated nature, the lengthy shutdown period adds significant cost by loss of production to it.

In overall cost, Concept 3 is the most attractive one. The difference between the overall costs of all concepts is 7 %. Certainly, there is some degree of inaccuracy in the cost data, but as the same methods of estimation were used for all concepts, it is believed that the data represent the correct ranking between the concepts.

CAPEX / OPEX Comparison

Finally, an economic comparison of all revamp concepts is made, considering their CAPEX and OPEX.

The comparison is made by determination of the specific production cost, which means the cost of production per ton of ammonia. (Alternatively, also the net present value of all concepts could be determined.)

This requires converting the investment cost into an annuity by an economic model, consisting of interest rate and required payback period. As it is always the case, the result of the comparison can strongly depend on the economic model used.

To illustrate the influence of the model, different scenarios have been evaluated. The first one (low interest rate and long payback period) in principle favors capital intensive plants with low specific energy consumption The second one (high interest rate and short payback period) just favors the opposite, i.e. plants with comparatively low investment and higher energy consumption. The scenarios are combined with two different specific energy costs. The scenario parameters are listed in Table 4.

For operating cost, it is assumed that they are sufficiently covered by the contributions for the streams shown in Table 2. The assumption is justified that the other components related to personnel or maintenance are equal for all concepts. As the aim is only to determine the ranking between the concepts, they can be neglected. The study is made for two different gas cost as shown in Table 4. The cost for steam and electricity is derived from the gas price.

The resulting specific production cost (CAPEX plus OPEX) are summarized for different combinations of parameters in Table 5. It shall be mentioned again that these figures shall serve only to determine the ranking between the concepts. As the energy consumption of the plant used as the basis is not optimized, some figures appear high.

Parameter	Unit	Values
Interest rate	% per year	4 or 10
Required payback period	years	5 or 15
Energy cost	USD/MMBTU (LHV)	1.0 or 4.0
Operating days	days / year	350

Table 4: Parameters for economic scenarios.

Economic scenario			**Production Cost for Process Concept**		
Energy cost	**Annual interest rate**	**Payback period**	**1 Enlarged SMR**	**2 SR with enriched air**	**3 parallel ATR**
USD/MMBTU	%	years	USD/t	USD/t	USD/t
1.00	4	15	128	129	122
1.00	10	5	307	309	289
4.00	4	15	231	234	226
4.00	10	5	404	412	394

Table 5: Comparison of the specific production costs (CAPEX plus OPEX) for the three different revamp concepts under four different economic scenarios.

The ATR-based revamp Concept 3 with the lowest specific energy consumption (see Table 2) is also the one with the lowest investment cost. Hence, it is not surprising that it also shows the lowest overall production costs.

CO_2 Production

For all concepts, CO_2 is emitted by the following three sources:

- flue gas from reformer stack Inside Battery Limits (ISBL)
- flue gas from steam generation boiler stack Outside Battery Limits(OSBL)
- CO_2 stream from CO_2 removal unit.

Practically all carbon in the natural gas is finally ending up as CO_2 in one of the above streams.

Also as to the electricity consumption, a (virtual) CO_2 emission can be assigned because electricity production (inside or outside the plant) is linked to CO_2 emission. As the electricity consumption of all three process concepts is similar and fairly small compared to the natural gas consumption, for sake of simplicity, no CO_2 emission equivalent is assigned to electricity consumption.

Table 6 shows the thus determined CO_2 emission per ton of ammonia.

While the first two of the above listed streams from the stacks contain CO_2 in a concentration of only approx. 10 % at ambient pressure, the third one is more than 99.5 % pure CO_2 at slightly elevated pressure. Only this CO_2 can readily be used for production of urea fertilizer.

Stream	Unit	Plant concept		
		1	2	3
Total CO_2 generation	t CO_2 / t NH_3	2.10	2.09	2.06
CO_2 available for urea production	t CO_2 / t NH_3	1.16	1.23	1.22
Max. urea production	t/d	3449	3667	3650

Table 6: CO_2 emission per ton of ammonia and maximum possible urea production at 2180 t/d ammonia.

Considering the stoichiometry of urea formation from ammonia and CO_2 and the available amounts of the reactants, Concepts 2 and 3 show an increase of about 6 % in maximum possible urea formation compared to Concept 1 (see Table 6).

If one would like to achieve the same urea production by the other two concepts, their amount of usable CO_2 would have to be raised. Basically two options are available to do so:

- CO_2 can be washed out of the flue gas of reformer or boiler by absorption and desorption. The result is an almost pure CO_2 stream which could be added to the existing CO_2 stream. This solution adds high investment cost and a little operating cost to the process.
- More gas can be fed through the plant up to and including the CO_2 removal unit. This would produce more CO_2 in the process which is consequently separated in the CO_2 removal unit, thus increasing the CO_2 export stream. The surplus synthesis gas downstream of the CO_2 removal is fed to the reformer burners. This solution makes the reforming and CO_2 removal sections of the plant a little larger, increasing investment a little, but adds significant operating cost to them.

Both concepts have been applied already to revamps and new plants.

Summary

A detailed investigation has been carried out to assess the economic feasibility of three different concepts for a capacity increase by 30 % of an existing ammonia plant. The main difference between the three concepts is in the reforming section.

The aim of the investigation is to establish an economic ranking between the revamp concepts. Concept 3 which is based on an autothermal reformer (ATR) turns out to be the most attractive solution. The other concepts (enlargement of the existing steam reformer resp. operation of the secondary reformer with oxygen-enriched air) have higher overall costs. Responsible for this ranking are mainly the advantages of the ATR-based concept in overall capital costs compared with the other concepts. The process calculations show only moderate differences between the individual energy consumption figures.

The overall capital costs must include all costs associated with revamp implementation, including loss in production by longer plant shutdown. Especially Concept 1 (enlargement of the existing steam reformer) requires considerably more complicated implementation work which contributes to its higher overall cost.

References

[1] M. Papsch: Technische und wirtschaftliche Bewertung von Prozesskonzepten zur

Synthesegaserzeugung in Ammoniak-Anlagen, Bachelor Thesis, University of Applied Sciences, Krefeld (2011)
[2] J. Johanning, K. Noelker: Comparison of synthesis gas generation concepts for capacity enlargements of ammonia plants, Nitrogen + Syngas 2006 International Conference, Athens (2012)
[3] J. S. Larsen, D. Lippmann: The Uhde Dual Pressure Process – Reliability Issues and Scale Up Considerations, 47th Annual Safety in Ammonia Plants and Related Facilities Symposium,
San Diego, California (2002)
[4] F. Kessler et al., First application of Uhde's dual pressure ammonia process for revamping of the Duslo ammonia plant, Nitrogen + Syngas International Conference 2006, Vienna (2006)
[5] K. Noelker: Commissioning Experience of the World's Largest Ammonia Plant, ACHEMA International Conference 2009, Frankfurt a.M. (2009)