Refinery hydrogen management

Managing current hydrogen infrastructure and planning for future requirements requires careful selection of the best combination of recovery, expansion, efficiency improvements, purification and new H_2 supply options

Ruth A Davis and Nitin M Patel
Air Products and Chemicals Inc

Hydrogen management has become a priority in current refinery operations and when planning to produce lower sulphur gasoline and diesel fuels. Along with increased H_2 consumption for deeper hydrotreating, additional H_2 is needed for processing heavier and higher sulphur crude slates. In many refineries, hydroprocessing capacity and the associated H_2 network is limiting refinery throughput and operating margins. Furthermore, higher H_2 purities within the refinery network are becoming more important to boost hydrotreater capacity, achieve product value improvements and lengthen catalyst life cycles.

Improved H_2 utilisation and expanded or new sources for refinery H_2 and H_2 purity optimisation are now required to meet the needs of the future transportation fuel market and the drive towards higher refinery profitability. Many refineries developing H_2 management programmes fit into the two general categories of either a catalytic reformer supplied network or an on-purpose H_2 supply.

Some refineries depend solely on catalytic reformer(s) as their source of H_2 for hydrotreating. Often, they are semi-regenerative reformers where offgas H_2 quantity, purity, and availability change with feed naphtha quality, as octane requirements change seasonally, and when the reformer catalyst progresses from start-of-run (SOR) to end-of-run (EOR) conditions and then goes offline for regeneration. Typically, during some portions of the year, refinery margins are reduced as a result of H_2 shortages.

Multiple hydrotreating units compete for H_2 – either by selectively reducing throughput, managing intermediate tankage logistics, or running the catalytic reformer sub-optimally just to satisfy downstream H_2 requirements.

Part of the operating year still runs in H_2 surplus, and the network may be operated with relatively low H_2 utilisation (consumption/production) at 70 to 80%. Catalytic reformer offgas H_2 supply may swing from 75 to 85% H_2 purity. An H_2 purity upgrade can be achieved through some hydrotreaters by absorbing heavy hydrocarbons. But without supplemental H_2 purification, critical control of H_2 partial pressure in hydroprocessing reactors is difficult, which can affect catalyst life, charge rates, and/or gasoline yields.

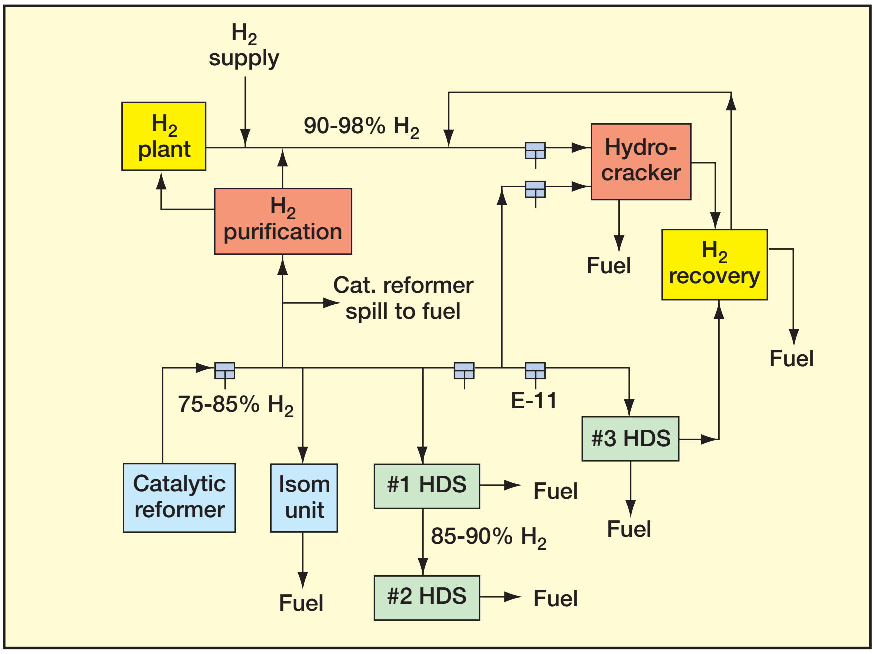

Figure 1 *Refinery hydrogen network*

More complex refineries, especially those with hydrocracking units, also have on-purpose H_2 production, typically with a steam methane reformer (SMR) that utilises refinery offgas and supplemental natural gas as feedstock. The SMR plant provides the swing H_2 requirements at higher purities (92–99+% H_2) and serves an H_2 network configured with several purity and pressure levels. Multiple purities and existing purification units allow for more optimised hydroprocessing operation by controlling H_2 partial pressure for maximum benefit. Typical H_2 utilisation is 85% to 95%. When the refinery SMR plant begins reaching its nameplate H_2 capacity during catalytic reformer H_2 production swings, hydroprocessing bottlenecks again reduce refinery throughput and operating margins. Furthermore, the energy bills associated with the refinery SMR plant's efficiency should be benchmarked to find cost savings that go right to the refinery's bottom line. SMR plant on-stream reliability also takes on a higher priority to maintain high margins.

A refinery H_2 management programme should be organised to meet some of the following objectives, depending on specific refinery configuration:

Maximise H_2 utilisation through increased recovery;

Decouple catalytic reformer operation with H_2 production needs;

Take advantage of higher H_2 purity within specific consumers;

Reduce costs of on-purpose H_2 production;

Improve on-purpose H_2 production reliability;

Expand on-purpose H_2 production;

Evaluate refinery H_2 plant shutdown economics;

Integrate with new industrial gas company H_2 supply.

Ultimately, the programme defines recommendations that balance overall costs with refinery benefits through implementing the best combination of recovery, expansion, efficiency improvements, purification and new supply options.

Network analysis

The refinery network analysis begins by developing an overall network model, defining the current network flow scheme, establishing relevant operating cases, and reconciling H_2 balance information for each limiting case. Figure 1 (previous page) shows a conceptual H_2 network configuration for the refinery sourced with catalytic reformer offgas and with an on-purpose H_2 plant.

The refinery H_2 distribution system supplies multiple hydroprocessing units: hydrocrackers, hydrodesulfurization, and isomerization. H_2-containing streams cascade through various H_2 consumer units, recovery units, purification systems, and sulphur removal steps. The network streams are driven by makeup and recycle compressors and include vents, bypasses and purge gases sending residual H_2 to fuel. Although many of these features are common to every refinery H_2 network, each H_2 infrastructure has evolved into a unique flow scheme.

Operating scenarios within the refinery are dynamic. Each operating case can establish different limitations on various H_2 network optimisation considerations. Therefore, all network stream process conditions are defined for several relevant cases such as:

—Daily operation: Maximum and typical crude rates; Crude slates: sweet/sour; light/heavy

—Catalytic reformer operation swings: Summer v. winter variable octane requirements; Catalytic reformer SOR through EOR catalyst performance; Regeneration schedule

—Hydroprocessing units turnaround schedules

—Future projections for crude flexibility and/or new H_2 consumers.

For each limiting case, typical process stream data is collected – pressure, min/max/average flow, complete composition: H_2, hydrocarbons, H_2S. One day's snapshot usually will not suffice and probably won't balance. Data reconciliation and closing the H_2 balance for each case is usually accomplished by double-checking flow and composition data for key streams during several timeframes, using LP model data, and making some reasonable engineering assumptions.

Many recent technical papers have discussed the systematic network analysis approach using tools such as pinch technology, utilisation analysis, hydroprocessing unit simulation, H_2 separation equipment design tools, and optimiser programmes to identify network improvements. There are many constraints on the optimisation exercise imposed by control philosophy, compression equipment, sulphur removal capabilities, utility infrastructure, and physical location of hydroprocessing units that must be factored into these methods. Therefore, the network analysis also involves a good dose of engineering experience and know-how.

Improvement options identified by the network analysis involve:

—Control and operational changes

—Network re-routing

—Equipment retrofits

—New or expanded H_2 recovery and purification systems such as membrane systems, PSA units, cryogenic systems.

However, for refineries with on-purpose H_2 production plants, the network analysis is not complete without a thorough evaluation of the SMR plant performance.

On-purpose H_2

Two process schemes employed in on-purpose H_2 generation plants are shown

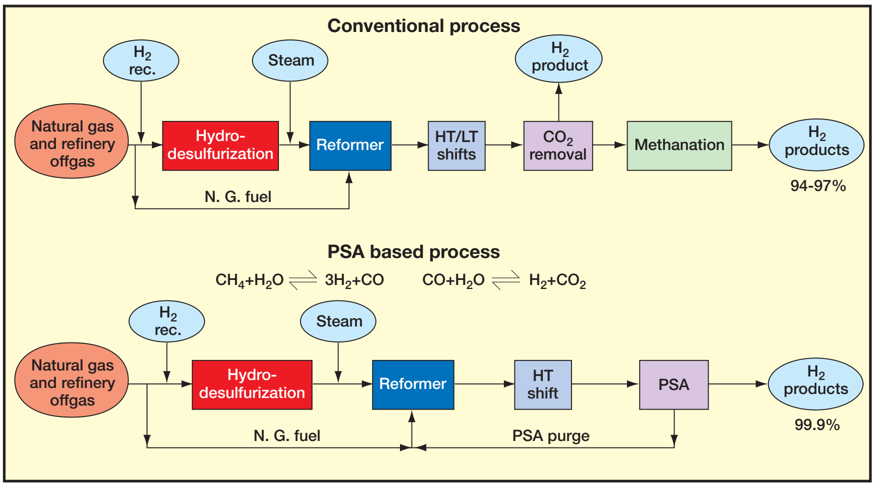

Figure 2 *Hydrogen production process schemes*

in Figure 2. H_2 is produced by steam reforming of light hydrocarbons supplied from refinery offgas feeds and natural gas. The endothermic reforming reaction is accomplished in an SMR by sending pre-heated feed gas mixed with steam through catalyst-filled tubes housed in a furnace. The resulting H_2 and carbon oxides are processed in shift reactors to convert carbon monoxide to CO_2 and produce additional H_2.

The two processes differ mostly in the H_2 purification section downstream of the SMR. Many existing refinery H_2 plants use a conventional process, which produces a medium-purity (94% to 97%) H_2 product by removing the CO_2 in an absorption system and methanating any remaining carbon oxides. Since the 1980s, most H_2 plants are built with pressure swing adsorption (PSA) technology to recover and purify the H_2 to purities above 99.9%. These PSA-based H_2 plants have higher efficiencies than conventional lower-purity plants due to additional export steam credits and more efficient reformer designs.

As new H_2 requirements push existing H_2 plants to maximum production, a comprehensive performance test and subsequent process modelling are required to determine the plant's maximum achievable H_2 capacity, identify the equipment bottlenecks and find inefficiencies in the existing operation. During the test programme, refiners can benchmark their SMR performance with an industrial gas supplier with a track record of operating and expanding many large H_2 plants by using ongoing best practices and incorporating proactive efficiency programmes.

Since many refinery H_2 plants utilise refinery offgas feeds containing H_2, the actual maximum H_2 capacity that can be synthesised via steam reforming is not clear from daily plant data. The H_2 content in offgas feeds can change due to operational changes in the hydrotreaters and/or in the recovery systems. A controlled performance test programme will reveal the maximum synthesis H_2 capacity and define the impact of variable offgas feed H_2 content on overall H_2production. Based on this information, the H_2 management programme can evaluate expansion options, efficiency improvements, reliability enhancements and the plant's optimum contribution in the future H_2 network requirements.

Figure 3 shows some results from a recent refinery H_2 plant performance test. The plant was operated at maximum possible production with various H_2 contents in the feed. The graph shows the overall contained H_2 product (blue line) and the synthesis H_2 from the reforming reaction (red line). Synthesis H_2 capacity is reduced with higher H_2 content feed as this H_2 "takes up" duty in the reformer.

At current operating conditions, the refinery H_2 consumers demand higher synthesis H_2 production than the reformer's current capabilities. In fact, the design synthesis H_2 datapoint shown indicates that the plant cannot operate at its design conditions due to furnace constraints. If future H_2 balance plans had utilised the design synthesis capacity and not accounted for feed H_2 content, a shortfall in H_2 production would have constrained the refinery throughput. During the reformer equipment assessment, a low cost retrofit opportunity was identified to increase synthesis H_2production by about 5% and meet refinery H_2 demands with a small capacity surplus remaining.

The reduction in synthesis H_2 with feed H_2 content also shows the opportunity for recovering offgas H_2 before sending it to the reformer as feedstock. At high recoveries, this would unlock additional H_2 supply availability for the refinery.

H_2 plant expansion

The limitations in an existing SMR H_2 plant are dependent on the plant's age, extent of previous capacity expansion work, utility availability, constraints in emission permits, and the flexibility of the refinery with respect to steam production, design margins, and available downtime for modifications.

Typical process and equipment limitations are found in the radiant and/or convection sections of the reformer furnace, with induced draught (ID) fan capacity, through system pressure drop constraints and with purification equipment capabilities.

Existing H_2 plant expansion from 5% up to 40% can be a cost effective way of fulfilling incremental H_2 needs. Some expansion options available for the SMR H_2 plant are presented in the following discussion and are generally ranked from no or low investment to higher capital requirements (in that order), beginning with reformer process optimisation, PSA unit improvements, reformer re-tubing, pre-reformer investment, upgrading the CO_2 removal system and post-reformer (oxygen secondary) investment. Overcoming

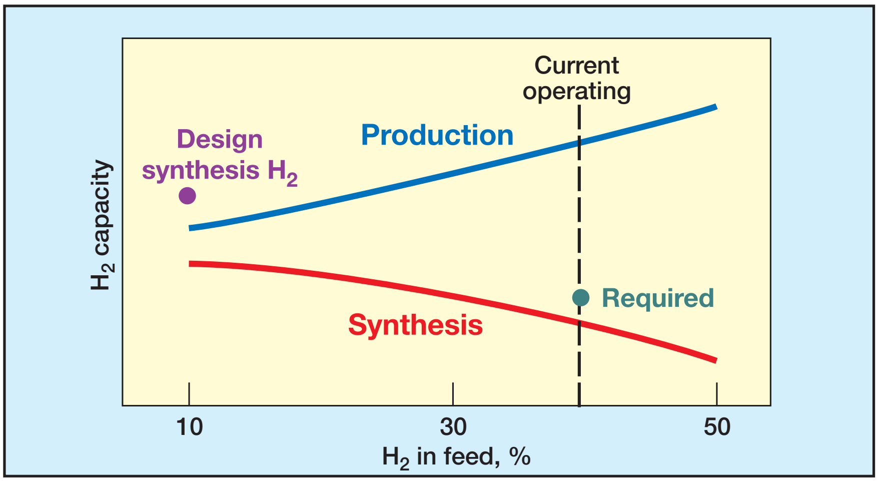

Figure 3 *Hydrogen plant performance results*

reformer firing-limitations is usually the starting point for most expansion option investigations. One reformer process optimisation strategy is to increase the mixed feed preheat temperature from convection section waste heat and, therefore, increase flow through the reformer at a given reformer firing rate. PSA-based plants operating at steam-to-carbon ratios lower than 3.5 can easily increase H_2 production.

Lowering the steam-to-carbon ratio in a conventional (non-PSA based) plant is usually not an option because it will increase methane leakage, which lowers the H_2 purity, and reduces the energy available for stripper reboiler in the CO_2 unit. If reformer firing is limited by ID fan capacity, excess air control and fan retrofits can be examined.

If the reformer uses refinery offgas (ROG) feeds with significant H_2 content, some or all of the ROG feeds can be sent to a new recovery system to reduce the inlet H_2 taking up reformer firing duty. The result is higher synthesis H_2 capability and higher overall H_2 availability from recovery, plus the reformer plant generation.

PSA modifications can be accomplished to increase H_2 production by improving H_2 recovery or increasing its capacity. PSA related options include adsorbent replacement, cycle adjustments, reducing purge gas backpressure, and adding additional PSA vessels.

New catalyst tubes that are made with micro-alloy metallurgy can be installed in the reformer furnace (reformer re-tubing). Larger ID micro-alloy tubes can withstand design pressure differentials at higher skin temperatures while providing excellent tube integrity with 10 to 20 years of operating life. Therefore, the new tubes result in 5% to 10% additional H_2 capacity by increasing reformer throughput and firing.

In a pre-reformer, adiabatic steam-hydrocarbon reforming is performed outside the fired reformer in a vessel containing high nickel catalyst. The heat required for the endothermic reaction is provided by hot flue gas from the reformer convection section. Since the feed to the fired reformer is now partially reformed, the SMR can operate at an increased feed rate and produce 8–10% additional H_2 at the same reformer load. An additional advantage of the pre-reformer is that it facilitates higher mixed feed preheat temperatures and maintains relatively constant operating conditions within the fired reformer regardless of variable refinery offgas feed conditions.

In conventional H_2 plant processes, increased reformer firing and throughput proportionally increases the quantity of CO_2 sent to the absorption system, which is why an upgrade to the CO_2 removal system should be considered. Depending on the system design, CO_2 removal capabilities may need to be expanded by solvent changeout, tower internal replacement or new parallel equipment.

If high purity oxygen is available at the refinery, an oxygen secondary reformer (post reformer) can be installed immediately downstream of the primary reformer furnace to expand overall H_2 production by up to 40%. A portion of the reforming load is shifted from the primary SMR to the secondary reformer. The oxygen secondary reformer provides a low methane slip and the ability to lower the outlet temperature of the primary SMR, providing a means of processing more feed gas in the primary reformer without increasing the reformer-firing rate.

The secondary reformer vessel is a refractory lined carbon steel vessel housing an oxygen burner in its top neck and a fixed catalyst bed. Installation of a secondary reformer usually requires significant changes to the CO_2 removal system and possibly the waste heat boiler. In conventional H_2 plants with methanation back-ends, H_2 product purity can be increased up to 98%. The economics are generally dependent on a reliable source of low cost oxygen and can piggyback off oxygen requirements for sulphur plant and FCC debottlenecking.

Efficiency improvements

Energy consumption for feed and fuel is the largest cost component for H_2 produced from large SMR based H_2 plants. At 50 million standard cubic feet (scfd) H_2 capacity or greater, energy costs account for over 60–70% of the cost. Many of the expansion option strategies previously described will also reduce the energy expenditure, expressed in Btu/scf of H_2 produced. During the H_2 plant evaluation, opportunities for energy efficiency improvements can be identified, which may include:

—Reforming process optimisation such as reduced steam/carbon ratio and new inlet/outlet temperature setpoints
—Furnace optimisation such as excess air control, higher radiant efficiency, and improved waste heat recovery
—CO_2 removal system energy reduction
—H_2 PSA recovery enhancements.

With energy pricing at $4.00–$5.00/million Btu, significant operating costs savings can be achieved by focusing on SMR plant energy efficiency improvements when formulating a H_2 management programme. For a conventional 50 million scfd plant with a current net efficiency of 480Btu/scf H_2, process optimisation improvements with minimum investment can reduce energy consumption by 10–20Btu/scf. This energy reduction translates into $0.70 to $1.75 million saved in energy bills over a year. Nominal capital investment could result in efficiency improvements of up to 30Btu/scf H_2, which provide $2.1 to 2.6 million/year savings.

Replacing aging facilities

Another potential consideration is the replacement of the existing aging H_2 plant because of poor efficiency and high maintenance costs, especially when future H_2 requirements dictate either a major refinery H_2 plant expansion and/or new H_2 supply. If the refiner chooses to continue operating an old SMR plant, capital spending may be required to modernise the plant with new environmental controls (such as NO_X emissions) and reformer tube replacement for a long future run-life.

The difference between the conventional plant design and a new, high efficiency SMR plant at 50 million scfd can be up to 20% higher energy consumption. Using a savings of 70Btu/scf and typical current energy costs at $4.00 to $5.00/million Btu, annual savings for the new design result in as much as $4.5 to $5.5 million/year. As previously described, an expansion project can incorporate efficiency improvements that recuperate a portion of the energy penalty of the older asset. However, the overall H_2 costs after expending capital for a major H_2 plant expansion has to be evaluated against new H_2 supply costs. This is especially important when the expansion does not completely satisfy the full incremental increase in H_2 requirements and new supply is still required. The evaluation must take into account that new supply H_2 becomes less expensive as volumes increase.

Benefits

Finally, the H_2 management programme quantifies the economic benefits realised in the refinery for each recovery, purification, and production improvement option, categorised for both no, or low, capital and higher capital execution plans. A well-designed H_2 management programme has been proved to discover valuable benefits for refinery operation of up to $1 million/year to greater than $10 million/year. Some benefits have immediate impact for current action plans; others require capital investment for implementation to be achieved in future operation.

Potential benefits that can improve refinery current operations with no, or low, capital investment and less than a two-year payback include:

—Decoupling semi-regenerable catalytic reformer operation from H_2 network requirements
—Increasing hydrotreater catalyst life
—Improving hydroprocessing unit product values and reducing existing H_2 plant energy consumption costs
—Maintaining high refinery throughput year-round.

Decoupling catalytic reformer operations from H_2 network requirements should be considered. The catalytic reformer is operated for its primary purpose, which is optimal octane production. This eliminates or at least minimises octane giveaway during winter operation while maximising reformer catalyst cycle length. Conversely, the refinery can maintain high throughput in the hydroprocessing network regardless of a catalytic reformer H_2 supply shortage due to its operating conditions.

If H_2 recovery can meet these objectives, it eliminates the option to make a significant capital investment to convert to continuous catalytic regeneration. Improved H_2 utilisation during the summer octane run can allow for processing less expensive, heavier and more sour crude slates.

Hydrotreater catalyst life is a strong function of H_2 partial pressure. Optimum H_2 purity at the reactor inlet extends catalyst life by maintaining desulphurisation kinetics at lower operating temperatures and reducing carbon laydown. Typical purity increases resulting from H_2 purification equipment and/or increased H_2S removal as well as tuning H_2 circulation and purge rates, may extend catalyst life up to about 25%. The refinery benefits from lower catalyst recharge costs, which can be several million dollars per charge.

Reducing shutdown frequency can also decrease collateral lost production during changeout shutdowns, which add up to significant lost refinery margins. However, improved catalyst life needs to fit into overall refinery shutdown schedules to realise its benefits.

Major refinery margin improvements are available when H_2 systems are optimised in units that are directly or indirectly responsible for gasoline production: high conversion hydrocrackers and cat feed hydrotreaters (CFHT). Higher H_2 partial pressures in the hydrocracker units result in lower operating temperatures and product quality "uplift" to higher gasoline fraction volumes. Increasing hydrocracker makeup H_2 purity by 2–3% can increase C_5+ liquid yields by several hundred bpd. For example, at typical uplift values, a 200bpd C_5+ yield increase translates into about $1 million annual increased margin.

The value of improved CFHT operation is that it indirectly results in

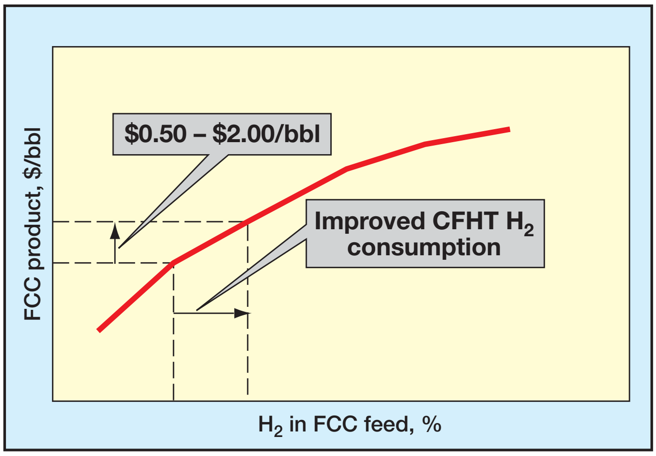

Figure 4 Benefits of CFHT hydrogen system and FCC product value improvements

higher gasoline selectivity from the FCC unit. When optimum H_2 purity and H_2 circulation rates are established in the CFHT, its product H_2 content increases through additional aromatics saturation. Total FCC unit product value can be increased by \$0.5 to \$2/bbl (Figure 4). For many FCC units, this benefit totals over \$10 million/year in increased revenues.

Putting a programme together to improve existing H_2 plant energy consumption costs can save 10 to 30Btu/scf H_2 produced. Efficiency improvements immediately reduce energy bills by several million dollars/year, depending on H_2 plant operating production.

Maintaining high refinery throughput year-round can be achieved by improving existing H_2 plant onstream reliability and debottlenecking its production when it is reaching nameplate capacity.

Future H_2 requirements may require higher capital investment to produce additional H_2 and improve hydrotreater performance and configuration. There are benefits to getting the most from this necessary capital expenditure, including:

—Meeting lower sulphur fuel requirements at minimum cost with hydrotreater unit reconfiguration, which can reduce future H_2 consumption increases

—Defining new H_2 generation requirements with some precision, so future operation is not H_2 constrained.

—Meeting new H_2 demands with existing H_2 plant expansion strategies.

Current H_2 infrastructure

Refinery A has a relatively complex refinery network where cat reformer H_2 offgas is supplemented by on-purpose H_2 from an SMR plant. Purge gases from the hydrocracker and some hydrotreaters are sent to a H_2 recovery system with relatively low recovery at about 70%. The control system also spills about 15% of the catalytic reformer offgas to fuel. Currently, the SMR H_2 plant is not running at nameplate capacity and can handle the relatively low H_2 utilisation of 75% as shown in Table 1.

However, expanding the existing H_2 recovery system and improving its performance can increase utilisation to 83%. This translates into a reduction in the SMR H_2 plant production requirements by about 13 million scfd, saving \$2 million to \$2.5 million/year in energy costs alone.

Future H_2 requirements

Refiner B has a refinery H_2 network with H_2 utilisation already reaching 95%+. This is achieved by strategic installation of H_2 recovery systems combined with using refinery offgas as H_2 plant feed. Mostly low pressure purges and vents are sent to fuel with little opportunity for additional H_2 recovery. The challenge is to find the best combination of expansion, recovery, and supply to meet future ULSD requirements that require 30 million SCFD additional H_2. The options include:

—The existing H_2 plant has 15 million scfd additional H_2 capacity and runs at 480Btu/scf net efficiency; process optimisation could reduce energy consumption modestly to 470Btu/scf

—The H_2 plant can be expanded to produce 4 million scfd additional H_2 with low capital investment; a higher investment project is estimated to increase H_2 capacity by 10 million scfd, but neither expansion alone will provide the entire 30 million scfd requirement

—A low cost equipment modification would recover an additional 2 million scfd H_2

—The remaining H_2 requirement will be supplied by a new industrial gas company H_2 plant dedicated to this refinery.

Refinery A: hydrogen utilisation

		Current operation H_2 million scfd	H_2 recovery revamp H_2 million scfd
H_2 Production			
H_2 plant		Base	Base–13.3
Cat reformers		Max summer	Max Summer
	Total	138.0	124.7
H_2 demand			
Hydrotreaters		47.0	47.0
Hydrocracker		50.0	50.0
Isomerisation		6.0	6.0
	Total	103.0	103.0
H_2 to recovery			
Purges to recovery		26.0	36.0
H_2 recovered		18.0	31.3
Offgas to fuel		8.0	4.7
H_2 to fuel			
HT purges to fuel		17.0	17.0
Cat reformer spill		10.0	0.0
H_2 recovery offgas		8.0	4.7
	Total	35.0	21.7
H_2 utilisation		**74.6%**	**82.6%**
H_2 plant savings			
H_2 production decrease, million scfd			13.3
Plant cost savings @ \$5.00 million Btu \$million/yr			2.5
Plant cost savings @ \$4.00/million Btu\$million/yr			2.0

Table 1

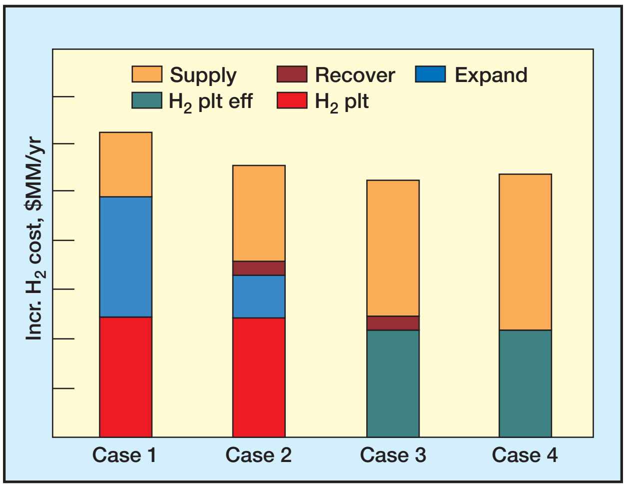

Figure 5 *Refinery B: future hydrogen requirements; option comparison*

Figure 5 compares several combinations of the previously discussed options to determine the lowest incremental H_2 cost for future requirements:

Case 1: First reactions could be to maximise expansion and minimise new H_2 supply. Spend the capital to expand the H_2 plant by 10 million scfd and run it full out to provide 25 million scfd additional H_2. Buy 5 million scfd H_2 from an "over-the-fence" supplier. This results in the highest incremental H_2

cost of almost \$3.00/1000scf.

Case 2: Opt for the lower capital cost H_2 plant expansion and ramp up the H_2 plant to 19 million scfd. Add-in the recovery project, which only provides 2 million scfd more H_2, but at the lowest cost of all options. Buy 9 million scfd H_2 from a supplier. There is a significant reduction in incremental H_2 cost to about \$2.60/1000scf, which is a savings of over \$3 million/ year.

Case 3: Only execute the small recovery project and buy 13 million scfd H_2 supply. Run the H_2 plant at nameplate capacity, but work with a H_2 plant service provider to install an ongoing efficiency programme. Even if a 10Btu/scf improvement can be sustained, incremental H_2 cost is reduced to \$2.50/1000scf, another \$1 million/year improvement.

Case 4: Note that with no capital project execution within the refinery, an efficiency improvement programme can be implemented on the existing H_2 plant and buy 15 million scfd H_2. Incremental H_2 costs are only slightly higher at \$2.55/1000scf.

Case 5: Consider shutdown economics. Replacing the existing H_2 plant with a new, higher efficiency large H_2 plant for the entire H_2 supply could reduce overall H_2 costs to \$2.35–\$2.45/1000scf.

For the incremental approach, strategies 3 or 4 are looking like the best choices, depending on whether the refinery decides to execute a capital project as part of the programme. Ultimately, understanding the bottlenecks, expansion capabilities and efficiency improvement opportunities, and shutdown economics considerations in the existing H_2 plant is key to determining the best overall solution for future H_2 requirements.

Conclusion

In the clean fuels environment, hydroprocessing capacity limits refinery throughput and operating margins. Optimised H_2 infrastructure must be a high priority and depends on a proactive H_2 management programme. At first, the programme establishes maximum H_2 utilisation by implementing cost-effective recovery and reconfiguring networks to take advantage of higher quality H_2 benefits.

Today's clean fuels refinery is more dependent on on-purpose H_2 production. Existing H_2 generation plants must be evaluated to establish their current maximum capacity, identify cost-effective expansions, and set up energy efficiency and on-stream reliability programmes. For future requirements, the best strategies for low cost H_2 require the right combination of expansion, efficiency improvements, recovery and new supply. Finally, increased refinery margins and cost savings identified from a well-designed H_2 management plan need to be maintained at optimal levels through an ongoing H_2 monitoring programme.

Ruth A Davis *is manager of hydrogen services and solutions technology for Air Products and Chemicals Inc, Allentown, Pennsylvania, USA. She has 25 years of process engineering experience, providing process management for H_2 production and contributing to cryogenics, adsorption and membrane separation system applications, and is responsible for technology strategy supporting the global hydrogen services business. She holds an MS degree in chemical engineering from University of Delaware.*

E-mail: davisra2@airproducts.com

Nitin M Patel *is senior engineering associate, H_2/syngas technology, at Air Products & Chemicals. He has over 27 years of process engineering experience in hydrogen production, including technical leadership in all aspects of H_2/syngas technology, and is responsible for technology strategy formulation in support of the company's global hydrogen production services. He holds an MS degree in chemical engineering from Georgia Institute of Technology.*