Uhde

Comparison of synthesis gas generation concepts for capacity enlargement of ammonia plants

-

by

-

Dr. Klaus Noelker and Dr. Joachim Johanning

Uhde GmbH

Dortmund, Germany

Prepared for Presentation at

NITROGEN & SYNGAS 2012 International Conference

Athens

20 – 23 February 2012

ABSTRACT

Capacity enlargements after some time in operation offer plant operators the chance to increase turnover and revenue in moderate steps. In many cases, capacity enlargements of ammonia plants are economical and therefore attractive to plant operators. Also, they involve considerably smaller risks than the erection of a new plant since the overall investments are moderate, project implementation takes less time and the amounts of extra product for which customers have to be found is limited.

Capacity enlargements up to about 15% can usually be realized with moderate modifications to a plant by mobilizing the reserves in the majority of the process units. Only a few equipment items constituting bottlenecks require more significant measures.

Larger capacity increases beyond this tend to require more substantial measures, especially if the plant had been subjected to a revamp before. As this makes the capacity increase considerably more expensive in relative terms, i.e. related to the additional amount of product, it is of prime importance to select the most cost effective solution.

The paper investigates the economics of various syngas generation concepts for a 30% capacity increase. Among the compared alternatives are the enlargement of the existing steam reformer and an autothermal reformer (ATR) in parallel to the existing syngas generation. In order to generate comparable consumption figures, the impacts of the different revamp concepts on the plant's energy balance have been included in the investigation

Contents:

INTRODUCTION

Newly errected production plants are in general specificly designed for a certain nameplate capacity. Normal operation at or close to design capacity usually dominates plant life with respect to time and therefore overall operating costs. Hence, it is desirable that all equipment items in this case operate at least close to their optimum.

For certain pieces of equipment this is not always possible due to competing demands related to operational flexibility. Also, individual capacity reserves for certain main equipment as required by the customer as well as engineering regulations and standards have to be accounted for. Still, plant designers will always seek the best compromises within these boundary conditions and tune the design for the nameplate capacity.

In most cases all this leads to moderate additional capacities of the plants beyond nameplate capacity which can be exploited without additional measures. Nevertheless, once a plant has been in operation for some time operating companies often become interested in additional product as they can find a market for it. Upgrading of an existing plant allows to increase capacity in moderate steps compared to a new plant and mitigates the inherent risks concerning erection time, costs and market conditions.

Since the original extra capacity at this stage usually has been realized already and the plant is operated at full load additional measures have to be taken to further increase capacity. All process steps and each individual piece of equipment need to be looked at to identify and eventually remove the bottlenecks preventing the desired increase in plant output. The offsites and utilities section of the plant needs to be checked also.

Usually, capacity expansions up to about 15% above original design capacity can be realized with moderate modifications. Only a limited number of main equipment items have to be either modified, replaced or backed up by additional equipment. Within the rest of the main equipment the remaining spare capacity is activated. In many cases such capacity enlargement show reasonable economic viability in respective studies and are subsequently implemented.

Larger capacity expansions beyond this – which then target production levels of 125 - 130% related to nameplate capacity – commonly require much more substantial measures. This makes such capacity extensions considerably more expensive both in absolute as well as relative terms. In cases where a plant upgrading of this kind is envisaged it is in general advisable to target a more substantial capacity increase. The cost for the additional equipment as well as the associated construction costs are by no means directly proportional to the capacity increase. Hence, capacity upgrades of this kind tend to become ever more economical with larger add-on capacity.

Also, it is of prime importance to select the most cost effective technical concept. This paper reports the results of an extensive investigation regarding the capacity expansion of an existing ammonia plant. The investigation is limited to the process plant itself with the main focus on the synthesis gas generation section. The offsites and utilities section has not been analyzed in detail.

SCOPE OF THE INVESTIGATION

As addressed above a substantial capacity expansion of a process plant requires the detailed investigation of all process steps in order to identify and eventually remove the bottlenecks.

An existing ammonia plant was chosen as the basis for the investigation. The capacity of the reference plant at the time of this study was 1680 mtpd. 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. Hence, the assumption appears justified that no significant potential for additional capacity is available in the existing plant equipment.

To be able to determine individual consumption figures for each of the investigated process alternatives – which allow a comparison between them – the entire process plant as well as the steam system of the reference plant have been included in the calculations. The main focus of this investigation is on the gas generation section. Therefore, only one process concept was used to extend the capacity of the ammonia synthesis.

BASICS OF CAPACITY ENLARGEMENTS

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

- Passage through the process gas flowpath

  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 volume flowrates associated with the envisaged capacity expansion would result in an almost 70 % higher pressure drop across the synthesis gas generation section.

  Since the syngas compressor has to take a larger process gas flow from a reduced front end pressure to synthesis pressure, it is inevitably the critical item in this respect. Hence, it is worthwhile to look at options to mitigate the effect of higher flowrates on pressure losses.

  In principal, a higher natural gas inlet pressure could compensate the additional pressure drop in the gas generation section. However, in most cases the potential for this will be rather limited as the equipment design pressures are usually directly related to their original operating pressure levels.

  With respect to pressure drop the arrangement of additional equipment parallel to existing equipment is obviously preferable compared to sequential arrangement.

However, parallel arrangements require more elaborate piping and additional flow control devices to assure the desired flow distribution.

Also, in some cases changes of reactor concepts from axial to radial flow of the process gas through the catalyst beds may still be an option.

- Transfer of the larger amounts of heat

  The larger heat duties which have to be transfered to or extracted from the process gas can be met via expansions of the 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 simply moderate drops in plant energy consumption.

- Analyses of the various reaction steps

  Every catalyst deteriorates and looses activity with time. Larger flowrates speed up this ageing process. Hence, if no further measures are taken, reductions in service life of the catalyst charges are to be expected after revamp implementation.

  This effect can be compensated for via a more active catalyst or larger catalyst volumes. Obviously, what can be done and what is economical to do depends on the individual state of each reactor. In many cases – as in the one reported in this paper – it may be appropriate to leave the CO conversion and the methanation reactors as they are and accept the reduction catalyst service life.

  Since the increased flowrates inflict higher fluid forces to the catalyst particles in the beds it should be checked for each reactor that this effect does not lead to significant movement of the top layer of the catalyst bed or even fluidization and the associated grinding effect, generating dust and mechanical deterioration.

- Separation of species down to required levels

  In several process steps of an ammonia plant separation of undesired species from the process gas takes place. The most critical area is the CO2 removal. However, there are several other unit operations such as the desulphurization as well as the removal of water in the process air compression, downstream of the CO conversion or during synthesis gas compression and last but not least 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 sharp rises in concentraion of the species to be removed. Also, some undesired effects may be observed such as entrainment of liquid into the process gas.

COMPARED PROCESS CONCEPTS

An existing ammonia plant was selected as reference plant for this investigation to base the process calculations on real plant parameters and to generate realistic figures for specific energy consumption. As can be derived from the block flow diagram in Fig. 1 the plant's process concept is fairly conventional.

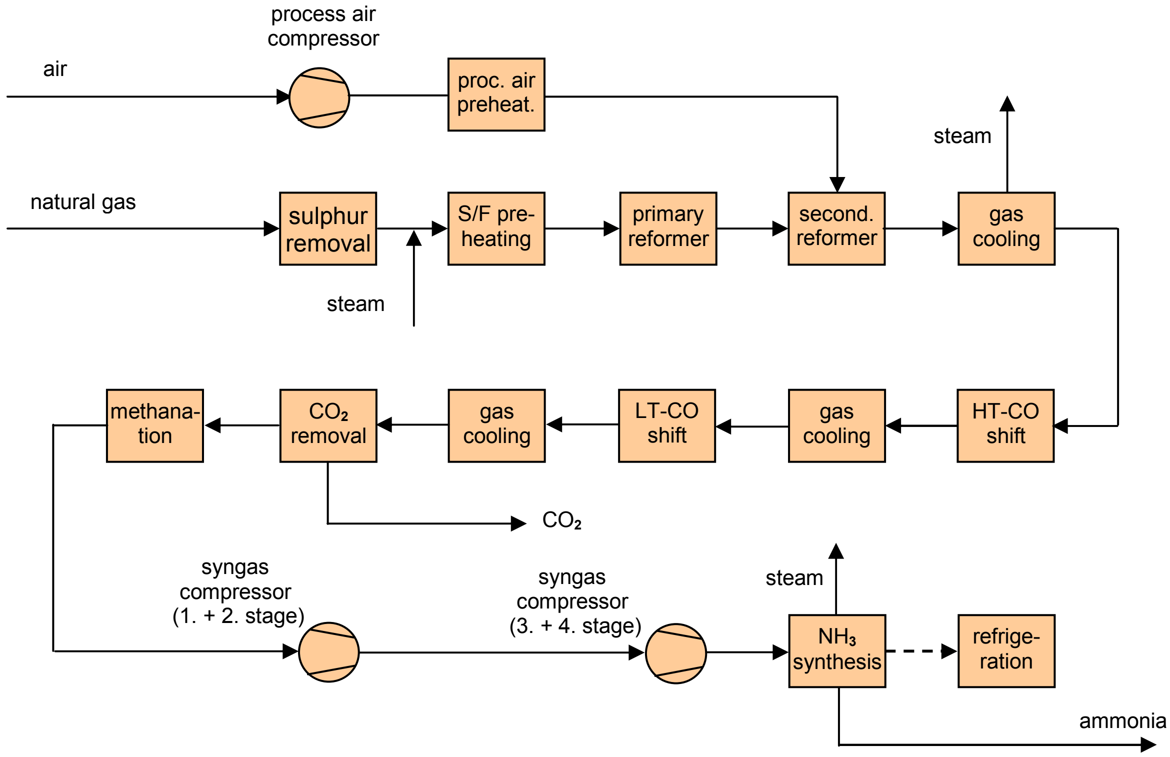

Fig. 1: Block flow diagram of the reference plant

One of the main targets of this investigation is to determine individual consumption figures for each process alternative included in this investigation. In an ammonia plant numerous interconnections exist between the various process steps via the waste heat utilization and the steam system. This demands that all plant parts have to be represented in the calculations.

Since the main focus of this work is to look explicitly on concepts for the syngas generation section, only one process concept has been employed to rise the capacity of the ammonia synthesis. The basis of this approach is the Uhde-Dual-Pressure-Concept. Fig. 2 contains a block flow diagram of the upgraded plant's back-end. It involves an additional once-through (OT) synthesis converter on an intermediate pressure level between synthesis generation and ammonia synthesis.

For the envisaged relatively large capacity expansion it can be assumed that the synthesis compressor will not be 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, essentially taking the additonal 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 final cooling train. This sequence of coolers takes the process gas temperature down to a level sufficiently low enough to separate most of the generated ammonia via condensation from the synthesis gas.

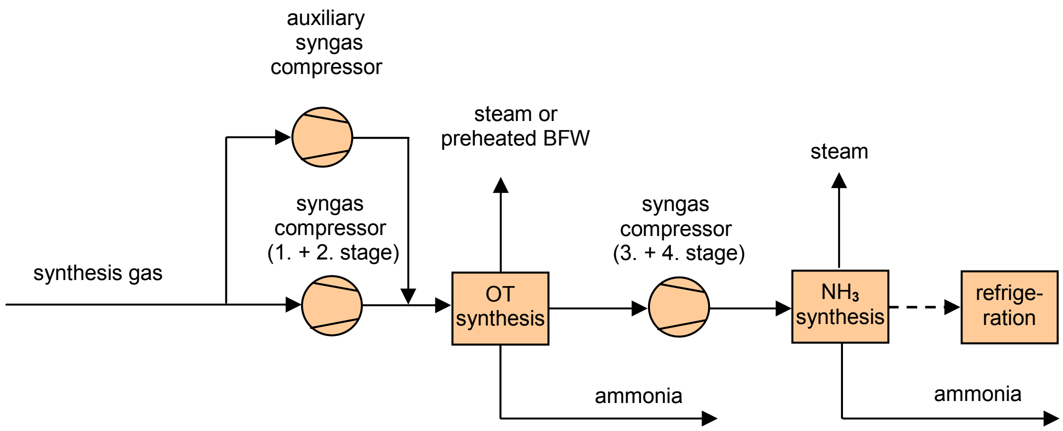

Fig. 2: Capacity expansion scheme for the ammonia synthesis

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 of the gas transfered to the ammonia synthesis is the same as in the original plant. Also, the composition is basically the same. Exceptions are moderate variations of the inerts depending on the revamp concept and an additional minor ammonia content.

To provide the additional synthesis gas for the extra ammonia produced in the OT synthesis a likewise substantial expansion of the plant's synthesis gas generation capacity is required.

The most critical and cost intensive part of the synthesis gas generation is the reforming section. Apart from a mere enlargement of the existing primary and secondary reformer several well publicized alternative solutions are in principal applicable. With respect to time and budget, the following three process alternatives have been selected for this comparison:

I. Enlargement of existing primary / secondary reforming section
II. Secondary reformer operation with oxygen-enriched air
III. Autothermal Reformer (ATR) parallel to existing reforming section

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

Concept I: Enlargement of existing primary / secondary reforming section

Fig. 3 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 as well as the additional equipment.

Without modification the desulphurization would respond to the larger natural gas flow with a reduced service life of the adsorber beds and increased pressure loss. 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 the two reactors remains essentially unchanged. Hence, both reformers have to transfer approximately 30% larger heat duties to the process gas. In principal, the following options are available to lift the overall heat duty of an externally fired steam reformer:

- increased average heat flux
- larger heat transfer area
  - 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 principal a viable measure to increase the heat exchange area of a new reformer. However, for an existing reformer this option is not practical as it would entail a complete rework of the oven box including a relocation of the entire inlet manifold arrangement. Hence, the addition of a suitable number of tubes was considered the only feasible option.

The additional reformer tubes could either be placed in a separate oven box or integrated into the existing primary reformer. A separate oven 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 oven 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 certainly is a fairly complicated exercise. Nevertheless, it can be done if the reformer offers certain design features and it has been carried out.

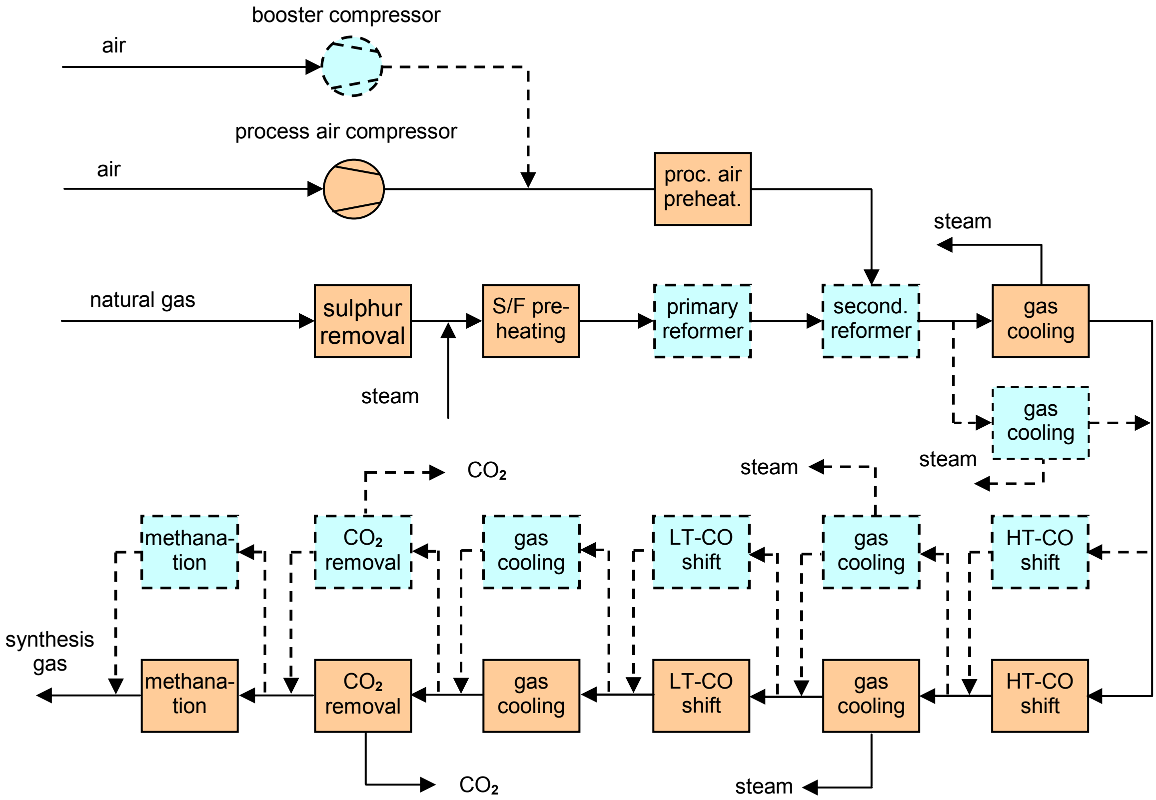

Fig. 3: Block flow diagram of revamp concept I

To cope with the larger process gas flowrate and the associated larger heat duty, the secondary reformer has to be replaced. An auxiliary compressor must be installed parallel to the existing process air compressor to provide the additional amount of process air. 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 reforming section have to be backed by additional parallel heat exchangers or replaced by single larger units. CO conversion and methanation reactors have been checked with respect to catalyst volumes. They can cope with the increased process gas flow without modification, albeit with significant reductions in catalyst lifetimes. Hence, it would essentially be a budget driven decision whether the existing converters are kept, backed by booster reactors or replaced by single larger vessels.

The CO_2 removal section requires an additional absorber to cope with the extra process gas.

Concept II: Secondary reformer operation with oxygen-enriched air

Fig. 4 again contains a block flow diagram of this process alternative. Only those process units are shown in the figure which are different compared to concept I. The basic approach of concept II is to provide the additonal heat duty for the larger process gas flow via the secondary reformer. 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.

The oxygen required for air enrichment must be imported from outside or provided by an air separation plant within battery limits. Generating the oxygen within battery limits is obviously 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 Capex/Opex comparison.

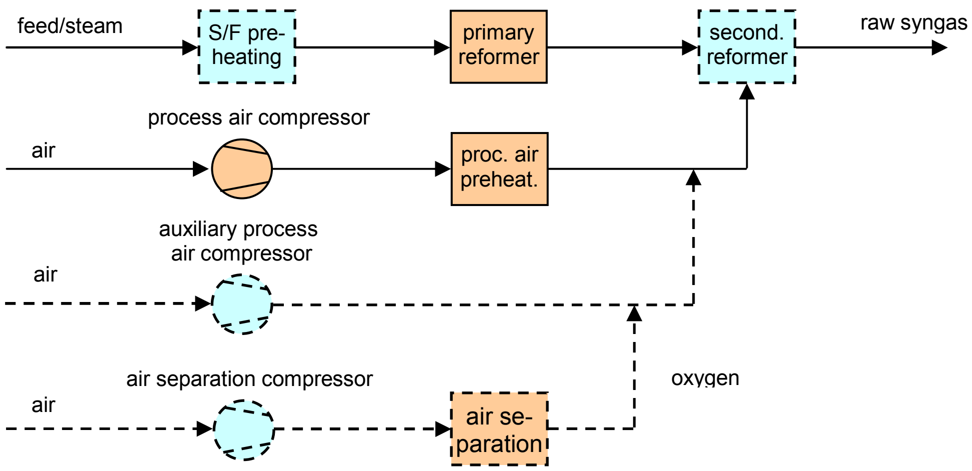

Fig. 4: Block flow diagram of revamp concept II

The existing primary reformer of the reference plant remains essentially unchanged. Also, the modifications required for the other process steps in the synthesis gas generation section are fairly similar to concept I.

The process gas of concept II contains more CO and CO_2 than the gas of concept I due to the autothermal supply of the extra reformer duty. Hence, the loads for the CO shift converters and for the CO_2 removal unit are somewhat higher compared to concept I.

Concept III: ATR parallel to existing reforming section

The basic approach of this alternative is a stand alone ATR parallel to the existing reforming section. Fig. 5 contains the block flow diagram of this process concept. The existing reforming section remains essentially unchanged. Since the additional reforming is done entirely through autothermal reforming, this concept requires more oxygen than concept II.

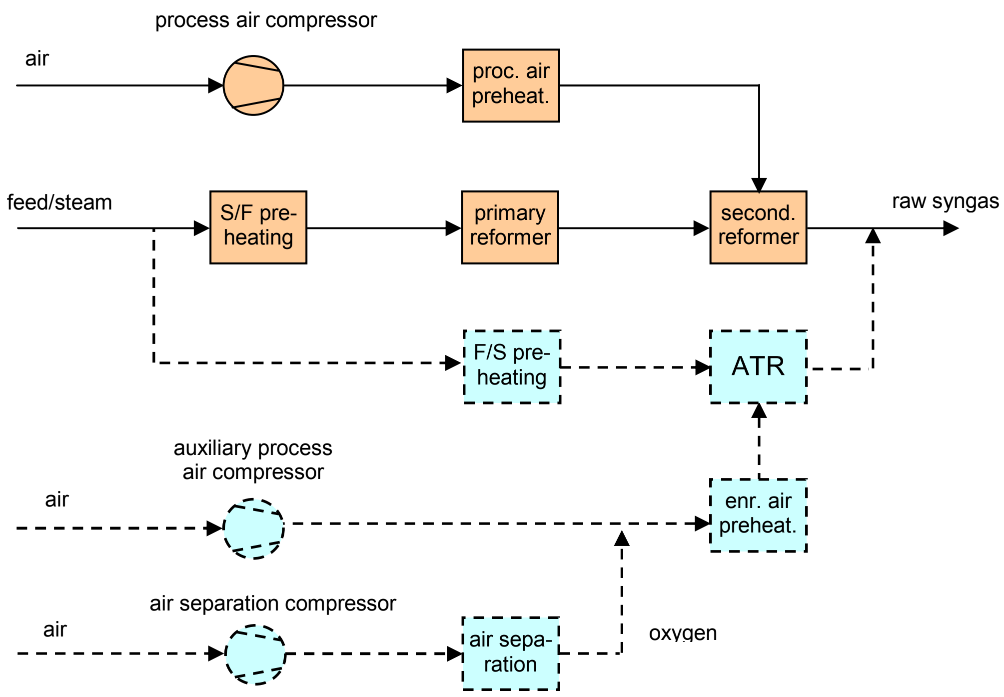

Fig. 5: Block flow diagram of revamp concept III

RESULTS OF THE ECONOMICAL COMPARISON

To assess the individual economic viability of each revamp concept, a complete Capex / Opex comparison has been carried out. Both operating costs and capital investment have been calculated individually for each revamp concept.

Operating cost comparison

AspenPlus-based material and heat balances have been prepared for each revamp concept to determine the individual operating costs. These balances include the entire waste heat utilization including the steam generators and all major steam consumers. Hence, the energy consumption figures generated through this investigation are very accurate.

Based on the material and energy balances individual energy consumption figures for each revamp concept have been calculated. The utility streams listed below pass battery limits and are associated with dedicated energy contents entering or leaving the plant.

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

Table 1 contains a representation of these utility streams for each revamp concept in terms of their energy content. It outlines the contribution of each stream to the overall specific energy consumption of the revamp concepts. The overall consumption figures are shown in the bottom line of the table. Feed and fuel gas are represented via their LHV and the MP steam via its specific enthalpy. The imported electrical energy has been transformed into the thermal energy required to generate it via an overall efficiency of the steam system of 30%. Hence, the thermal energy equivalent of 1 kWh electrical energy is 0,012 GJ.

Table 1: Individual specific energy consumption of the revamp concepts

		revamp concept		
		I	II	III
utility	**unit**	**enlarged SMR**	**SR with enr. air**	**ATR**
feed gas	Gcal / t_{NH3}	5,45	5,80	5,77
fuel gas	Gcal/ t_{NH3}	3,17	2,85	2,81
imported MP steam	Gcal / t_{NH3}	0,50	0,41	0,38
electrical power	Gcal / t_{NH3}	0,23	0,28	0,27
purge gas	Gcal / t_{NH3}	-0,41	-0,43	-0,42
overall spec. cons.	**Gcal / t_{NH3}**	**8,94**	**8,91**	**8,81**

Table 1 reveals that the differences in specific energy consumption between the revamp concepts are in the order of 1,5 %.

The absolute values of the consumption figures are relatively high compared to the values achieved by newly built ammonia plants which are in the range of 6,9 - 7,3 Gcal/t_{NH3}. It has to be mentioned, that the values listed in the table have to be considered as relative figures which give the correct ranking between the revamp concepts with respect to energy consumption. The revamp parts are fully integrated with the existing plant with respect to energy utilization to achieve best possible overall consumption figures. This makes it impossible to separate the revamp parts from the existing plant with respect to energy consumption and determine individual figures for the revamp parts alone.

Capital Investment Comparison

Based on the process parameters determined in the material and energy balancing of the revamp concepts a first assessment of the capital investment associated with the individual concepts has been carried out. The methodology applied shall be explained briefly. It is commonly used in investigations of this kind when more detailed and costly calculations methods can not yet be justified.

The methodology centers around the procurement costs of the main equipment items. It assumes, that the other cost components involved in the erection and commissioning of a production plant are directly related to the cost of the main equipment.

In our case, the procurement costs of the individual equipment items have been calculated via the well known formula

$$K_{\mathbf{rc,i}} = K_{\mathbf{bc}} * (p_{\mathbf{rc,i}} / p_{\mathbf{bc}})^{\alpha} * (V_{\mathbf{rc,i}}/V_{\mathbf{bc}})^{\beta} * (P_{\mathbf{rc,i}} / P_{\mathbf{bc}})^{\gamma} \qquad (1)$$

$K_{\mathbf{rc,i}}$ and $K_{\mathbf{bc}}$ are the procurement costs for the individual pieces of equipment in the revamp concepts and the respective base components from our cost data base. p represents the maximum pressure in the components, V the maximum actual volume flowrate and P the thermal or mechanical power transfered, produced or consumed by the equipment.

Obviously, the numerical values applied for the exponents α, β and γ determine the effect of each physical parameter on the cost scaling. If set to 0, the respective parameter becomes insignificant with respect to equipment cost. Quite often the value of 0,67 is mentioned in the literature. For exponent β and relatively simple vessels a relationship between capacity and material requirement can be formulated. In general, however, the value is empirical and has turned out to be a reasonable choice in many cases.

The other major components and activities associated with building a plant such as

- engineering (eng)
- instrumentation (msr)
- electrical (el)
- piping (p)
- procurement and expediting (Pex)
- civil (civ)
- erection (er)
- commissioning (com)

are in their majority related to individual equipment items. It is common experience, that there is a direct relationship between the costs of a piece of equipment and the other cost components listed above. Hence, the overall cost a piece of equipment will contribute to the erection cost of a plant can be expressed as

$$K_{c,i} = f_{eng} * f_{msr} * f_{el} * f_p * f_{pex} * f_{civ} * f_{er} * f_{com} * K_{rc,i} \quad (2)$$

The f_k are cost escalation factors taking account of the other cost component listed above. The entire erection costs for the individual revamp concept j (j = I, II, III) can finally be derived from the equation

$$K_j = \Sigma K_{rc,i} \quad (3)$$

It is also common experience, that the relationships between the cost of the equipment items and the other cost components are not the same for each type of equipment and not as rigid as one would like. Hence, there is a considerable uncertainty margin. An advantage of the method is that, because of the relatively large number of equipment items a major part of the individual errors are equalled out through the integral effect. However, a considerable amount of engineering would be required to increase the precision of this capital cost assessment. It would have gone far beyond the sope of this investigation and should rather be the next step.

Table 2 contains a list of the main equipment items included in the capital cost assessment for each revamp concept.

The capital cost calculations in the end lead to the individual erection cost for the revamp concepts listed in Table 3.

An important part of the real cost of a capacity expansion of this size is the loss of production directly associated with the actual implementation, i.e. the tie-in and commisioning of the revamp part. Obviously, erection time for the revamp section as a whole would extend over several months. A complete shutdown of production for such a long period would render any revamp concept totally uneconomical and could not be tolerated.

Fortunately, the major part of the erection work can usually be carried out with the existing plant in operation. Only the final tie-ins require a standstill of the entire plant. In general, this will be combined with a planned major shutdown, in which some more time consuming service operations are scheduled.

However, an analysis of the revamp concepts with respect to the activities associated with implementation reveals significant differences between them. As the concepts vary mainly in the reforming section, the main differences are related to this equipment. Concept III requires only tie-ins at relatively cold and therefore non-critical piping. It appears justified, that no extra time beyond the scheduled length of the shutdown is required.

Concept II requires the installation and tie-in of a new secondary reformer. Hence, the assumption seems justified that concept II would demand at least one additional week for this work.

Table 2: Main equipment items represented in the capital cost assessment

	revamp concept		
	I	II	III
main equipment item	**SMR en-largement**	**SR with enrich. air**	**ATR**
auxiliary air compressor	x		x
air separation unit		x	x
steam reformer oven box expansion	x		
combustion air fan	x	x	
flue gas fan	x		
secondary reformer replacement / modif.	x		
autothermal reformer			x
fired heater			x
process air preheating	x	x	
combustion air preheating	x	x	
feed / steam preheating coil	x	x	
natural gas preheating coil	x	x	x
steam generator	x	x	x
steam drum	x	x	x
OT synthesis	x	x	x
CO_2 absorber	x	x	x
auxiliary synthesis gas compressor	x	x	x

Concept I requires difficult structural work to enlarge the oven 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 another four weeks.

The additional shutdown periods have been turned into capital costs via the assumption, that one day of production for a plant of this size is at least equivalent to 600.000 USD in revenue. The value is based on fairly conservative figures (ammonia market price 400 USD/mt, energy cost 3,0 USD/MMBTU). This leads to the individual implementation cost for each revamp concept listed in the second last line of Table 3.

Table 3 outlines that, neglecting the rather specific implementation of this revamp concept, concept I appears to be the most attractive solution for this kind of capacity enlargement. However, taking into account the real cost of implementation this picture is completely reversed and revamp concept III becomes the most promising one with respect to capital costs. Concept I actually drops into second position with about 4% higher overall capital costs. Concept II turns out to be the least attractive one with about 7 % higher capital costs compared to concept III.

Table 3: Capital cost of the revamp concepts in USD

	Revamp concept		
	I	**II**	**III**
plant section / cost component	**enlarged SMR**	**sec. ref. with enriched air**	**ATR**
synthesis gas generation	80,4	98,4	93,7
OT synthesis	71,2	71,6	69,1
steam system	2,1	2,2	2,1
reformer waste heat section	3,9	2,9	3,1
overall erection cost	**157,6**	**175,1**	**168,0**
implementation cost	16,8	4,2	---
overall capital cost	**174,4**	**179,3**	**168,0**

Capex / Opex Comparison

Finally, a Capex / Opex oriented comparison of the three revamp concepts has been carried out. There are several ways to set up such a comparison. The general rule is, the more data about operation related costs can be included, the closer the result will be to the actual production cost of the facility.

Since the aim of this investigation is primarily to establish an economical ranking between the revamp concepts only the utilities costs listed in Table 1 are taken into account. The other operating cost components related to e.g. personnel or maintenance are assumed to be fairly similar for all revamp concepts. Hence, they would add equal absolute margins to the operating costs but would not take influence on the ranking.

To illustrate the influence of annual interest rate and payback period for rented capital on the specific production cost two different scenarios have been evaluated. The first one (low interest rate / long payback period) in principal favours capital intensive plants with low specific energy consumption. The second one (high interest rate / short payback period) just the opposite, i.e. plants with comparatively low investment and higher energy consumption. The scenarios are combined with two different specific energy costs (0,75 / 3,0 USD/MMBTU). The results are listed in Table 4.

The ATR-based revamp concept III with the lowest specific energy consumption is also the one with the lowest investment costs. Hence it is not surprising that it also shows the lowest overall production costs.

Table 4: Specific production costs of the revamp concepts in USD/t_{NH3}

			revamp concepts		
			I	II	III
specific energy cost	annual interest rate	payback period	enlarged SMR	SR with enriched air	ATR
USD/ MMBTU	%	yrs.	specific production cost in USD / t_{NH3}		
0,75	5	15	128	130	123
	15	5	329	336	316
3,0	5	15	212	210	203
	15	5	412	416	395

CONCLUSION AND OUTLOOK

A extensive investigation has been carried out by ThyssenKrupp Uhde to assess the economic viability of three different concepts designed to increase the synthesis gas generation capacity of an existing ammonia plant. The expansion of the synthesis gas section constitutes a vital part of an overall production enlargement of the plant.

The aim of the investigation was to establish a ranking between the revamp concepts with respect to their individual economics. The ATR-based concept III turns out to be the most promising solution with the concept I (enlargement of the existing steam reformer) in second position.

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 have rendered only moderate differences between the individual energy consumption figures.

The overall capital costs should include the costs associated with revamp implementation. Especially concept I (enlargement of the existing steam reformer) requires considerably more complicated and laborious implementation work which demands a longer plant shutdown and the related loss in production.