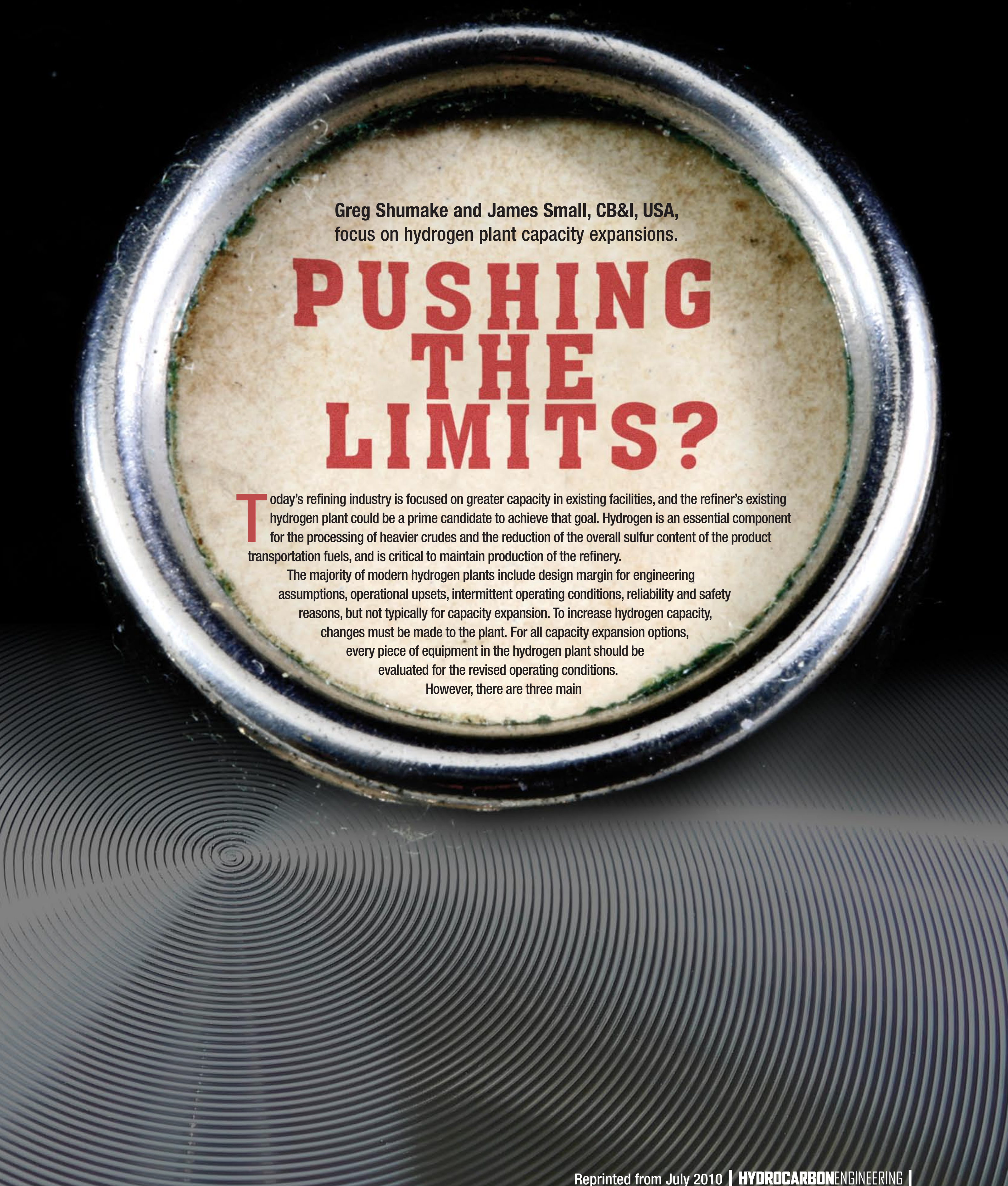

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focus on hydrogen plant capacity expansions.

PUSHING THE LIMITS?

Today's refining industry is focused on greater capacity in existing facilities, and the refiner's existing hydrogen plant could be a prime candidate to achieve that goal. Hydrogen is an essential component for the processing of heavier crudes and the reduction of the overall sulfur content of the product transportation fuels, and is critical to maintain production of the refinery.

The majority of modern hydrogen plants include design margin for engineering assumptions, operational upsets, intermittent operating conditions, reliability and safety reasons, but not typically for capacity expansion. To increase hydrogen capacity, changes must be made to the plant. For all capacity expansion options, every piece of equipment in the hydrogen plant should be evaluated for the revised operating conditions. However, there are three main areas of focus for debottlenecking a hydrogen plant: reforming and shift conversion, product purification, and process gas cooling.

Reforming and shift conversion

The reformer is the most important piece of equipment in the hydrogen plant. The primary design margin for the reformer is the temperature margin between the design temperature of the reformer tubes and the maximum calculated tube wall temperature. A reformer can sometimes be pushed with additional process gas throughput and fuel usage, but attention must be focused on the temperature of the reformer tubes. The life of the reformer tubes can be substantially reduced if the tube wall temperature exceeds the design temperature (Table 1). For this reason, tube wall temperature surveys of the reformer are recommended on a daily basis. Typically, a hydrogen plant can yield an approximate increase in capacity up to 10% with this approach. A refiner must decide if a loss in tube life is permissible in order to reach the additional capacity. Note that in some instances tubes are designed with a minimum sound wall thickness. This means the tube may be designed for a longer tube life than indicated. Depending on the design conditions, the required tube wall thickness may be thinner than the minimum sound wall thickness. This will result in a larger design margin at the standard tube life or a longer tube life at the standard design margin. This condition will allow for additional throughput while maintaining the standard design margin and tube life.

The Larson-Miller (Equation 1) is used to design the reformer tubes, based on a minimum required tube life. Table 1 illustrates the effect of temperature on the life of the reformer tubes.

$$\text{Log}_{10}S = -0.0062{*}P^2 + 0.2955{*}P - 1.5426 \qquad (1)$$

$$P = T(22.96 + \text{Log}_{10}t) * (10^{-3})$$

Where:

S - stress (Mpa).

P - Larson-Miller parameter.

T - tube wall temperature (°K).

t - time (hrs).

Figure 1. Existing two cylindrical reformer hydrogen unit (left) revamped by the addition of a third cylindrical reformer (right).

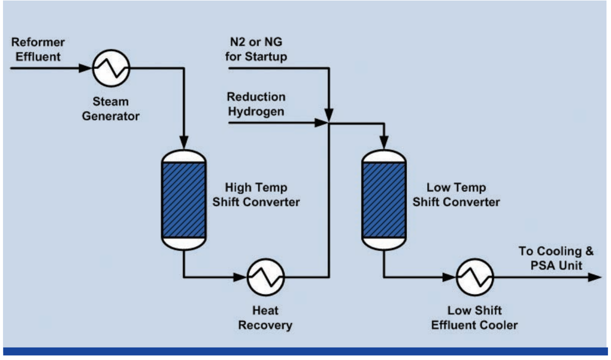

Figure 2. Low temperature shift converter layout.

The results in Table 1 are based on a tube wall temperature margin of 50 °F for 100 000 hrs tube life, which are industry typical values. Varying the tube wall temperature by +/-25 °F illustrates the effect that temperature has on the tube life.

On lower capacity hydrogen plants (<10 million ft^3/d H_2/ 11 166 NCMH) that use smaller cylindrical reformers, it may be possible to add an additional reformer to increase the overall plant capacity. A North American refiner, working with CB&I, was able to increase its plant capacity from 6 million ft^3/d H_2 (6699 NCMH) to approximately 9 million ft^3/d H_2 (10 048 NCMH) by the addition of a third cylindrical reformer (Figure 1).

Another method to debottleneck the reforming capacity of the hydrogen plant is to increase the feed preheat temperature of the feed gas and steam mixture to the reformer. This decreases the absorbed duty of the reformer, thereby allowing additional capacity for firing. Many plants are designed to preheat the reformer feed to approximately 950 °F (510 °C) in the waste heat recovery unit. Increasing this temperature to 1050 °F (566 °C) can yield an approximate increase in capacity up to 7%, assuming there are no pressure drop limitations through the unit. Equipment changes for this approach would include a new feed preheat coil and possible changes to the inlet piping system to the reformer. This change will also decrease steam production because of the shift in duty to feed preheating.

A hydrogen plant uses the water gas shift reaction in Equation 2 to increase the hydrogen yield of the unit.

$$CO + H_2O = H_2 + CO_2 \qquad (2)$$

Most hydrogen plants employ a high temperature shift converter downstream of the primary reformer. In these instances, the addition of a low temperature shift converter (Figure 2) can yield an approximate increase in capacity of up to 5% by shifting most of the residual CO to hydrogen. Equipment modifications to the hydrogen plant would include the additional shift converter vessel and an effluent exchanger. The additional equipment would increase the overall pressure drop of the unit and would ultimately reduce the final hydrogen product pressure. One thing to keep in mind with the low temperature shift converter is that special attention must be given to its startup. The low temperature shift

Table 1. Tube wall temperature effect

Design tube wall	
Temperature (°F)	Tube life (hrs)
1675	47 500
1650	100 000
1625	216 500

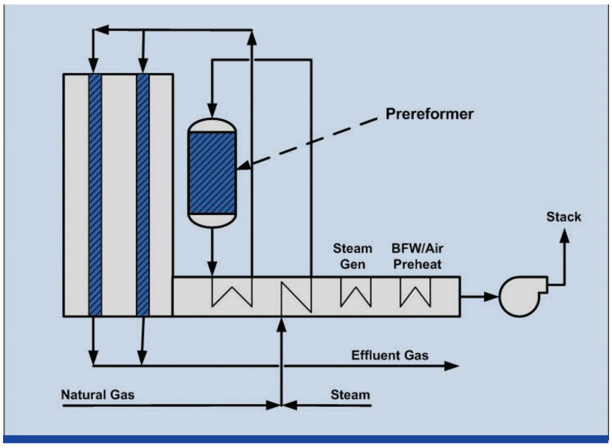

Figure 3. Preformer layout.

Figure 4. Navajo Refining's cylindrical unit (left) utilised secondary feeds to the PSA. Excess offgas was exported and utilised as fuel for the box reformer (centre).

converter will have a startup loop installed that will use either natural gas or nitrogen for catalyst reduction on the initial run and subsequent catalyst changeouts. The low temperature shift catalyst is poisoned easily with chlorides or sulfur compounds in comparison to the high temperature shift catalyst.

The addition of a prereformer to an existing hydrogen plant can also increase the capacity of the plant. The prereformer can alleviate a portion of the absorbed duty from the primary reformer since it partially reforms the feedstock stream. However, since the reforming capacity is all in the same heat envelope, additional capacity is gained only by reheating the prereformer effluent. This creates additional firing capacity in the primary reformer. In Figure 3, a prereformer effluent convective coil has been inserted into the waste heat recovery unit to allow for an increase in plant capacity. The addition of a prereformer has a twofold effect:

- Reduces the heat load on the primary reformer.
- Reduces the steam make of the plant.

The amount of heat load reduction is dependent upon the types of feedstock. For a plant using a natural gas feedstock, the use of a prereformer could reduce the absorbed duty of the primary reformer by approximately 6%, whereas for a naphtha fed plant, the absorbed duty could be reduced by approximately 11%. This reduction in absorbed duty then allows for increased capacity of the plant.

Steam production is reduced since a portion of the waste heat duty is shifted from steam generation to the reheating of the prereformer effluent. The amount of steam reduction is also dependent on the types of feedstock used in the plant. For a natural gas fed plant, the prereformer is typically endothermic. For a heavy hydrocarbon feed such as naphtha, the prereformer is typically exothermic.

Product purification

In modern hydrogen units, the final product purification is accomplished by the use of a pressure swing adsorption unit (PSA). The PSA is a multi bed system that operates on a repeating cycle having two basic steps: adsorption and regeneration. The product hydrogen flows through the PSA while the impurities are adsorbed and then released as offgas fuel back to the reformer. There is typically little design margin in the design of a PSA unit. Therefore, the PSA is a vital piece of equipment that will need to be evaluated for capacity expansion options.

There are several methods for increasing the capacity of the PSA unit. There may be updates to the PSA operating programme that could improve the operation of the unit. The adsorbent may be changed out for improved adsorbent materials. In addition, the mixture of the adsorbent materials can be modified. These changes can typically yield an approximate increase in capacity of up to 2%. If additional PSA capacity is required, more extensive modifications can sometimes be made to the unit. Most PSA units operate in pairs of adsorber vessels. Therefore, additional adsorber vessels could be installed to increase the capacity of the unit. This option would include an overall evaluation of the accompanying PSA valve skid and pipe headers. Some of these components may also require modification for this option.

In most refineries, there are various offgas streams from other refining units that are rich in hydrogen such as a hydrotreater offgas or perhaps a hydrocracker purge stream. If the PSA unit has to be revamped, then it may be possible to design the modified PSA unit to purify the alternate streams to increase the overall hydrogen production of the unit. However, there are limits to using these alternate streams to the PSA. The use of an alternate feed stream to the PSA will increase the offgas that is returned to the reforming furnace where it provides the majority of the heat input. In addition, the heating value of the offgas is impacted by the alternate streams and the reformer burner design would have to be evaluated for these changes. The additional offgas from the PSA combined with the potentially higher heating value could result in more firing than the primary reformer requires. This could limit the flow of alternate streams being sent to the modified PSA unit, unless there are options for exporting the excess low pressure offgas. Navajo Refining added this capability to one of its new CB&I designed hydrogen plants in 2005 (Figure 4). The alternate streams yielded 4 million ft^3/d (4465 NCMH) of additional hydrogen.

Process gas cooling

The PSA has an optimum temperature operating range of 90 - 120 ˚F (32 - 49 ˚C). The performance of the PSA unit will decrease at higher inlet temperatures. For capacity expansion options, the final synthesis gas cooling prior to entering the PSA is typically a bottleneck. The final synthesis gas cooling usually consists of an air cooler followed by a cooling water trim cooler. Therefore, both exchangers would have to be evaluated at the higher throughput rates. There may be options to increase the performance of the coolers without the need for replacement. For process gas air coolers, there are ways to maximise the air flow through the exchanger. In some cases, it may be possible to increase the fan blade pitch, thus increasing the amount of air flow the fan is able to move. If the fan blade pitch is already maximised, then other scenarios could include purchasing a larger motor and possibly a different set of sheaves. For process gas coolers using cooling water, there are a couple of areas to maximise the use of the cooling water. In some cases, the cooling water flow may be regulated by the use of restriction orifices. The removal of these flow orifices would increase the flow through the exchanger. Another method to increase the performance of the exchanger is to chill the cooling water prior to entering the exchanger.

Conclusion

Today's demand to meet tighter emission requirements for transportation fuels will require additional hydrogen. In some cases incremental hydrogen demand can be met through the revamp of an existing hydrogen plant. The options presented above provide several solutions to refiners for increasing the hydrogen plant capacity. These options can be done alone or in combination to further increase the capacity, but at a capital cost and probable reduction in product pressure. For a successful hydrogen plant expansion, a thorough evaluation should be performed not only for each piece of equipment but also plant utility usage and emission impacts.

References

1. SHUMAKE, G. and SMALL, J., 'Mixing Things Up in Hydrogen Plants', *Hydrocarbon Engineering*, November 2006, p 43 - 48.
2. SHUMAKE, G. and ABU DIAB, T., 'Operation Hydrogen', *Hydrocarbon Engineering*, February 2006, p 43 - 48.
3. HSU, C.S. and ROBINSON, P.R., editors; Practical Advances in *Hydrocarbon Processing*. Springer: New York, 2005, p. 323 - 369.
4. BOYCE, C.; RITTER, R.; CREWS, M.A.; 'Is It Time For A New Hydrogen Plant', *Hydrocarbon Engineering*, February 2004.