Hydrogen management has become a priority in current refinery operations and when planning to produce lower sulphur gasoline and diesel fuels. Along with increased H2 consumption for deeper hydrotreating, additional H2 is needed for processing heavier and higher sulphur crude states. In many refineries, hydprocessing capacity and the associated H2 network is limiting refinery throughput and operating margins. Furthermore, higher H2 purities within the refinery network are becoming more important to boost hydrotreater capacity, achieve product value improvements and lengthen catalyst life cycles.

Improved H2 utilisation and expanded or new sources for refinery H2 and H2 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 H2 management programmes fit into the two general categories of either a catalytic reformer supplied network or an on-purpose H2 supply.

Some refineries depend solely on catalytic reformer(s) as their source of H2 for hydrotreating. Often, they are semi-regenerative reformers where offgas H2 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 H2 shortages.

Multiple hydrotreating units compete for H2 – either by selectively reducing throughput, managing intermediate tankage logistics, or running the catalytic reformer sub-optimally just to satisfy downstream H2 requirements.

Part of the operating year still runs in H2 surplus, and the network may be operated with relatively low H2 utilisation (consumption/production) at 70 to 80%. Catalytic reformer offgas H2 supply may swing from 75 to 85% H2 purity. An H2 purity upgrade can be achieved through some hydrotreaters by absorbing heavy hydrocarbons. But without supplemental H2 purification, critical control of H2 partial pressure in hydroprocessing reactors is difficult, which can affect catalyst life, charge rates, and/or gasoline yields.

More complex refineries, especially those with hydrocracking units, also have on-purpose H2 production, typically with a steam methane reformer (SMR) that utilises refinery offgases and supplemental natural gas as feedstock. The SMR plant provides the swing H2 requirements at higher purities (92-99% H2) and serves an H2 network configured with several purity and pressure levels. Multiple purities and existing purification units allow for more optimised hydroprocessing operation by controlling H2 partial pressure for maximum benefit. Typical H2 utilisation is 85% to 95%. When the refinery SMR plant begins reaching its nameplate H2 capacity during catalytic reformer H2 production swings, hydprocessing 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 H2 management programme should be organised to meet some of the following objectives, depending on specific refinery configuration:

- Maximise H2 utilisation through increased recovery;
- Decouple catalytic reformer operation with H2 production needs;
- Take advantage of higher H2 purity within specific consumers;
- Reduce costs of on-purpose H2 production;
- Improve on-purpose H2 production reliability;
Expand on-purpose H\textsubscript{2} production;
Evaluate refinery H\textsubscript{2} plant shutdown economics;
Integrate with new industrial gas company H\textsubscript{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\textsubscript{2} balance information for each limiting case. Figure 1 (previous page) shows a conceptual H\textsubscript{2} network configuration for the refinery sourced with catalytic reformer offgas and with an on-purpose H\textsubscript{2} plant.

The refinery H\textsubscript{2} distribution system supplies multiple hydrotreating units: hydrocrackers, hydrodesulfurization, and isomerization. H\textsubscript{2}-containing streams cascade through various H\textsubscript{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\textsubscript{2} to fuel. Although many of these features are common to every refinery H\textsubscript{2} network, each H\textsubscript{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\textsubscript{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\textsubscript{2} consumers.

For each limiting case, typical process stream data is collected – pressure, min/max/average flow, complete composition: H\textsubscript{2}, hydrocarbons, H\textsubscript{2}S. One day's snapshot usually will not suffice and probably won't balance. Data reconciliation and closing the H\textsubscript{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, hydrotreating unit simulation, H\textsubscript{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 hydrotreating 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\textsubscript{2} recovery and purification systems such as membrane systems, PSA units, cryogenic systems.

However, for refineries with on-purpose H\textsubscript{2} production plants, the network analysis is not complete without a thorough evaluation of the SMR plant performance.

**On-purpose H\textsubscript{2}**

Two process schemes employed in on-purpose H\textsubscript{2} generation plants are shown...
in Figure 2, H₂ is produced by steam reforming of light hydrocarbons supplied from refinery offgas 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 H₂ and carbon oxides are processed in shift reactors to convert carbon monoxide to CO₂ and produce additional H₂.

The two processes differ mostly in the H₂ purification section downstream of the SMR. Many existing refinery H₂ plants use a conventional process, which produces a medium-purity (94% to 97%) H₂ product by removing the CO₂ in an absorption system and methanating any remaining carbon oxides. Since the 1980s, most H₂ plants are built with pressure swing adsorption (PSA) technology to recover and purify the H₂ to purities above 99.9%. These PSA-based H₂ plants have higher efficiencies than conventional lower-purity plants due to additional export steam credits and more efficient reformer designs.

As new H₂ requirements push existing H₂ plants to maximum production, a comprehensive performance test and subsequent process modelling are required to determine the plant’s maximum achievable H₂ 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 H₂ plants by using ongoing best practices and incorporating proactive efficiency programmes.

Since many refinery H₂ plants utilise refinery offgas feeds containing H₂, the actual maximum H₂ capacity that can be synthesised via steam reforming is not clear from daily plant data. The H₂ 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 H₂ capacity and define the impact of variable offgas feed H₂ content on overall H₂ production. Based on this information, the H₂ management programme can evaluate expansion options, efficiency improvements, reliability enhancements and the plant’s optimum contribution in the future H₂ network requirements.

Figure 3 shows some results from a recent refinery H₂ plant performance test. The plant was operated at maximum possible production with various H₂ contents in the feed. The graph shows the overall contained H₂ product (blue line) and the synthesis H₂ from the reforming reaction (red line). Synthesis H₂ capacity is reduced with higher H₂ content feed as this H₂ “takes up” duty in the reformer.

At current operating conditions, the refinery H₂ consumers demand higher synthesis H₂ production than the reformer’s current capabilities. In fact, the design synthesis H₂ datapoint shown indicates that the plant cannot operate at its design conditions due to furnace constraints. If future H₂ balance plans had utilised the design synthesis capacity and not accounted for feed H₂ content, a shortfall in H₂ production would have constrained the refinery throughput. During the reformer equipment assessment, a low cost retrofit opportunity was identified to increase synthesis H₂ production by about 5% and meet refinery H₂ demands with a small capacity surplus remaining.

The reduction in synthesis H₂ with feed H₂ content also shows the opportunity for recovering offgas H₂ before sending it to the reformer as feedstock. At high recoveries, this would unlock additional H₂ supply availability for the refinery.

**H₂ plant expansion**

The limitations in an existing SMR H₂ 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 H₂ plant expansion from 5% up to 40% can be a cost effective way of fulfilling incremental H₂ needs. Some expansion options available for the SMR H₂ 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 CO₂ removal system and post-reformer (oxygen secondary) investment. Overcoming
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 H₂ 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 H₂ purity, and reduces the energy available for stripper reboiler in the CO₂ 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 H₂ content, some or all of the ROG feeds can be sent to a new recovery system to reduce the inlet H₂ taking up reformer firing duty. The result is higher synthesis H₂ capability and higher overall H₂ availability from recovery, plus the reformer plant generation.

PSA modifications can be accomplished to increase H₂ production by improving H₂ 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 H₂ 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 H₂ 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 H₂ plant processes, increased reformer firing and throughput proportionally increases the quantity of CO₂ sent to the absorption system, which is why an upgrade to the CO₂ removal system should be considered. Depending on the system design, CO₂ 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 H₂ 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 CO₂ removal system and possibly the waste heat boiler. In
conventional H₂ plants with methanation
back-ends, H₂ plant 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 H₂ pro-
duced from large SMR based H₂ plants.
At 50 million standard cubic feet (scfd)
H₂ 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 H₂ produced. During the H₂ plant
evaluation, opportunities for energy
efficiency improvements can be identi-
fied, 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
— CO₂ removal system energy reduction
— H₂ PSA recovery enhancements.

With energy pricing at $4.00–
$5.00/million Btu, significant operating
costs savings can be achieved by focus-
ing on SMR plant energy efficiency
improvements when formulating a H₂
management programme. For a conven-
tional plant, capital spending may be
required to modernise the plant with a cur-
rent net efficiency of 480Btu/scf H₂,
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 invest-
ment could result in efficiency improve-
ments 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 catalyt-
ic reformer operation from H₂ network
requirements
— Increasing hydrotreater catalyst life
— Improving hydroprocessing unit
product values and reducing existing H₂
plant energy consumption costs
— Maintaining high refinery through-
put year-round.

Decoupling catalytic reformer opera-
tions from H₂ network requirements
should be considered. The catalytic
reformer is operated for its primary pur-
pose, which is optimal octane produc-
tion. This eliminates or at least minimises
cycle length. Conversely, the
refinery can maintain high throughput in
the hydroprocessing network regard-
less of a catalytic reformer H₂ sup-
ply shortage due to its operating condi-
tions.

If H₂ recovery can meet these objec-
tives, it eliminates the option to make a
significant capital investment to convert
to continuous catalytic reforming (CFHT).
Improved H₂ utilisation during the sum-
mer octane run can allow for processing
less expensive, heavier and more sour
crude slates.

Hydrotreater catalyst life is a strong
function of H₂ partial pressure. Opti-
mum H₂ purity at the reactor inlet
extends catalyst life by maintaining
desulphurisation kinetics at lower oper-
ating temperatures and reducing carbon
delay. Typical purity increases result-

ing from H₂ purification equipment
and/or increased H₂S removal as well as
tuning H₂ 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 chaotic cut shutdowns, which
add up to significant lost refinery mar-
gins. However, improved catalyst life
needs to fit into overall refinery shut-
down schedules to realise its benefits.

Major refinery margin improvements
are available when H₂ systems are opti-
mised in units that are directly or indi-
rectly responsible for gasoline produc-
tion: high conversion hydroccra-
ters (HCC), cat and thermal hydrocen-
ters (CFHT). Higher H₂ partial pressures
in the hydrocracker units result in lower
operating temperatures and product
catalyst “uplift” to higher gasoline frac-
tion volumes. Increasing hydrocracker
makeup H₂ purity by 2–3% can increase
$0.50 to $2.00/bbl. Fuel production.

Figure 4 Benefits of CfHT hydrogen system and FCC product

tional purity improves

$0.50 – $2.00/bbl

H₂ in FCC feed, %

Improved CFHT H₂ consumption

Figure 4 Benefits of CfHT hydrogen system and FCC product value improvements
higher gasoline selectivity from the FCC unit. When optimum H₂ purity and H₂ circulation rates are established in the CFHT, its product H₂ 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 H₂ plant energy consumption costs can save 10 to 30Btu/scf H₂ produced. Efficiency improvements immediately reduce energy bills by several million dollars/year, depending on H₂ plant operating production.

Maintaining high refinery throughput year-round can be achieved by improving existing H₂ plant onstream reliability and debottlenecking its production when it is reaching nameplate capacity.

Future H₂ requirements may require higher capital investment to produce additional H₂ 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 H₂ consumption increases.
— Defining new H₂ generation requirements with some precision, so future operation is not H₂ constrained.
— Meeting new H₂ demands with existing H₂ plant expansion strategies.

**Current H₂ infrastructure**

Refinery A has a relatively complex refinery network where cat reformer H₂ offgas is supplemented by on-purpose H₂ from an SMR plant. Purge gases from the hydrocracker and some hydro-treaters are sent to a H₂ recovery system with relatively low recovery at about 70%. The control system also spills about 15% of the catalytic reformer off-gas to fuel. Currently, the SMR H₂ plant is not running at nameplate capacity and can handle the relatively low H₂ utilisation of 75% as shown in Table 1.

However, expanding the existing H₂ recovery system and improving its performance can increase utilisation to 83%. This translates into a reduction in the SMR H₂ plant production requirements by about 13 million scfd, saving $2 million to $2.5 million in energy costs alone.

**Future H₂ requirements**

Refiner B has a refinery H₂ network with H₂ utilisation already reaching 95%. This is achieved by strategic installation of H₂ recovery systems combined with using refinery offgas as H₂ plant feed. Mostly low pressure purges and vents are sent to fuel with little opportunity for additional H₂ 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 H₂. The options include:

— The existing H₂ plant has 15 million scfd additional H₂ capacity and runs at 480Btu/scf net efficiency; process optimisation could reduce energy consumption modestly to 470Btu/scf.
— The H₂ plant can be expanded to produce 4 million scfd additional H₂ with low capital investment; a higher investment project is estimated to increase H₂ 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 H₂.
— The remaining H₂ requirement will be supplied by a new industrial gas company H₂ plant dedicated to this refinery.

**Figure 5** compares several combinations of the previously discussed options to determine the lowest incremental H₂ cost for future requirements:

**Case 1:** First reactions could be to maximise expansion and minimise new H₂ supply. Spend the capital to expand the H₂ plant by 10 million scfd and run it full out to provide 25 million scfd additional H₂. Buy 5 million scfd H₂ from an “over-the-fence” supplier. This results in the highest incremental H₂.
cost of almost $3.00/1000scf.

**Case 2:** Opt for the lower capital cost H₂ plant expansion and ramp up the H₂ plant to 19 million scfd. Add-in the recovery project, which only provides 2 million scfd more H₂, but at the lowest cost of all options. Buy 9 million scfd H₂ from a supplier. There is a significant reduction in incremental H₂ 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 H₂ supply. Run the H₂ plant at nameplate capacity, but work with a H₂ plant service provider to install an ongoing efficiency programme. Even if a 10Btu/scf improvement can be sustained, incremental H₂ 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 H₂ plant and buy 15 million scfd H₂. Incremental H₂ costs are only slightly higher at $2.55/1000scf.

**Case 5:** Consider shutdown economics. Replacing the existing H₂ plant with a new, higher efficiency large H₂ plant for the entire H₂ supply could reduce overall H₂ 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 H₂ plant is key to determining the best overall solution for future H₂ requirements.

**Conclusion**

In the clean fuels environment, hydroprocessing capacity limits refinery throughput and operating margins. Optimised H₂ infrastructure must be a high priority and depends on a proactive H₂ management programme. At first, the programme establishes maximum H₂ utilisation by implementing cost-effective recovery and reconfiguring networks to take advantage of higher quality H₂ benefits.

Today’s clean fuels refinery is more dependent on on-purpose H₂ production. Existing H₂ 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 H₂ require the right combination of expansion, efficiency improvements, recovery and new supply. Finally, increased refinery margins and cost savings identified from a well-designed H₂ management plan need to be maintained at optimal levels through an ongoing H₂ monitoring programme.

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