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The Uhde Dual Pressure Process – Reliability Issues and Scale Up Considerations

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Prepared for Presentation at

The 47th Annual Safety in Ammonia Plants and Related Facilities Symposium

San Diego, California, USA, 16 – 19 September 2002

1. Introduction

With aiming at improved economics the capacity of new ammonia plants increased continuously over the decades. While the first Uhde ammonia plant built in 1928 had a capacity of 100 MTPD distributed on four reactors, nowadays single train plants are typically around 2,000 MTPD and projects for 3,000 MTPD or beyond are promoted.

Over the past 80 years some step changes in plant capacity can be observed reflecting the development of the process. One of the major steps was M.W. Kellogg´s introduction of the single train concept based on centrifugal syngas compressors in the 1960´s enabling a shift of the capacity to 600 STPD. During the following decades some process variations were introduced, e.g. ICI´s AMV and LCA processes or C.F. Braun´s purifier process, all of these aiming at improved process economics rather than a significant capacity increase.

As early as 1971 Uhde commissioned a 1,400 MTPD plant at a loop pressure of 225 bar. In 1991 the next generation of Uhde ammonia plants started with the BASF Antwerp plant having a nameplate capacity of 1,800 MTPD (now operating at 2,060 MTPD). For this plant design a second ammonia converter was introduced into the loop, the scale up experiences from this project led to design considerations taken into account for the next capacity generation at 3,300 MTPD and beyond.

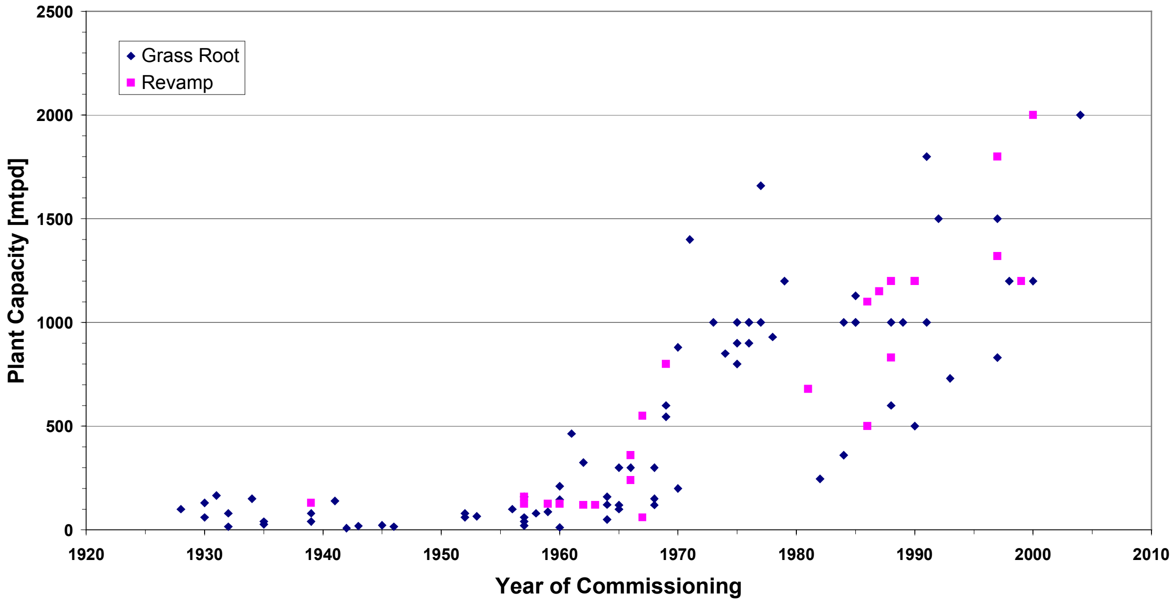


Figure 1: Capacity Development of Uhde ammonia plants

When targeting larger capacities it is not sufficient just to consider what technically can be built or realised, also the risk of the scale up has to be considered for the overall economics of a project, especially when a step change in production capacity is foreseen.

It is the objective of this paper to give an assessment of these risks and to show measures for reducing these. At very large capacities the demand for plant reliability is even more important due to the large amount of capital involved and the significant production losses during any shutdown which will be difficult to compensate for with ammonia from the market.

2. Scale Up risk assessment

When scaling up chemical processes one should distinguish between the typical laboratory to production plant – scale up and the further enlargement of production capacities in commercial production plants. The first mentioned typically involves a mini-plant / pilot-plant and larger demonstration units to prove that the risks resulting from the dimension effect on the chemical processes are minimised and similarity laws can be applied. Scale up factors for such a process developments can be in the range of 10 - 10,000.

Once being in the commercial scale dimension effects become more and more negligible compared to other risks. These are basically introduced from the need of larger equipment which bear the risk of design errors, unexpected mechanical problems / failures or difficulties in the manufacturing process.

In a typical risk assessment the risk is defined as the product of the consequences (C) and the probability (P) of a failure, i.e.:

$$R = C * P$$

In this paper we would like to focus on the consequences on the equipment and production only, for a complete risk assessment also injury to people and environmental impact have to be considered. With the following table the consequences can be classified:

Consequence Level	**Equipment Damage**	**Cost of non-availability**
Low	< 5,000 US$	< 1,500 US$ / shift
Medium	5,000 – 50,000 US$	1,500 – 15,000 US$ / shift
High	> 50,000 US$	> 15,000 US$ / shift

Concerning the consequence level the cost of non-availability (i.e. production loss or increased operating cost) will typically be higher than the equipment damage itself. The given cost limits represent a typical large scale ammonia plant.

The probability of failure should be evaluated and graded as low, medium and high as well. The probability values are estimated incidents of defects/damage/degradation per year provided the equipment is operated in its designated operating conditions. The probability is based on the theoretical failure mode depending on type of equipment, process data and material used for the particular media, and real life experience.

The consequences and probability factors can be plotted in a matrix, which will then identify the associated risk.

Consequence				
	High	Medium High Risk +	Medium High Risk +	High Risk ++
	Medium	Medium Risk O	Medium Risk O	Medium Risk O
	Low	Very Low Risk -	Low Risk -	Low Risk -
		Low	**Medium**	**High**
	Probability			

In many cases when major design changes or new processes for ammonia plants were introduced in the past the first plants of these new generations faced significant difficulties in the start up phase and first years of operation due to the increased probability of design problems when introducing changes to a proven design.

Based on this experience the importance of references is increasingly considered by production companies, engineering companies and project financing institutions. With respect to commercial scale ammonia plants scale up factors have therefore been in the order of 1.3 to 1.5 in the past. By applying these principles and considering a capacity of 2,000 MTPD as proven technology a capacity increase to 2,600 to 3,000 MTPD seems reasonable within the next decade unless other measures for risk reduction are applied.

3. Plant Reliability

When considering the plant reliability a focus has to be laid on number of plant shutdowns and shutdown time, the latter being crucial for the economics of the plant.

About 80% of all unplanned shutdown time can be contributed to equipment related failures, the remaining being basically electrical/instrumentation failures or human error.

A survey on the distribution of shutdown reasons with respect to the plant section is given in fig.2-(ref. 1). These data are in line with earlier investigations, i.e. the reliability issues of ammonia plants did not really change over the past 15 years.

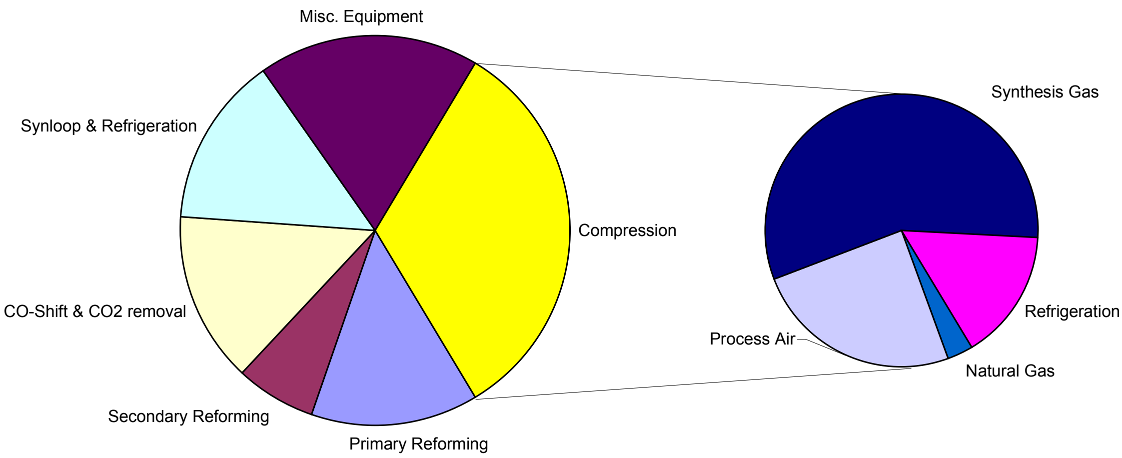


Figure 2: Distribution of plant sections causing a shutdown

About one third of all shutdowns are caused by the machinery and a closer look at these shows that the synthesis gas compressor has the most significant impact. This compressor is very critical in respect of its rotor-dynamics due to the high speed and large bearing span being a result of the difficult compression task requiring a number of impellers. Furthermore the sensitive sealing system is contributing considerably to the downtime. Depending on the type of failure occurred the consequences will be medium, in most cases even high. Therefore the syngas compressor can be seen as one of the most critical equipment in the ammonia plant. According to the survey this turboset contributes to more shutdowns than the entire primary reforming section.

For the risk evaluation of large capacity ammonia plants besides the syngas compressor also the other critical units should be addressed, basically the reforming section and the synthesis loop.

4. General scale up considerations for ammonia plants

The ammonia plant consists of various unit operations, some of which are more critical than others. The conventional process utilising primary reforming followed by secondary reforming, shift, carbon dioxide wash and methanation to produce synthesis gas for the ammonia synthesis loop has now been used for decades and will probably still find use in the large scale plants, which are being considered now.

There are of course alternative ways to produce synthesis gas for the synthesis loop and they have advantages in some area but also disadvantages in other areas. Few of the alternative processes have been used for ammonia production and one should therefore carefully evaluate the process risk before abandoning the conventional process.

The table below lists the areas of concern both for design and for scale-up considerations.

Equipment	**Critical Design**	**Critical Scale Up**
Reforming Section	++	-
Process Gas Cooler	+	+
Process Air Compressor	+	O
Shift Reactors	--	O
Front End Piping	--	-
CO2 removal	-	+
Syngas Compression	++	++
Synthesis Equipment	+	+
Synthesis Piping	O	++
Refrigeration Compressor	+	O
(++: very high, +: high, O: medium, -: low, --: very low)		

While the reforming section is of critical design it is not too difficult to scale the section up to a capacity larger than 4,000 MTPD, at least for a top fired box design. In the frontend of the ammonia plant we notice that the process gas cooler and the carbon dioxide removal section are the areas of greatest concern when scaling up. The shift reactors and process air compressor are moderately critical when considering larger capacities.

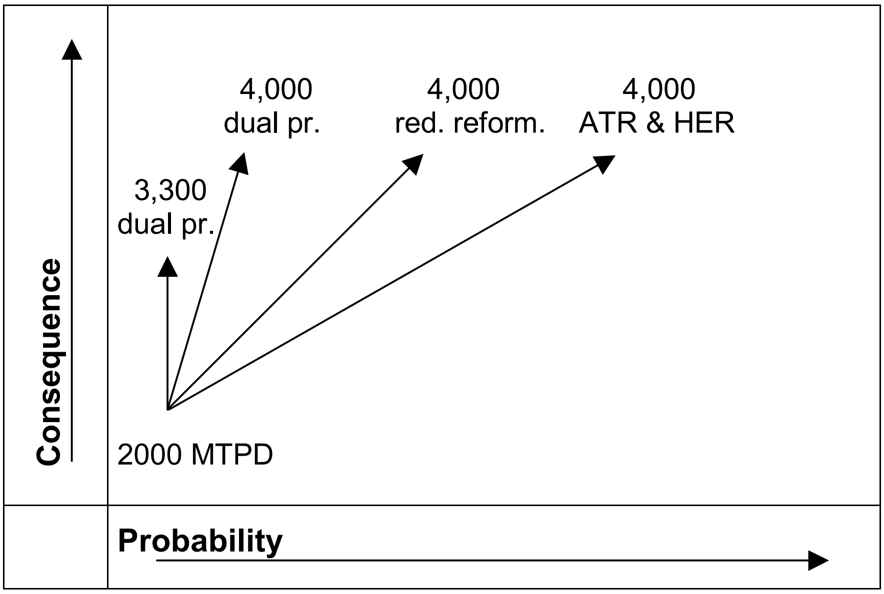
Compared to the conventional process, a process which utilises extra load on the secondary reformer by either oxygen enrichment or excess air to the secondary reformer and consequently a smaller or no primary reformer will in the frontend have a larger carbon dioxide wash, one of the critical areas for scale up due to fabrication and transportation of the large columns issues. When adding oxygen to the secondary or auto thermal reformer, the potential for over-reduction of the high temperature shift increases.

A process with heat exchange reforming will normally replace the heat input from primary reformer with additional oxygen and/or air to the auto thermal reformer coupled with a fired heater for preheat of feeds. These processes also need a larger carbon dioxide removal section. By eliminating a fired reformer these processes eliminate a section, which is critical to design but not critical in terms of scale-up – at least to capacities in the range of up to at least 4,000 MTPD. On the other hand the heat exchange reformer schemes do reduce the amount of steam raised from the ammonia process and in that respect offer a benefit in smaller pipe sizes in the steam system.

For the time being it is hard to see that there is a process, which is more suitable for the large scale plants than the conventional process which is being built today. The evaluation of the different options is summarised in the table below:

	Conventional Process		**Reduced Primary Reforming**	**Autothermal Reforming**		**Heat Exchange Reforming**	
				without tubular reformer	with tubular reformer	Oxygen blown ATR	Air blown ATR
			with Purifier	with ASU	with ASU	with ASU & N_2 wash	without ASU, with Purifier
Status	Proven Process		Proven Process	Proven for Syngas	Proven for Syngas	Proven for very small scale	Some-what similar to O_2 case
Frontend Size	base		larger	larger	larger	smaller	larger
HT-Shift				very critical unless lower presure	critical unless lower pressure		
CO_2 wash			larger	significantly larger	larger	significant-ly larger	larger
Syngas			almost inert free			inert free	almost inert free
Syngas Compres-sion & Synthesis	Scale Up	Dual Pres-sure					
Compres-sion	critical	proven	reduced, proven	critical	critical	reduced, proven	reduced, proven
Synthesis	critical	proven	moderate risk	critical	critical	moderate risk	moderate risk

Considering the risk of the different overall process schemes and applying the consequence/probability diagram the following evaluation can be made.



In general the higher the capacity and the further away from the proven process schemes the higher will be the risk. The processes, which utilise autothermal reforming schemes, must for the time being be associated with a higher level of risk when considered for the ammonia process due to the relatively little use in the field of ammonia production. Instead of a process based on ATR alone it is more likely that oxygen enrichment will be used in the secondary reformer in cases where the limits of the tubular reformer have been reached. Due to economic disadvantages of an ATR design (when considering also the air separation unit) compared to a conventional reforming scheme it can not be expected, that commercial plants in the range of 2,000 – 3,000 MTPD will be based on autothermal reforming in the near future, so there will be no moderate scale up in capacity when changing to an ATR design.

Piping

For the design of the equipment nozzle loads introduced by the piping have to be considered. Besides axial forces as a result of the thermal expansion bending moments have to be taken into account. These bending moments have typically a more significant effect, especially when introduced in the longitudinal direction of the equipment. The stiffer the piping is the higher will be the nozzle loads and the more measures have to be taken to relax these, e.g. stiffening pads on the equipment or re-routing of piping. The stiffness of the piping is proportional to its moment of inertia which is a function of its outside and inside diameter:

$$I \sim (d_O{}^4 - d_I{}^4)$$

While the low pressure piping is basically affected by the nominal diameter the high pressure piping also suffers from the wall thickness. Since for large capacity plants the equipment is also larger, the nozzle forces can be increased but often the limits for nozzle forces for rotating equipment make it necessary to modify the piping system around those pieces of equipment.

With respect to piping it is possible to design but pipe sizes may see a relative increase more than just the capacity factor due to the need for more flexibility in the pipe system. This of course also adds to the cost of the plant and results in a higher pressure drop of the process.

For very large plants the availability of standard piping elements for the synthesis loop is of significant importance since standard flanges or fittings in the required 1500# rating are only available up to 24”. Everything beyond this limit would require a special design with own authority approval and thus would drive plant cost up considerably.

Reforming

For a top fired box reformer the impact of higher capacity is not large. For example, increasing the plant size from 2,300 MTPD to 3,300 MTPD increases the number of row by two from six to eight and each row has now three more tubes. The heat flux and tube sizes are unchanged between the two cases. Flow distribution in the reformer, both on the process side and on the combustion side needs to be checked but there are very reliable computer programs available to perform such tasks. These tools are already being utilized today.

Modular construction of the waste heat section is possible up to 4,000 MTPD. Larger tubing in the waste heat section will be used as needed for the process and steam coils.

The secondary reformer represents no technical problem being a cold-walled refractory lined reactor. Even larger types of this equipment are considered as autothermal reformers for GTL plants.

The process gas cooler is in plants constructed by Uhde a horizontal fire-tube boiler and at 3,300 MTPD is at its maximum size, which can be build today. It is anticipated that this size limitation will be relaxed once a demand for plants larger than 3,300 MTPD materializes. For plants larger than 3,300 MTPD it will, in the conventional process, be necessary to use two parallel boilers. This is of course technically possible and should not introduce significant additional risk; parallel boilers have been used before in many plants.

Shift section and CO_2 removal

In the shift section advances in catalyst manufacturing and technology have reduced the required catalyst volumes substantially. The main issue is more transportation of the low temperature shift converter to and at the site. In the CO_2 removal section the circulation rate increases with plant capacity and larger pumps often require higher NPSH and that leads to elevated columns or parallel and smaller pumps. While increasing the elevation of the column adds cost parallel pumps can reduce reliability unless proper stand-by pumps are provided. Single absorbers and regenerators are possible in plant sizes up to 4,000 MTPD at least for the conventional process (i.e. as long as the primary reforming duty is not reduced).

The semi-lean solution pumps have to be configured already as 3 x 50% below 3,000 MTPD, and this configuration is able to be handle capacities beyond 4,000 MTPD in a conventional frontend.

Synthesis Gas Compression

The synthesis gas compressor is by far the most critical piece of equipment in the ammonia plant. As stated previously it accounts for about 20% of shutdowns in ammonia plants. Over the last thirty years the centrifugal synthesis gas compressor has been used in plants from about 600 MTPD to now about 2,100 MTPD. The smaller compressors operated at higher speeds in the range of 14,000 rpm and the larger compressors now operate at about 10,000 rpm. This is related to limits for tip speeds of the compressor and turbine.

The largest LP-casing in operation today has been used in plants from 1,100 MTPD up to 2,000 MTPD. At 3,000 MTPD it will be necessary to move to the next casing size if the process is simply scaled up. Alternatively, the process can be modified slightly and an existing machine can be used.

Synthesis and catalyst

The size of the synthesis loop is a function of the conversion per pass across the ammonia converter(s). In synthesis loops where high conversion is desired the result is most often that three catalyst beds distributed in two converters is used. Since the conversion to ammonia is enhanced by low temperature at the exit of the catalyst bed, it has been considered to cool the catalyst beds by raising steam directly within the beds as opposed to external cooling as is the practice today. These steam raising reactors have so far not been considered viable due to the risk of boiler feed water entering the catalyst bed directly. This can be considered a case, where the consensus is that the benefit is not worth the risk even though such reactors have been used successfully in for example methanol synthesis loops.

A very effective way to get higher conversion per pass would be with a catalyst with lower ignition temperature than those currently in use. Such a catalyst has not yet emerged. The high activity catalysts, which are known, appear to have ignition temperatures similar to magnetite catalyst and are able to improve conversion per pass by extending the high end conversion. However, at that end, the chemical equilibrium limits the conversion.

5. Dual Pressure Process

To overcome the bottlenecks of standard piping and to reduce the compression demands for large capacity ammonia plants Uhde and Synetix presented the Uhde Dual Pressure Process in 2001 (ref. 2) This process is characterised by a once through ammonia synthesis located at an intermediate pressure level upstream of the ammonia synthesis. The new flowsheet delivers a capacity of 3,300 MTPD using well tried and tested equipment. It also provides the basis for even larger plants, e.g. 4,000 MTPD.

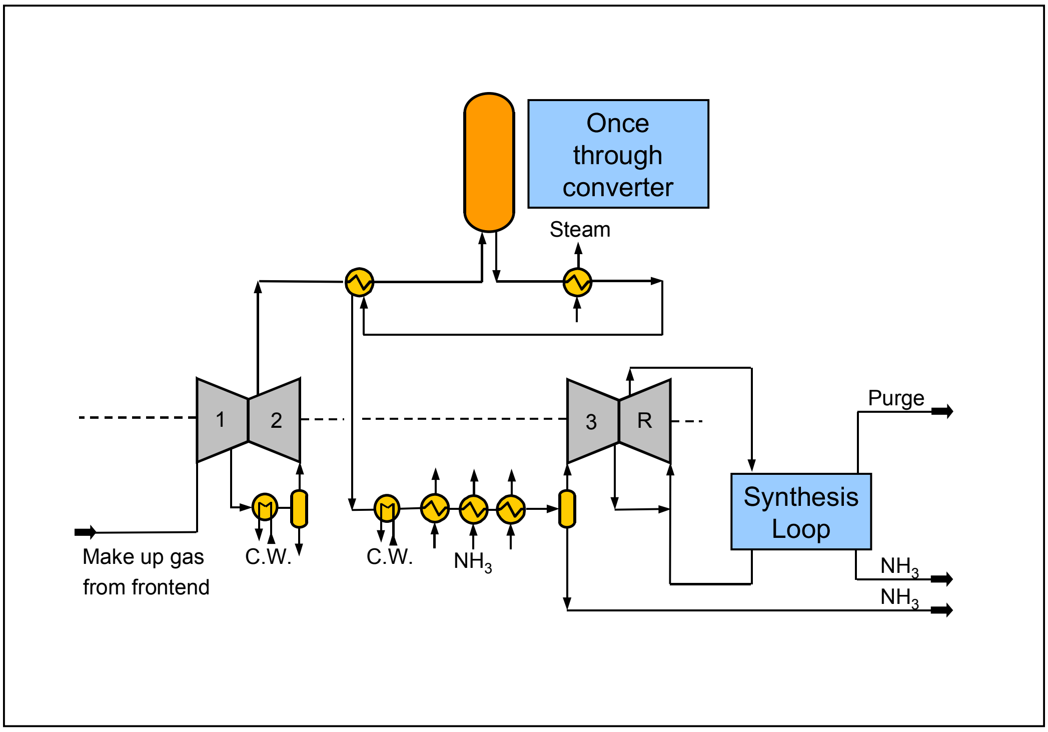


Figure 3: Principle of the Dual Pressure Process

Conversion of synthesis gas to ammonia at a low pressure requires a special high activity catalyst. The first of the generation of low pressure catalysts, Synetix's AMV catalyst, has been in commercial operation since the mid 1980s and has proved to be extremely reliable and trouble-free. With the basic material being magnetite a continuous future low pricing compared to some novel precious metal catalysts is ensured.

The ammonia synthesis configuration (in respect of which patent applications have been filed) in such a process consists of the following stages:

1. Compression of make-up gas is carried out in two steps, first in a two-stage inter-cooled compressor. This is the LP casing of the synthesis gas compressor. At the discharge of the compressor the pressure is about 110 bar. The pressure range is comparable to the low-pressure processes, which operate with a single-casing synthesis gas compressor. A three-bed, inter-cooled, once-through converter in this location can produce about a third of the total ammonia. Included with the make-up gas and fed to the reactor is recovered hydrogen and nitrogen from the purge gas recovery unit. This unit can be either a membrane unit or a cryogenic unit.

2. The effluent from the converter is cooled and ammonia product is separated. Final cooling is done in stages against ammonia chilling. About 85% of the ammonia produced is separated from the gas, which is further compressed to the synthesis loop operating pressure.

3. The HP compressor casing thus operates at a much lower temperature than normally seen. The benefit of the deep chilling is that each impeller in the first stage of the HP casing produces more head than is typical for an inter-cooled synthesis gas compressor. That offsets most of the pressure drop throughout the once-through converter.

4. The ammonia synthesis loop operates in the normal pressure range of up to 210 bar. The high synthesis loop pressure is achieved by the combination of the chilled second casing of the synthesis gas compressor and a slightly elevated front-end pressure. The front-end pressure required is within the commercial ammonia plant experience of Uhde.

The detailed flowsheet shows that several measures were taken to restrict the number of additional equipment to a minimum by combining chillers having the same refrigeration level. Furthermore the equipment downstream of the make up gas synthesis fulfils a dual function, it is used for separation of the product from the make up gas synthesis as well as chilling of the synthesis gas for the next compression step.

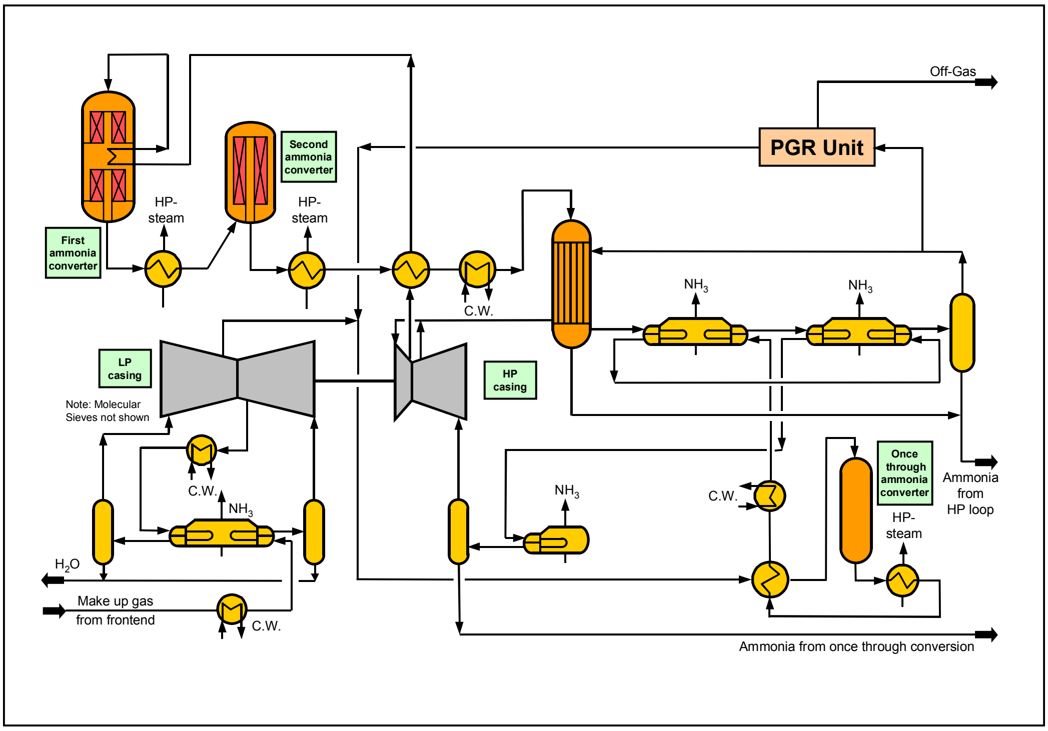


Figure 4: Uhde Dual Pressure Process

By removing ammonia product at the intermediate pressure, the optimum inert level in the high-pressure synthesis loop is reduced somewhat in comparison with today's norm. That is because only about two thirds of the purge needs to be recompressed to the synthesis loop pressure. About one third is converted to ammonia and removed as product at the pressure level the purge gas is rejected to. The lower inert level improves the operating conditions in the high-pressure ammonia converter; further capacity increase of the synthesis loop is, therefore, possible. Another benefit is that the higher dew point of the reactor effluent stream enables the condensation of ammonia product to be carried out more effectively with more condensation occurring against cooling water. Overall, the result is an energy-efficient ammonia plant design – a study using climatic data typical of Middle East conditions indicates an ISBL consumption of about 7.1 Gcal/tonne (-33°C product).

Thanks to the low inerts level in the once-through conversion, the partial pressure of the reactants is higher than in existing low-pressure loops. Therefore the amount of high-activity catalyst in the dual-pressure process is much less than in the existing low-pressure processes. Another advantage is that, in contrast to a high-pressure once-through converter, the first catalyst bed in the dual-pressure process is thermodynamically limited to an acceptable temperature, thus eliminating the need for ammonia injection to moderate the temperature rise over fresh catalyst.

6. Risk Assessment of the Uhde Dual Pressure Process

Following the evaluation scheme of the scale up considerations the individual risk assessments within the units of the dual pressure process are given below.

With respect to piping front end and synthesis piping should be considered separately. Due to the normal capacity scale up in the front end by 65% also the piping diameters would have to be increased by approx. 30% but under consideration of the higher frontend pressure the increase is in the order of 23% only. Availability of standard piping elements is not an issue, due to the stiffer piping some lines have to be changed in their routing and higher nozzle forces have to be taken into consideration. With respect to machinery the discharge line of the process air compressor has to be checked, but compared to other process schemes requiring increased process air the risk is rather low.

The synthesis piping benefits from the fact that it remains basically a 2,000 MTPD synthesis for an overall production of 3,300 MTPD. So availability of standard piping elements, piping stiffness and nozzle loads are in the referenced range and do not contribute to an increased risk. It should be highlighted that the maximum synthesis piping diameter required in the 3,300 MTPD plant is 20”. With the availability of also 24” standard flanges and fittings the synthesis has a scale up potential of beyond 4,000 MTPD.

In the reforming section no risk arises out of the design of a larger furnace box since significantly larger reformers have been built before. Compared to the 3,300 MTPD reformer the largest reformer based on Uhde design has about 100% more tubes. What has to be considered in the risk evaluation is the 3 bar higher reforming pressure. A higher reforming pressure has usually a detrimental effect on the reliability of the reformer outlet piping and leads to an increased potential for metal dusting in the HP steam superheater upstream of the HT Shift. With respect to the reformer outlet Uhde reformers have been equipped for decades with a cold outlet manifold, i.e. hot pigtails are avoided. This design has proven to be extremely reliable and one of Uhde´s ammonia plants is in operation with an even higher reforming pressure trouble free since 1989. Concerning the metal dusting issue the distribution of steam generation and steam superheating downstream the secondary reformer was shifted back towards increased steam production already for the recent plants. By keeping the metal temperatures of the superheater outside the metal dusting region a safe measure of prevention is ensured at the cost of a slightly increased energy consumption until a long term proven economic metallurgical solution is offered.

For the CO Shift section the higher reforming pressure poses an additional risk for the HT shift catalyst with respect to overreduction of the catalyst and the subsequent formation of hydrocarbons. As part of the development programme for the new process, Synetix addressed that demand and tested their latest generation HT shift catalyst, KATALCO 71-5, to confirm its suitability under the proposed conditions.

Concerning the CO_2 removal unit the risk is basically introduced by transport and handling dimensions of the columns. Since it can be expected that these very large capacity plants are located close to the sea side with corresponding export facilities, transportation of the equipment to the construction site should not be seen as too critical.

A capacity of 3,300 MTPD was chosen for the new flowsheet to utilise a synthesis gas compressor of a size currently in use in today's 2,000-MTPD plants. Based on the process concept, the design was fine-tuned in close collaboration with Nuovo Pignone to optimise the compressor and turbine design. The low-pressure casing is identical in size to that used in a new 2,000 MTPD plant. The high-pressure casing is the same as is in use in two plants built by Krupp Uhde in the early 1990s. The whole compressor train for the dual-pressure process is actually smaller with respect to casing size than some which are in operation in 2,000-MTPD plants. By having a number of references for the most delicate equipment in an ammonia plant the risk of the capacity leap from 2,000 MTPD to 3,300 MTPD is reduced considerably.

The introduction of a once through reactor could pose a risk by changing the process configuration. However, the application of make up gas reactors is not new, they have been used in ammonia plants before. The only difference to the previous application is that they have been used directly upstream of the synthesis loop, i.e. at the high pressure. By shifting the reactor to the intermediate pressure level it is increased in size but even larger reactors have been built as loop reactors before at similar pressure levels (CIL - Terra Courtright and Zhong Yuan AMV plants). On the other hand the risk of overheating the first bed is eliminated at the intermediate pressure since the reaction is limited thermodynamically.

Due to the fact that in the 3,300 MTPD dual pressure process the ammonia synthesis has to produce only two thirds of the overall production the synthesis equipment is essentially of the same size as in a 2,000 MTPD plant and can therefore be considered as proven design.

There are two main contributions to the refrigeration duty; chilling downstream of the once-through converter and chilling in the high-pressure loop. The overall refrigeration requirement per tonne of ammonia increases slightly, by about 10%. A single turbine-driven compressor will be used for all refrigeration duties. Conventional, well-proven machine designs can be utilised for both the compressor and the steam turbine, i.e. the higher refrigeration capacity does not contribute significantly to the risk increase of the scale up.

7. Conclusion / Summary

A step change in ammonia plant capacity always involves a certain risk by introducing new and un-referenced equipment. Especially critical equipment of an ammonia plant are the reforming section and the synthesis gas compressor. Concerning the piping especially the synthesis piping has to be evaluated in detail, not only based on its limitation of standard piping elements to 24" but also with respect to increased stiffness and nozzle loads at larger diameters and wall thickness.

Since significantly larger top fired reformers have already been built for other processes the reformer can be excluded from the scale up risk assessment. With the dual pressure process the syngas compression requirements are reduced considerably and production load is shifted from the synthesis loop to the once through synthesis. The risk of designing a 3,300 MTPD plant via the dual pressure process reduces the scale up risk considerably since it can be built with all proven high pressure equipment. Also a further scale up to 4,000 MTPD or beyond involves less risk based on the lower compression and synthesis equipment/piping requirements compared to other processes.

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