



**Autothermal reforming:
a flexible syngas route with future potential**

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ABSTRACT

The autothermal reforming technology has been widely discussed in the past on conferences and in technical magazines. Compared to the steam methane reformer (SMR) with its multitude of tubes the autothermal reformer (ATR) offers a much simpler mechanical setup and superior operational robustness. The ATR can provide a wider syngas composition range than the SMR, especially significantly lower $H_2:CO$ ratios.

Numerous ATRs are in operation worldwide. However, most of them operate as secondary reformers in ammonia and methanol plants in collaboration with steam reformers. Only a few true stand alone ATRs have been realized up to now. For ammonia and methanol plants the technology has so far been considered uneconomical. Natural gas based GTL plants, for which the ATR is the best choice, have not been built in large numbers as yet.

A comprehensive evaluation of the economic viability of an ATR based ammonia or methanol plant must take into consideration that the ATR requires oxygen. It also provides significantly less waste heat than the SMR for process media preheating and steam generation. So far, the additional operating costs for the utilities and the capital cost for the required equipment have eliminated the cost advantages of the stand alone ATR for ammonia plants and apparently also for methanol plants.

However, with ever growing plant capacities the competitiveness of ATR based plants is improving continuously. Mainly responsible for this are the different building-cost-to-capacity relationships of SMRs and air separation plants.

Uhde established an ATR development programme some years ago. A pilot plant has been built and is in operation since June 2009. There the applicability and robustness of the Uhde ATR concept is being validated.

The paper describes the pilot plant concept, its main design features, experiences made during erection and commissioning and outlines the extensive test program. Also, an ATR based ammonia process flowsheet is presented. Finally, an economic comparison between a conventional SMR-based ammonia plant with an ATR-based plant for a large capacity is discussed.



INTRODUCTION

Gas mixtures composed mainly of hydrogen and carbon monoxide are commonly addressed as synthesis gases. They are the base for important processes such as the syntheses of ammonia and methanol.

As required by the reactions which take place in the syntheses the syngases vary considerably in composition. Table 1 lists the most important processes and names the main requirements with respect to syngas composition.

Table 1: Synthesis gas composition requirements of downstream processes

process	synthesis gas composition requirements
ammonia	$(\text{H}_2 + \text{CO}) / \text{N}_2 \approx 3.0$
methanol	$(\text{H}_2 - \text{CO}_2) / (\text{CO} + \text{CO}_2) \approx 2.0$
hydrogen	max. H_2 content
gas-to-liquids	$\text{H}_2 : \text{CO} \approx 2.0$

Given their main constituents hydrogen and carbon monoxide synthesis gases can be processed from practically any carbon containing feedstock such as natural gas, naphtha, coal or petrol coke plus water via cracking and (steam) reforming. Coking has also been an important source.

Due to the heats of formation of the species involved all these processes are highly endothermic. The transfer of heat to the process gases to force the desired reactions can be effected via heat transfer through surrounding walls or by internal (autothermal) heat generation through partial combustion of the process gas.

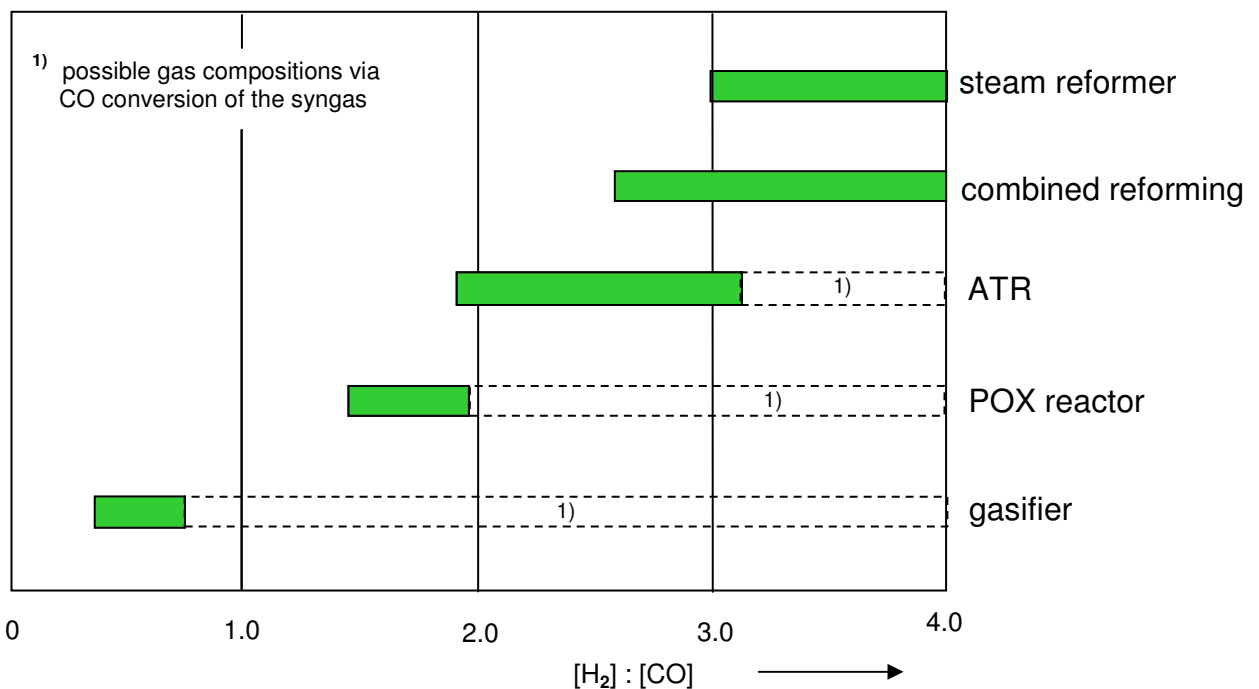
A number of syngas generators are available, competing with each other for the different downstream processes. Table 2 names the most important syngas generator concepts and differentiates between externally and autothermally heated devices.

As Table 2 illustrates, most of the syngas generators operate on hydrocarbons as feedstocks. The coke oven basically extracts the volatile hydrocarbon components from the charged coal (dry distillation). It delivers a syngas composition not too different from the steam reformer, albeit with a significantly higher hydrocarbon content.

**Table 2: Synthesis gas generator types**

syngas generator	feedstock
<u>external heating</u>	
steam reformer	natural gas, naphtha
coke oven	coal
<u>autothermal (internal) heating</u>	
autothermal reformer	natural gas
partial oxidation reactor	natural gas
gasifier	coal, petrol coke

By nature, the syngas generators operating with autothermal heating have significantly higher CO and CO₂ contents in their syngases than the generators with external heating, i.e. much smaller H₂ : CO ratios. In syngas generators with external heating the carbon oxides are to be found in the flue gases of the external heating.

**Fig. 1: Composition ranges provided by syngas generators**

This feature is most significant in the gasifier, where practically no hydrogen is provided by the feedstock. The hydrogen has to come entirely from cracked water mole-



cules, which requires more energy compared to the cracking of hydrocarbons and hence is associated with higher CO and CO₂ production. This is illustrated by the graph shown in Figure 1. The figure shows the individual ranges of H₂ : CO ratios in the syngases which can be provided by the various syngas generators.

In general, matching the individual syngas-based processes with the syngas generators which can provide the required gas compositions straight away should give the most economical solutions. However, as discussed briefly below, the required gas compositions can not always be delivered by one syngas generator type on its own.

- Ammonia synthesis

The ammonia synthesis requires an H₂ : N₂ ratio in the order of 3.0. Obviously, this could not be realized by a steam reformer on its own. However, it can be met easily by combined reforming using an air blown secondary reformer. The version where the secondary reformer provides just the stoichiometric amount of nitrogen required in the synthesis has been the dominant syngas generation concept in ammonia plants for the last fifty years.

Other versions operating with an excess of nitrogen in the secondary reformer and a downstream process step to remove the surplus nitrogen have also been realized in considerable numbers. A stand alone ATR is also capable of feeding an ammonia synthesis. It has to be operated with oxygen-enriched air or pure oxygen combined with a subsequent process step to introduce the required amount of nitrogen.

- Methanol synthesis

Steam reforming alone can not provide the syngas composition according to Table 1 best suited to the requirements of the methanol synthesis. A quite substantial purge from the synthesis or other means, e.g. a pressure swing absorption (PSA), are required to remove the surplus hydrogen. Pure autothermal reforming generates a syngas too low on hydrogen, requiring CO conversion and subsequent CO₂ removal.

Again, combined reforming can generate the optimum gas composition for the methanol synthesis straight away, leading to a significantly better consumption figure for the plant. Other advantages such as a smaller steam reformer size and a higher operating pressure in the syngas generation section are also associated with combined reforming. Hence, although requiring pure oxygen, combined reforming appears to be the superior alternative for larger plant capacities.

- Hydrogen

Hydrogen plants aim for maximum hydrogen production from a given amount of feedstock. Hence, most of these plants are based on steam reforming although several plants based on partial oxidation reactors are still in use. However, as ATR and POX reactors require pure oxygen, steam reforming is nowadays considered the more economic option. If this is valid for larger capacities in excess of 100,000 Nm³/h, e.g. for heavy oil upgrader plants, remains to be seen.



- Gas-to-liquids

As illustrated by Fig. 1 the ATR is the syngas generator best suited to Fischer-Tropsch (FT-) syntheses as it provides the required gas composition straight away. Besides using pure oxygen it is technically feasible to operate the ATR on plain air and feed the nitrogen containing syngas to the FT-synthesis. However, this obviously increases the flowrates in the entire process considerably and does not seem to be the most economic solution.

This brief excursion into the basics of syngas production shows that autothermal reforming has already been an established element of synthesis gas production for quite some time. It nevertheless wants to draw attention to the fact that – with the exception of the stand-alone ATR in GTL plants – all other autothermal reformers so far operate as secondary reformers on pre-reformed gas or mixtures of pre-reformed gas with natural gas. In addition, the effective steam-to-carbon ratios for these secondary reformers are also quite high.

Both the presence of considerable amounts of hydrogen in the pre-reformed gases and the high steam-to-carbon ratios substantially reduce the risk of soot formation in the ATR's combustion zones. Hence, operating experience gained with secondary reformers can not really be considered as a reliable basis for the design of "pure" ATR applications.

Recent years have brought a renewed interest in ATR based synthesis gas generation for ammonia plants. This is closely linked to further increases in plant capacity well beyond today's maximum of 3,300 mtpd realized by Uhde in the SAFCO IV and Ma'aden ammonia plants. As mentioned above in context with methanol plants, steam reformers are coming closer to their technical limits in this capacity range. The ATR technology appears to be a promising alternative here.

This paper presents Uhde's assessment of the economic viability of this approach and outlines what Uhde has done in recent years to prepare itself for this scenario.

AMMONIA PLANTS BASED ON AUTOTHERMAL REFORMING

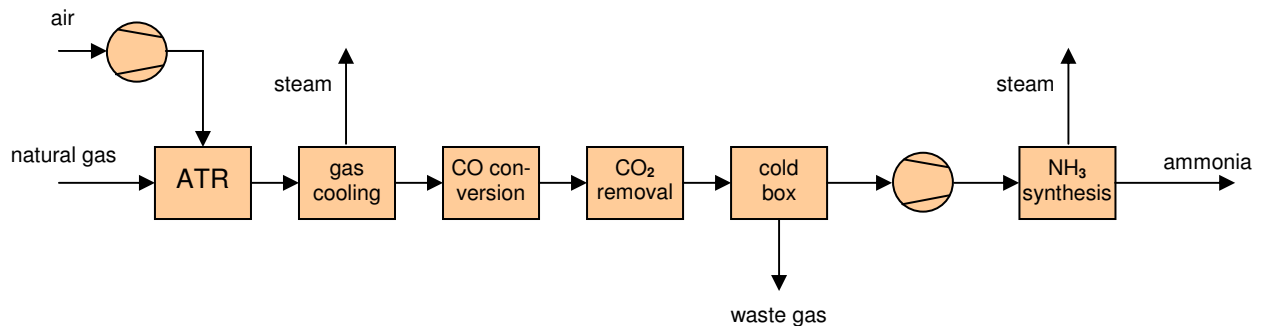
Flowsheet options

An autothermal reformer can be operated with the entire range of oxidator compositions between plain air and pure oxygen. The amount of oxygen which has to be provided by the oxidator is defined by the required reforming heat duty. It does not depend significantly on the overall oxidator composition. The process steps required downstream of the ATR for syngas preparation are however closely linked to the oxidator composition. Below, the consequences of the three most important oxidator compositions plain air, oxygen-enriched air with stoichiometric nitrogen content and pure oxygen on the process configurations as well as on equipment sizes are discussed briefly.

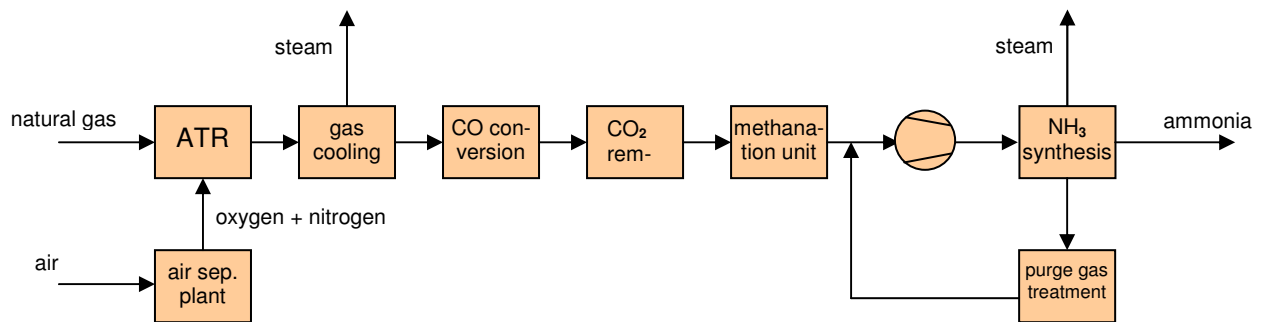
Fig. 2 contains simplified block flow diagrams of the process variations. Obviously, the numbers of process units are not significantly different between the three concepts.



Option 1 – plain air as oxidator



Option 2 – oxygen enriched air as oxidator



Option 1 – pure oxygen as oxidator

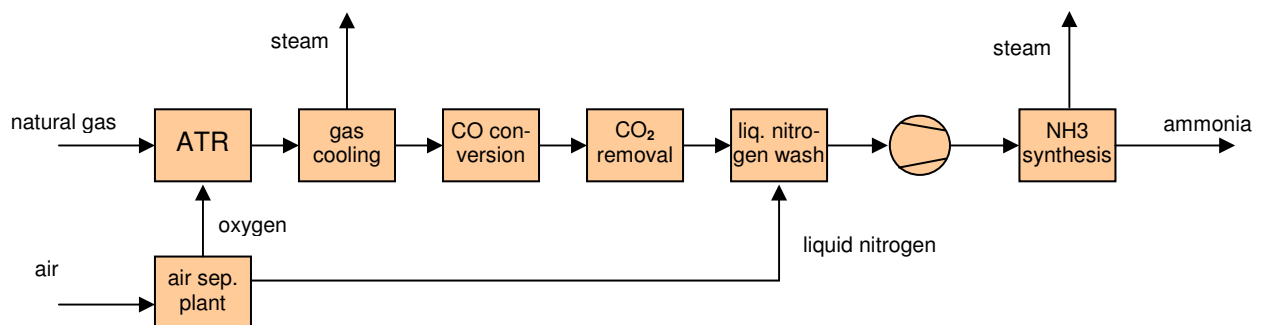


Fig. 2: Flowsheet options of ATR-based ammonia plants



- Option 1 – plain air

Using plain air results in a large nitrogen surplus in the syngas. Hence, a cryogenic unit is required to extract the excessive nitrogen upstream of the syngas compressor or a very large purge stream from the ammonia synthesis. Obviously, this concept is associated with the largest flowrates and equipment sizes at least in the synthesis gas generation section.

- Option 2 – oxygen enriched air

The option requires an air separation unit. The air flowrate is specifically selected to provide the correct amount of nitrogen for the ammonia synthesis. The oxygen coming in with the air and the added amount together meet the required reforming heat duty. Thus, the capacity of the air separation is minimized.

- Option 3 – pure oxygen

In this process concept only marginal amounts of nitrogen depending on the oxygen purity are introduced into the ATR. Hence, flowrates in the synthesis gas generation section take on their smallest values. The nitrogen required in the ammonia synthesis is introduced into the syngas just upstream of the syngas compressor for example via a liquid nitrogen wash unit.

A natural advantage of the process concepts using final cryogenic units in the syngas preparation is the practically inert-free synthesis gas. Without accumulation of inerts the size of the equipment in the ammonia synthesis can be reduced in the order of 10 %.

Option 1 operates with significantly larger flowrates than option 2 in the syngas generation section. This may be compensated by a somewhat simpler cryogenic unit compared to the air separation, smaller flowrates in the synthesis and the missing purge gas treatment. Hence, the two options may have fairly similar investment costs.

Comparing option 3 with option 2 the following can be pointed out. The capacity of the air separation plant has to be about 40 % larger. A nitrogen wash unit is considerably more complicated than a methanation section. Whether this is fully compensated by the somewhat larger equipment of the versions 1 and 2 remains doubtful at least. Hence, option 3 is likely to have the highest investment costs among the three process versions discussed.

Economic comparison with steam reforming based plants

The economics of ATR based ammonia plants compared to plants using conventional tubular steam reformers has been addressed in a number of papers in recent years in more or less detail [1], [2]. The projections about possible investment cost reductions vary between 20% and none. With respect to operating costs an even larger range between 30 % reduction and none can be derived from the references. These assessments do not always reveal which cost components are taken into account and hence can not really be compared.



Uhde's assessment outlined below leads to the conclusion, that any definite figure is misleading since plant capacity is the decisive parameter. The break even point indeed appears to be in the area of 4000 mtpd as it has been reported by other authors. However, with the limited accuracy possible at this moment a range of uncertainty, at least in the order of ± 500 mtpd, should be expected.

An economic comparison between the two plant concepts must look at both operating and investment costs.

Operating cost

About 65 % of the specific energy consumption of a modern ammonia plant is transferred to the ammonia product as chemical energy. The remaining 35 % are needed to enforce the required process steps, i.e. via preheating and compression of the process media. The energy balance of an entire ammonia plant is fairly complex. As both parts are affected by the quite significant changes in the process, the consumption side as well as the available waste heat have to be looked at.

Table 3 contains a comparison of the consumption side. The table lists the most important energy consumers in ammonia plants for both plant concepts. The right hand side column of the table indicates the change in energy consumption which has to be expected for the respective process unit in the ATR-based plant. It has to be pointed out however that the individual energy consumptions of the consumers listed in the table differ significantly. By far the most important consumers are the reformer, the waste heat section/fired heater and the air separation.

The feed flowrate in the ATR-based plant is about 25 % larger compared to a steam reformer based plant as it contains between 50 and 60 % of the fuel flow for the reformer burners. Oxidation of this additional amount provides the heat for the reforming reactions. Due to the resulting increase in natural gas flowrate the related energy consumptions in the syngas generation section are likely to go up, though not always in the same order as indicated in Table 3.

Due to the larger process gas flowrate in the syngas generation section of the ATR-based plant more heat is available from this process stream. The surplus can be used to raise the additional steam required for the air separation.

The large amount of heat contained in the flue gas of the primary reformer is usually utilized for the process media preheating addressed in Table 3 as well as steam superheating. In the ATR-based plant this has to be supplied by a fired heater. Due to the increased steam production more heat for steam superheating has to be supplied by the fired heater in addition to the increased heat demand for feed and feed/steam preheating.

The heater operates with a smaller specific flue gas flowrate than the waste gas duct of the steam reformer. Hence, stack loss as well as heat duty for burner air preheating are reduced compared to the steam reformer.



In summary, our calculations carried out so far show an additional energy consumption for the ATR-based ammonia plant in the order of 6 % for the process version operating on enriched air (option 2). For the version operating on pure oxygen (option 3) a somewhat higher additional consumption in the order of 10 % has to be expected.

Table 3: Comparison of primary ammonia plant energy consumers

energy consumers		add. energy consumption ATR plant [%]
SMR-based plants	ATR-based plants	
reformer (feed + related fuel)	reformer feed	0
waste heat section (related fuel)	fired preheater	+ 15
feed compression	feed compression	+ 25
feed preheating	feed preheating	+ 25
feed/steam preheating	feed/steam preheating	+ 7
process air compression	air separation	+ 150 / + 188 ¹⁾
process air preheating	process air preheating	0 ²⁾
synthesis gas compression	synthesis gas compression	0
ammonia vapour compress.	ammonia vapour compress.	0

1) enriched air / pure oxygen

2) reduced preheating temperature with regard to higher oxygen content

Investment cost

No ATR-based ammonia plant has been built as yet. Hence, a comprehensive cost data base, comparable to that available for conventional plants, does not exist. Therefore, the qualitative analysis presented below is used to derive a cost relation between the two plant concepts.

Table 4 lists the main process steps in an ammonia plant and the respective units in the conventional plant based on steam reforming as well as in the ATR-based plant. The right hand side column of the table indicates whether the respective unit in an ATR based plant is expected to have lower (↓), equal (~) or higher (↑) investment costs. To keep the focus on the most relevant process units the ATR process concept based on enriched air is used for this comparison.

This qualitative approach illustrates that the two process steps “oxidator preparation” and “reforming” are decisive for the investment cost relation between the two plant



concepts. Minor cost differences can be expected for each other process step as equipment numbers, sizes and complexities vary moderately throughout the entire flowsheets. However, excluding the fundamentally different process steps “oxidator preparation” and “reforming” it can be expected that these differences will not all be in favour of one plant concept and hence equal each other out.

Table 4: Investment cost comparison per process step

process step	SMR plant	ATR plant	unit cost ATR : SMR
oxidator preparation	compressor	air separation	↑
desulphurization	preheater / reactor		~
process gas preheating	flue gas duct	fired heater	~
reforming	steam reformer	—	↓
	secondary reformer	ATR	~
process gas cooling	steam generator / superheater		~
CO conversion	2 converter / 2 BFW preheater		~
CO ₂ removal	aMDEA, alternatively physical solvent		~
methanation	preheater / reactor		~
syngas compression	2-casing turbo compressor		~
ammonia synthesis	3 converter dual-pressure synthesis		~
refrigeration	compression type		~
purge gas treatment	hydrogen recovery		~ / ↓ 1)

1) option 2 "enriched air" / option 3 - "pure oxygen" with inert free syngas

The cost structure for a steam reformer has a fairly linear relationship between capacity and unit cost. With few exceptions material and construction costs are directly proportional to the number of tubes, i.e. unit capacity.

The cryogenic air separation is by nature a single train unit. Hence, contrary to the steam reformer, a considerable reduction of the specific investment cost with growing capacity can be expected. This characteristic is illustrated in Fig. 3. The steam reformer cost curve is derived from Uhde's respective database. The figures for air separation plants are derived from proposals provided by plant vendors.

The cost data have been deliberately related to the SAFCO IV reformer costs as it does not make sense to discuss absolute figures in this context. The actual costs depend very much on the region where the plant is to be built, specific requirements and the overall economic situation.

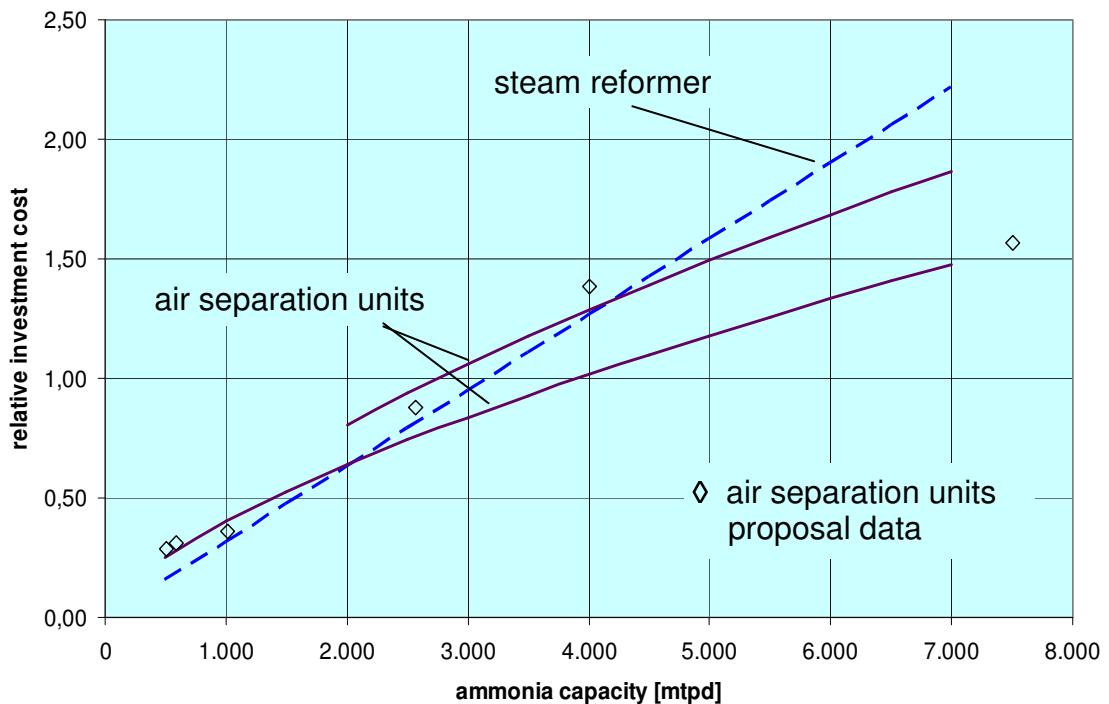


Fig. 3: Investment cost relation steam reformer vs. air separation plant

The figure illustrates that with the cost data available at present a single break-even capacity can hardly be determined. The cost figures of air separation units may be interpreted conservatively or optimistically by the continuous curves as shown in the figure. Obviously, there is a considerable capacity range of uncertainty. However, the figure indicates that – even with a conservative approach – beyond plant capacities of 4000 mtpd an ATR based plant is likely to have growing investment cost advantages compared to a steam reforming based plant.

UHDE'S ATR TECHNOLOGY

Description of Uhde's ATR test facility

The partial, i.e. substoichiometric combustion in autothermal reforming is in principal likely to generate soot. For any set of operating conditions defined by feed gas composition, pressure and steam-to-carbon ratio a minimum operating temperature exist below which soot is generated. Steam and hydrogen in principal reduce the soot forming potential, i.e. move the boundary for soot-free operation to lower operating temperatures. The minimum temperature also depends on the burner design and the flow conditions in the combustion zone.



Uhde has extensive experience in the design of secondary reformers for ammonia plants. These vessels operate under similar pressure and temperature, but in the following two important aspects enjoy considerably milder conditions compared with pure autothermal reformers.

Secondary reformers process partly reformed gases which contain significant amounts of hydrogen and operate with quite high effective steam-to-carbon ratios. Hence, they do not provide a basis to determine the minimum possible operating temperatures with respect to soot formation. This temperature is directly related to the amount of oxygen supplied. Therefore, to minimize plant operating cost the ATR is to operate at the lowest possible operating temperature.

The heat generation rates of secondary reformers are relatively small. With the considerably higher heat generation in autothermal reformers, increased wear and tear on the burner and thermal insulation must be expected. They have to be taken care of by modified design features. Also catalyst bed design and size can be expected to vary.

To gain further insight into ATR operation and to produce the operational data required to adjust the design procedures, Uhde set up a development program several years ago. Subsequently to being successful in finding a cooperation partner in the Russian company Kuibyshev Azot, a test facility was designed, built and commissioned at Kuibyshev Azot's production site.

Fig. 4 contains the process flowsheet of the test facility. The production site usually operates on fairly light natural gas with low contents of higher hydrocarbons and sulphur. Hence, to simulate heavier feedstocks, the test facility was equipped with a device to add propane to the incoming natural gas.

Provisions were also made for the addition of a hydrogen rich stream to the natural gas to simulate a pre-processed gas. With regard to the low sulphur content a further desulphurization of the incoming natural gas was considered unnecessary. It was also decided to take the required process steam from an external source. The heaters for natural, feed/steam mixture and oxidator are powered electrically.

At the ATR gas outlet a water quench is installed. Hence, only moderate gas temperatures are present outside the ATR pressure vessel. The ATR design capacity can be best characterized by a syngas flowrate at reactor outlet in the order of 1000 Nm³/h at 30 bar operating pressure.

The test rig is equipped with optical light extinction measurement devices to detect soot in the gas phase as well as in the process condensate.

During the design phase of the ATR itself, considerable effort was spent to optimize the geometry of the combustion or partial oxidation (POX) zone, gas inlet velocity and swirl and to find the best position for the oxidator nozzles. CFD analysis has been employed extensively for this step. As a result of these studies two ATR reactor tops with completely different shapes of the combustion zone have been manufactured.

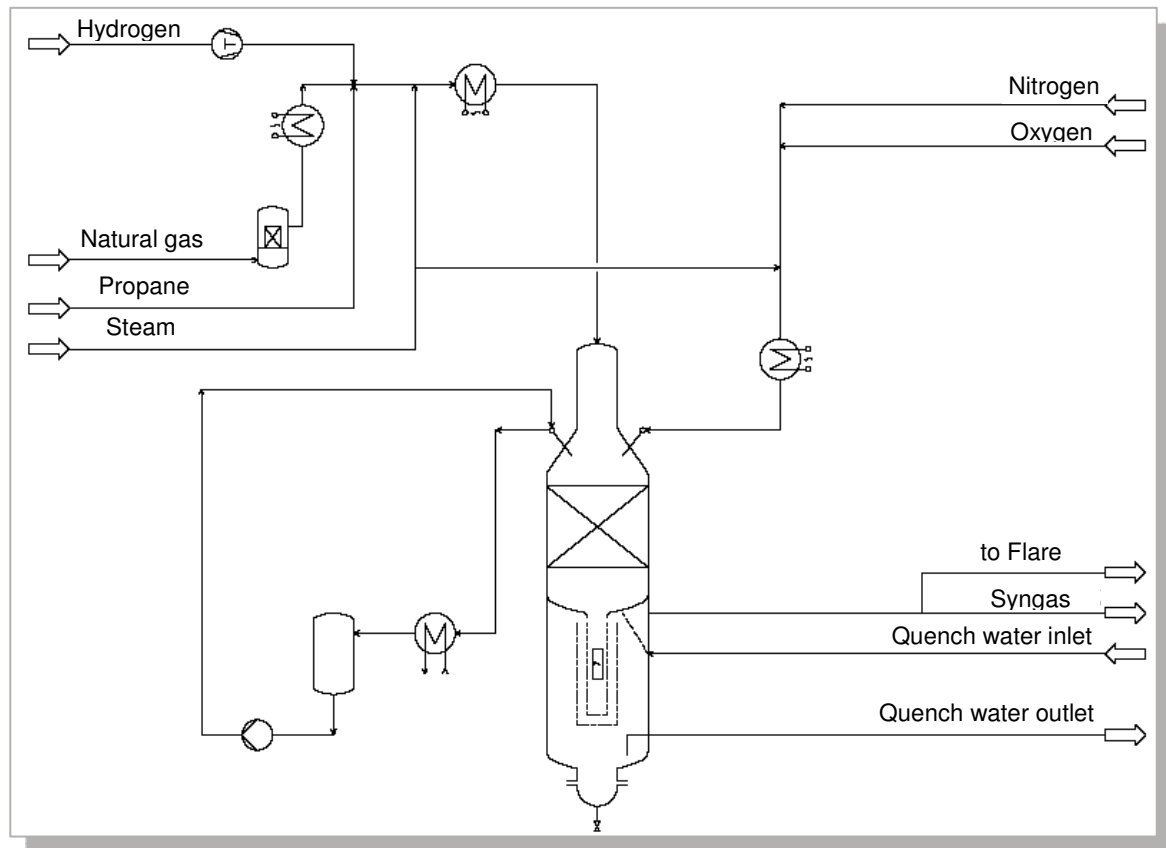


Fig. 4: Process flow diagram of the ATR test facility

With regard to the high design pressure and other restrictions it was decided early in the design phase to abandon the idea of installing an optical access to the combustion zone. Also, the number of nozzles which could be fitted to the reactor pressure vessel, is limited as each one is quite bulky due to the high design pressure.

Fig. 5 shows a CAD drawing of the test rig. The lower steel structures in the front contain the modules for burner cooling, hydrogen compression and feed preheating. The high steel structure at the back contains the ATR itself and the feed/ steam pre-heaters.

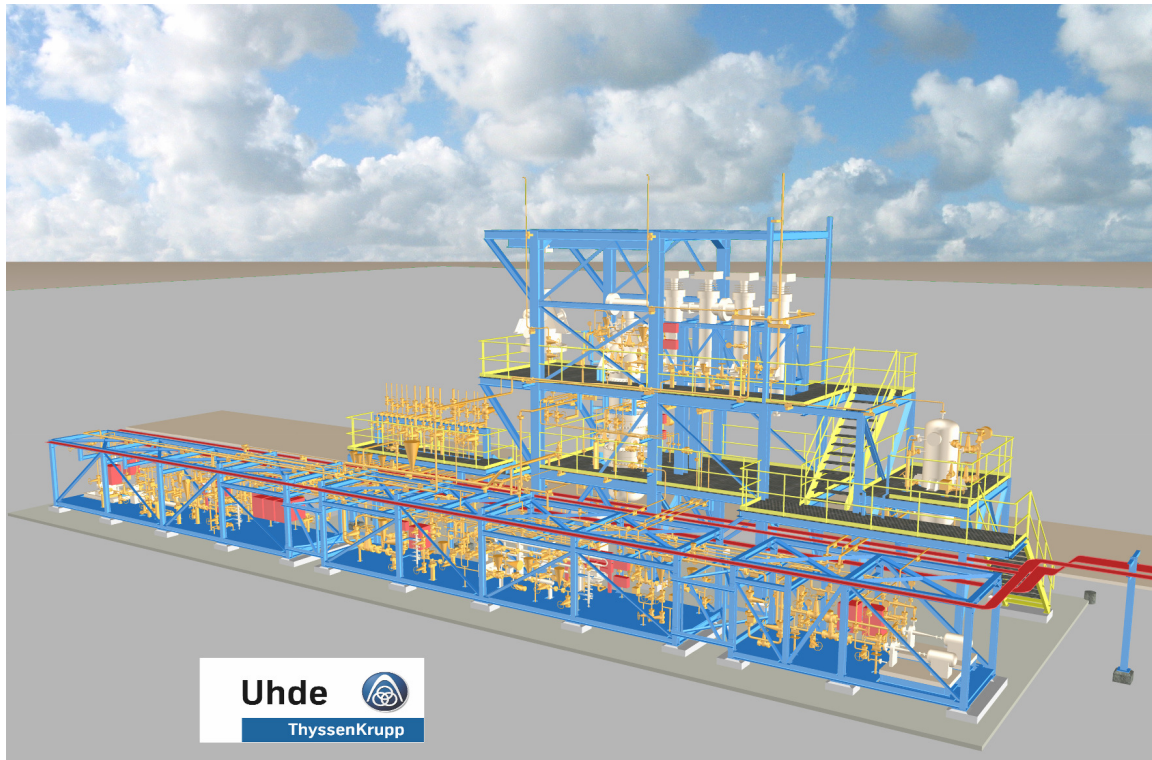


Fig. 5: CAD drawing of the ATR test rig

The process plant occupies an area of appr. 30 m by 11 m. The height of the central steel structure is about 10 m.

The main part of the test rig has been assembled on-site from prefabricated parts. This was considered more cost efficient than complete preassembly and transport of the relatively huge structure to the production site. The three modules in front were preassembled and transported as complete units.

Fig. 6 presents a photograph, showing the test facility in its final state of construction ready for startup. The picture contains a view on the facility from the opposite side compared to the CAD drawing in Fig. 5 with the main structure containing the ATR now in the front. On the right hand side of the picture the pipe rack for the utilities required by the test facility is visible.



Fig. 6: View of the ATR test facility ready for startup

Experiences during erection and commissioning

To obtain building and operating permission for the test rig turned out to be a time consuming and relatively costly procedure. A safety study had to be carried out to show conformity of the plant's design and instrumentation with russian regulations. Also, a special documentation conforming with russian standards had to be generated from the plants engineering documents.

Transportation of the material to the construction site did not present any problems.

Assembly of the test rig presented no major problem and could be continued even through the cold season. It should be mentioned, however, that the winter of 2008 to 2009 was unusually mild. Subsequently to the thorough execution of pre-commissioning and commissioning procedures the ATR was ignited for the first time in June 2009.

Operation of the optical soot detection devices has shown that the lowest soot content which can be detected in the samples taken from the quench water offtake is in the order of 1-3 ppm wt. Quench water and process gas mass flowrates are not really linked but usually of the same order. Hence, the soot concentration determined in the quench water should correspond roughly to the concentration in the process gas.

The minimum measurable soot concentration in the process gas named by the manufacturer of the measuring device should be in the order of 80 mg/Nm³ (0.1 ppm wt.). During the test runs the onset of soot always became apparent at first in the quench water. Hence, the actual sensitivity of the measurement device appears to be at least one order of magnitude lower than pronounced.



The target in the test runs so far has been the detection of the soot formation boundary. Therefore, as soon as the onset of soot was detected, the run conditions have been readjusted immediately to take the reactor away from the critical region. Any prolonged operation in the critical region always demanded a considerably longer period under soot free conditions to get the soot out again.

Operational Results

Depending on the process an autothermal reformer is to serve as sole synthesis gas generator it has to meet certain

- outlet gas requirements
 - composition (e.g. $[\text{CO} + \text{H}_2]/\text{N}_2 \approx 3.0$ or $\text{H}_2 : \text{CO} \approx 2.0$)
 - pressure
 - methane content

To meet these requirements with the feedstock available at the respective production site and present the most economical solution with respect to investment and operating cost the following parameters have to be selected carefully:

- primary design parameters
 - combustion zone shape
 - combustion zone gas residence time
 - inlet velocity feed/steam mixture
 - inlet swirl feed/steam mixture
 - inlet velocity oxidator
 - number of oxidator nozzles
 - oxidator nozzles direction
 - space velocity catalyst bed
- operating parameters
 - steam-to-carbon ratio
 - combustion zone temperature (→ oxygen-to-carbon ratio)
 - oxidator composition

Obviously, it is economically infeasible, to investigate all of the above mentioned parameters in an experimental program. The number of test runs and individual pieces of equipment to be manufactured would be simply too large.

With the capabilities of modern CFD calculations this could be considerably reduced. As discussed in chapter 3.1 an optimization of the geometrical parameters around the combustion zone has been carried out to determine the most promising combinations for two different combustion zone shapes.

On this basis a still quite extensive test program has been devised to cover the relevant ranges of the parameters listed in Table 5.

**Table 5: ATR test program**

parameter	unit	lower limit	upper limit
feedstock higher hydrocarbon content	%	2	14
N ₂ content oxidator	%	0	55
operating pressure	bar	20	30
steam-to-carbon ratio	–	0.5	3.0
combustion zone shape	–	A	B
rel. gas residence time comb. zone	%	50	100
combustion zone temperature	°C	1150	1250

As the first set of tests is specifically designed to determine the soot formation boundary, i.e. the minimum temperature for soot-free operation at given steam-to-carbon ratios, these tests were run without any catalyst in place to avoid damage to the catalyst.

These test runs carried out so far have delivered satisfying results. The performance of the ATR can be characterized by the three most important operating points depicted in Table 6. All points showed reasonable margins against the soot formation boundary despite rather unfavourable feedstock compositions, i.e. relatively high contents of higher hydrocarbons.

Operating point A represents the situation where an ATR generates a stoichiometric syngas composition for an ammonia synthesis operating on oxygen enriched air and a steam-to-carbon ratio of 3.0.

In point B the ATR is operating on pure oxygen. With the steam-to-carbon ratio of 2.0 the conditions tend to ammonia or methanol plants.

Point C represents ATR operating conditions typical for a syngas unit in a gas-to-liquid plant based on Fischer-Tropsch synthesis with an H₂ : CO ratio in the syngas in the order of 2.0.

Since no catalyst had been in place during these test runs the syngas still contains noticeable methane contents. Also, the relatively large specific heat losses of the still rather small test facility lead to extra CO₂ generation compared to a large production plant. Hence, gas compositions at the reactor outlet are not really significant and have therefore not being included in the table.

Table 6: Operational results of the test ATR

Operating point



Parameter	Unit	A	B	C
		NH ₃ syngas, enriched air	NH ₃ syngas, pure oxygen	FT syngas
<u>feed composition</u>				
CH ₄	mol%, dry	89.2	82.9	88.7
C ₂₊	mol%, dry	7.1	14.5	7.4
oxidator N ₂ content	mol%, dry	42.1	5.0	5.0
steam-to-carbon ratio	---	3.0	2.0	0.7
outlet temp. oxidat. zone	°C	1200	1250	1238
oxygen-to-carbon ratio	---	0.85	0.74	0.60
syngas pressure	bar abs	28.0	28.0	28.0
outl. temp. catalyst zone	°C	1086		1100

CONCLUSION AND OUTLOOK

So far only a very limited number of true stand-alone autothermal reforming applications have been realized though the technology has been investigated extensively for more than a decade now. As discussed above, the well established combination of autothermal reformers with steam reformers does not count in this respect. The operating conditions of these secondary reformers are considerably less severe than those of a true stand alone ATR.

The analysis presented in this paper investigates the reasons for this and illustrates – with the main focus on ammonia plants – why the conventional syngas generation based on combined steam and autothermal reforming still dominates newly built plants. The analysis also shows that the cost advantage of the conventional concept against ATR-based syngas generation is shrinking with increasing plant capacity.

The capacity of world scale ammonia plants has grown substantially since the early beginnings and markedly in the past 50 years. Though it is technically and logistically more challenging to provide the equipment for large capacity plants there is no indication that the economy of scale does no longer work for today's world scale capacities.

As discussed before it is difficult to determine the actual break-even capacity, where investment costs of both plant versions become equal. However, there are substantial arguments that for capacities beyond 4000 mtpd the ATR-based plants will have increasing advantages.

Such capacities are also feasible with steam reformer based plants as Uhde has shown with a concept presented on the AIChE ammonia safety symposium 2004 [3]. However, the ATR allows to increase the pressure level in the syngas generation sec-



tion considerably beyond the current limit of appr. 45 bar for steam reforming based plants. Hence, with reduced volumetric flowrates in the syngas generation section it will be much easier to realize larger capacities.

Uhde intends to prepare itself for this scenario. The test program for the ATR will be completed during the next months. The operating data will provide the basis to identify the most promising design and to update the design tools.

The currently feasible maximum operating pressure for the tests covers ATR applications for hydrogen production as well as GTL plants. It also provides a reasonable basis to judge the ATR behaviour under ammonia plant conditions.

Once the test results have been fully evaluated and transferred to the design tools Uhde intends to start offering its ATR technology to customers. In the beginning the focus will be on plants with medium size syngas capacities, e.g. in a revamp project, to limit the upscaling factor from the test facility. In the long run, however, the prime target are larger plant capacities.

Since the consumption figures of ATR based ammonia plants are in the same order as those of steam reforming based plants, there will be no significant differences in the overall CO₂ balances. However, a natural advantage of the ATR-based plant is that a much larger part of the generated CO₂ is always contained in the syngas under high pressure. It is separated in the CO₂ removal section without additional equipment and marginal extra energy expenditure. This allows full conversion of the entire ammonia production into urea. A steam reforming plant would require an expensive removal system to extract the CO₂ from the atmospheric reformer flue gases.

A complete separation of the entire CO₂ could also be realized in the ATR-based plant with little extra investment and an only moderate increase of the plants consumption figure provided the fired heater would be fed with syngas instead of natural gas.

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