# HDS benefits from plate heat exchangers

**Case studies illustrate benefits of plate heat exchangers versus shell and tube heat exchangers in hydrodesulphurisation units. Direct benefits include significant reductions in the operating duty of fired heaters and product coolers**

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In modern hydrodesulphurisation (HDS) units designed for producing ultra low sulphur diesel (ULSD), a growing number of refiners have selected plate heat exchangers (PHE) to minimise overall HDS capital and energy costs. Compared with traditional shell and tube (S&T) heat exchanger designs, some real-project case studies indicate that a modern 100 000bpsd HDS with PHEs will save the refiner some US$25–40 million in total costs during the project's first five years. Direct and indirect benefits of PHE selection are found as follows.

In HDS reactor feed/effluent (F/E) service, replacing a string of high-pressure S&T exchangers with a single, more efficient PHE can more than halve the operating duty of fired heaters and product coolers, saving both direct fuel cost and capital cost. Further, the simplified mechanical design of the CFE plus the lower Delta P of a smaller heater result in significantly lower pressure drop in the process loop, with power consumption at the recycle gas (RG) compressor cut by 20% to 40%.

Unit down-time costs for cleaning are also minimised, as uniform flow turbulence makes PHE exchangers in HDS service relatively slow to foul by salt or gum. When fouling does occur, as-new performance can be soon recovered by simple in-situ water-wash and coke-burn procedures.

Costly contamination of ULSD via cross-channel HE-leakage of high-sulphur feed becomes a virtual non-event with any new PHE. These tough, stainless steel (SS) exchangers are purpose-built and factory-tested to guarantee that cross leakage (if any) is below 5 parts in 10 million (less than 0.01ppm sulphur contamination on effluent assuming 20 000ppm sulphur feed).

The technology also spins off some less quantified but very real safety and environmental gains. These are found in a neater, more-compact HDS plant with less complex piping, lower maintenance, fewer big HP flanges to leak fugitive emissions and lower stack-discharge of GHG.

## Mechanical design

The optimised heat integration of a modern HDS unit with PHEs requires two exchangers, one in HP reactor F/E service, the other in LP stripper bottoms service.

Figure 1 shows the HP Packinox reactor F/E HE as a one-pass, true counter-flow plate pack inside a pressure vessel. The pack consists of thin stainless steel corrugated sheets formed by underwater explosion, stacked and welded together. With such plate packs, very large total heat-transfer surface areas are contained inside a relatively compact shell volume cf S&T exchangers. These PHEs have been built with areas over $15\,000m^2$ and duties over 100MW in a single shell.

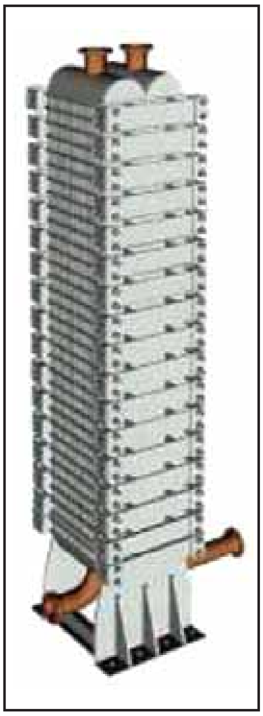

*Figure 1 (left) Reactor feed/effluent HE*
*Figure 2 (right) HDS stripper bottom HE*

As the PHE has no gaskets to soften and leak, it operates comfortably up to 550°C (1000°F). Maximum operating pressure is determined only by the design of the surrounding pressure vessel, and not by the design of the internal plate-pack. The vessel is simply pressurised with hydrogen from the discharge of the HDS recycle gas compressor. This point is always at the highest pressure in the reactor circuit and so the plate pack is always under external compression from a positive DP.

Thus, the plate bundle need only be designed for a differential pressure (design DP up to 30 bar approximately).

All Packinox reactor F/E exchangers for HDS units have thick restraining outer plates with tie rods as a precaution against any operational upset causing a "reverse DP" up to 2 bar. This device, plus some other simple passive safety measures built into the unit PID during Hazop review (check-valve at critical location, nitrogen purge location etc) give solid assurance against dangers of everse DP in operation and during turnaround.

Four bellows compensate for differential thermal expansion between the hot stainless steel pack and the relatively cool low alloy pressure vessel. Top and bottom end manholes are provided to facilitate bellows replacement in the unlikely event that this should ever be needed. Note that these bellows are in non-critical service, as their failure would only affect RG distribution.

ULSD quality is protected by ultra-tough PHE exchanger construction for HDS service. This construction employs thicker SS plates (between 1.5 and 2 times the thickness of the plates in Packinox's original PHEs for reformer service), heavier welding (principally at plate ends) and stiffer

HDS plate-corrugations to securely resist the greater crushing forces possible in ULSD units. In parallel, refined in-shop testing methods were developed to allow the detection of very minute leakage. Figure 2 shows the LP HDS stripper bottoms HE developed to recover stripper bottoms heat into stripper feed (cold high-pressure separator (CHPS) designs) or into reactor feed (hot high-pressure separator (HHPS) designs).

For moderate operating conditions (combined pressure/temperatures criteria), considerable cost is saved by eliminating the pressure vessel in favour of a simple set of cross-beams clamped by tie rods to compress the plate pack. Avoidance of expansion bellows also contains PHE cost.

## Improved heat recovery

The flow space between each Packinox PHE plate-pair works like a static-mixer channel fed by a long, narrow inlet slot. Unlike the 2cm span of each round-tube mouth in S&T exchangers, the 100cm span of each long-slot mouth in a Packinox exchanger is virtually impossible to block over its full length. Thus, turbulent two-phase flow can always enter and spread laterally to fill every PHE channel, with no troublesome dead zones. This ideal flow pattern generates overall heat transfer coefficients (OHTC) about twice the traditional S&T values for the same service.

Combining high OHTC, large surface areas and true counter-flow makes it technically and economically feasible for these exchangers to achieve very low hot approach temperature (HAT) values. Thus, these HDS exchangers deliver HAT between 12°C and 30°C, rather than the 50–80°C typical of most S&T exchangers. Smaller HAT means better waste heat recovery, so heaters burn less fuel, HP coolers are smaller and cheaper, required stripper feed heat-spikes are lower and ULSD product runs down cooler.

## Fouling in HDS F/E exchangers

Just as it lifts OHTC, the previously discussed static-mixer flow pattern also usefully retards fouling. Several HDS F/E exchangers treating a feed mix of atmospheric gasoils and FCC cycle oils have successfully demonstrated their ability to maintain steady high-heat recovery over a period of several years of operation. However, with heavy cracked feed from non-blanketed storage, any type of HDS F/E exchanger will inevitably suffer feed-side fouling from peroxide and olefinic gums deposits. PHEs show an inherent advantage in that respect, as their configuration allows easy de-gumming.

Although salt deposits have never been observed on high efficiency PHE cooling distillate HDS reactor effluent down to as low as 90°C, it is well established that ammonium chloride usually sublimes in this general service at about 170°C. One plausible theory for the lack of observed salt deposits in the PHE is that the high-turbulence of the largely liquid HC stream effectively scrubs the salts off the plate surfaces, sending them to the next process equipment downstream. The sublimation temperature depends on HDS intake levels of nitrogen and chlorine from organic N & Cl in feed, and of HCl in makeup gas.

With water-drained feed, water dew point in HDS reactor effluent is comfortably below HP-separator temperature, so any deposited salt would remain hot, dry and non-corrosive. Should evidence of dry salt fouling develop (steadily reducing HE duty and increasing HE delta-P), such fouling could be easily removed by an occasional water wash, either online or off-line.

PHE inlet faces on both hot side and cold side can be obstructed by slugs of particulate solids. These originate as construction debris, mill-scale from feed tanks and pipes, small ceramic chips from newly loaded reactor catalyst, or iron sulphide flakes dislodged from stripper column internals during a heat-spike upset. For all PHEs, fouling by entrained particulates is best avoided by installing a simple wire mesh screen filter on each inlet line. Packinox recommends and supplies designs for such filters, normally with a mesh size of 500 micron (0.5mm).

For the reactor feed inlet line, a 25 micron screen-size is preferred to also protect the HDS reactor. Otherwise the inevitable fines that sweep straight through any type of feed HE will start plugging the top layer of catalyst bed. Fines below 25 micron will be harmlessly washed through the reactor bed.

Without the recommended inlet screen filters, several early PHE exchangers did suffer unforeseen solids intake. The PHE face obstruction-raised dP without hurting heat duty. Manual cleaning during short shutdowns has been effective in some but not all cases, highlighting the importance of the inlet screens.

## PHE de-coking

To cope with gum fouling, an effective, in-situ, low-cost steam/air decoke method was derived from cleaning fired heater tubes and for regenerating catalyst beds. This patented, PHE-specific method includes a water-wash pre-step plus an effluent-side gas-flow as heat sink to avoid hot spots from feed-side coke-burn. During in-situ PHE decoking, the same unique flow pattern that enhances OHTC and retards fouling now ensures that the steam/air mix quickly reaches and cleans every square inch of PHE surface area. This happens reliably even if some plate-pack inlet or outlet slots are partly bridged by coke or particulates.

Refinery trials confirm this new method gives a superb cleaning result, with as-new heat transfer being fully restored. For example, a decoke trial in September 2002 records that when the PHE was new–fouled–de-coked, its HAT was 30–89–30°C. .

By contrast, individual tubes in S&T exchangers may totally block, ruling out in-situ steam/air decoking. This then demands messy exchanger opening followed by mechanical cleaning of individual tubes (by drilling and/or HP water-jet cutting) or heat-soaking the whole tube bundle in some large remote furnace. The final difficult S&T step is leak-tight re-joining of many large flanges.

All-welded PHE construction eliminates all S&T-type full body flanges in hot HP hydrogen service. This greatly reduces the risk of combustible vapour leaks and fugitive emissions of toxic $H_2S$. Refiners see this as a big step forward in plant safety and environmental

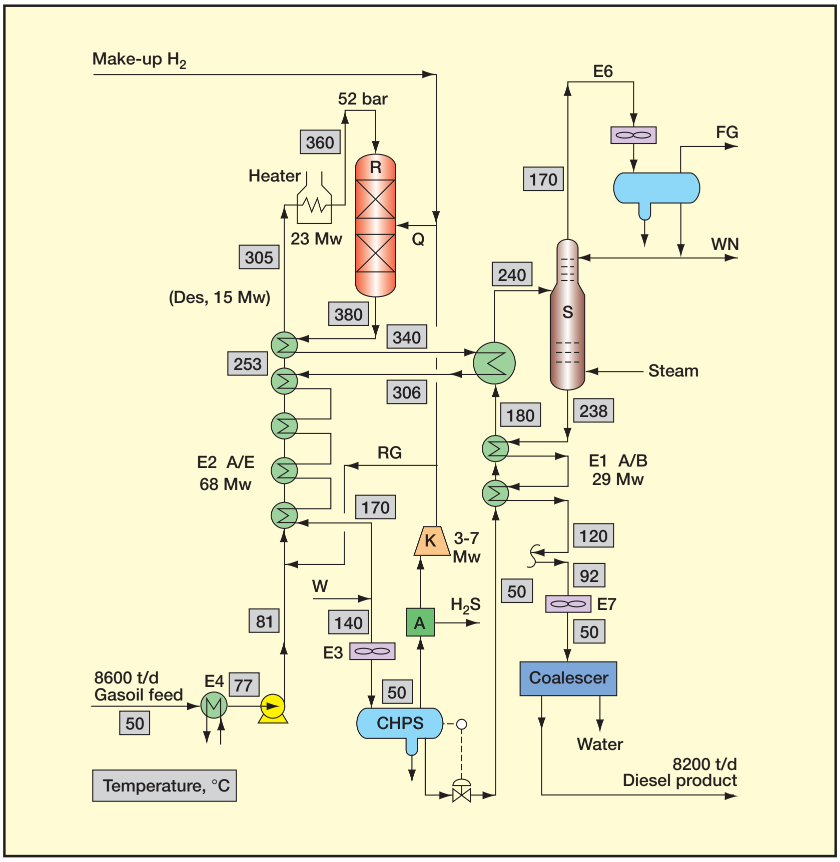


*Figure 3* Cold separator HDS – shell and tube HE

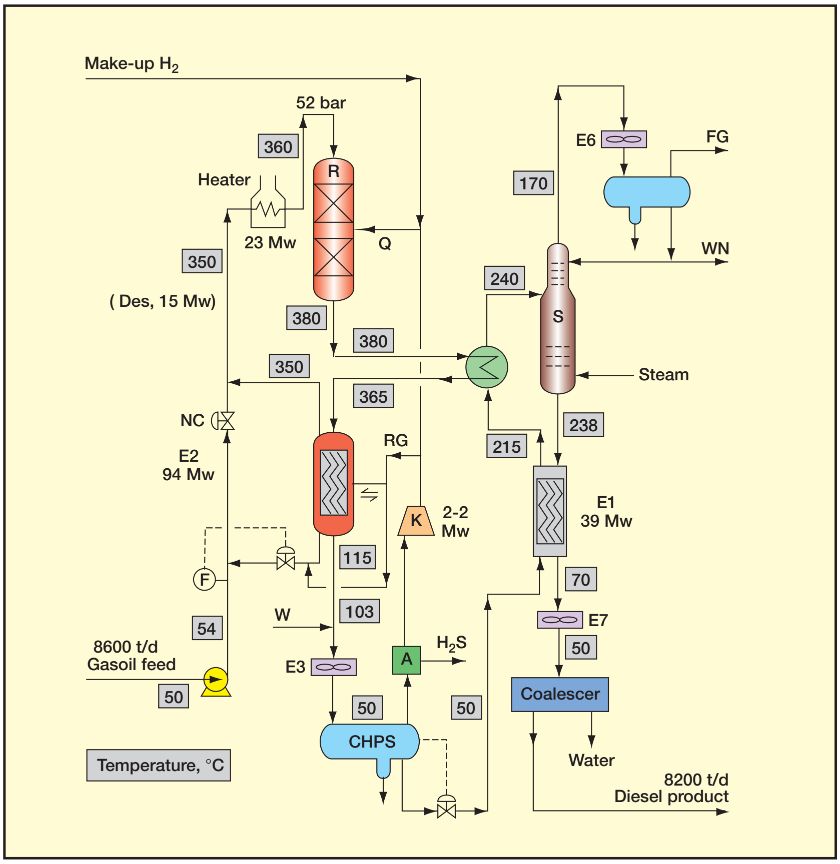


*Figure 4* Col separator HDS – Packinox PHE

performance. Stack emissions of $SO_X$, $NO_X$ and $CO_2$ are also reduced in line with lower heater duties. One PHE-based HDS-type unit with a high reactor exotherm now operates very well with its fired heater bypassed and stack emissions cut to zero.

Because the PHE shell, while designed for nearby process temperatures, is always substantially cooler than the PHE plate-pack, the pressure vessels can be designed with a generous mechanical safety margin.

## Avoidance of PHE corrosion

There are four corrosion types of most concern to users of HDS process equipment, including high-temperature corrosion (HTC), polythionic acid stress corrosion cracking (PTA SCC) and chloride stress corrosion cracking (SCC) of stainless steel (SS).

HTC by $H_2/H_2S$ or naphthenic acids is a concern against which alloy chromium content is the most important protection factor. The normal use of full austenitic stainless steel construction for the Packinox plate bundle considerably reduces HTC. For the same reason, the high temperature section of its low alloy steel vessel is clad with austenitic stainless steel.

PTA SCC is a concern for sensitised austenitic steel. It occurs, mainly during shutdown periods, in presence of water, oxygen and iron/chromium sulphide scales. To prevent this type of corrosion, it is necessary to use stabilised or low carbon stainless steel, and plate packs in that service use either 321 material selected at the low end of its carbon range, or 304L. In addition, tests performed on samples extracted from plates have shown a very low sensitisation level.

Chloride SCC of SS requires the complete combination exposure to chlorides, free water, dissolved oxygen, tensile stress and temperature between 60°C and 210°C. To avoid these harmful combinations, basic refinery equipment is installed to minimise concurrent entry of HCl, $O_2$, free-water and solids.

For cold reactor feed (which may still have some dissolved $O_2$ from non-blanketed tanks), the SCC-risk is largely removed by efficient water draining. Then, while still upstream of the PHE's SS plates, any remaining free-water haze disappears as the feed-mix is indirectly warmed by energy inputs from HDS charge pump and RG compressor.

For stripper feed from a CHPS, the oil phase is $O_2$-free because the reactor has converted any free oxygen to $H_2O$. Thus, even with water-haze and perhaps chloride traces, SCC cannot start because temperature ex-CHPS is too low. By the time stripper feed warms to well above 60°C in the LP PHE, any initial free-water haze is safely dissolved so that SCC is still not possible. For HHPS units, the LP PHE for stripper bottoms takes water-drained HDS feed from tankage or surge drum. As this oil has not yet seen any recycle gas (with potential HCl traces), no harmful chloride/free-water combination forms to contact PHE SS plates and promote SCC.

Pitting corrosion is galvanic corrosion, mainly under moist solid-deposits. This needs the combination of particulate deposits, chlorides, oxygen and free-water within the bundle as once occurred in a non-flushed PHE left opened-up in a damp, seaside atmosphere. Thus before opening for any prolonged shutdown, a PHE bundle should be washed on both sides with buffered $O_2$-free water to remove any small deposits that might promote pitting corrosion.

In summary, these few simple precautions should ensure years of corrosion-free service with PHEs: a surge drum with boot or a coalescer to remove free-water and debris from feed; fine-mesh feed filters to remove smaller entrained particles from the feed; alumina guard beds at the reformer to remove HCl traces from HDS fresh gas; gas-blanketing or floating roofs on feed tanks to prevent oxygen absorption.

Also, a buffered wash of both HE sides to remove possible small deposits should be done before opening to atmosphere.

**Overall heat balances: HDS case studies**

| All Cases : | Gasoil feed | 8600t/d | Reactor conditions: | |
|---|---|---|---|---|
| | Diesel prod | 200t/d | | In / Out |
| | Recycle gas | 320t/d | Temp, °C | 360 / 380 |
| | Quench gas | 210t/d | Press barg | 51 / 45 |

| HP Separator type | Cold Sep'r 50°C | | Hot Sep'r 250°C | |
|---|---|---|---|---|
| Design Case – figure : | 3 | 4 | 5 | 6 |
| Exchanger type | S&T | PHE | S&T | PHE |
| E1 Hot approach temp. °C | 58 | 23 | 69 | 27 |
| E2 Hot approach temp °C | 53 | 15 | 40 | 20 |
| Reactor circuit total ΔP bar | 19 | 11 | 13 | 9 |
| Plant heat balance MW | | | | |
| Heat In (to nearest MW) | | | | |
| - Charge heater | 23 | 5 | 10 | 0 |
| - Exotherm | 22 | 22 | 22 | 22 |
| - Electric drives (P1+K1) | 5 | 3 | 4 | 3 |
| - Stripping steam | 3 | 3 | 3 | 3 |
| Total In | 53 | 33 | 39 | 28 |
| Heat Out | | | | |
| - E3, High pr cond'r/cooler | 38 | 22 | 16 | 15 |
| - E6, Stripper condenser | 7 | 7 | 8 | 8 |
| - E7, Diesel rundown cooler | 8 | 4 | 15 | 5 |
| Total Out | 53 | 33 | 39 | 28 |
| Heat to feed mix to reach 360° | | | | |
| - Charge heater | 23 | 5 | 10 | 0 |
| - Elec drives | 5 | 3 | 4 | 3 |
| - E1, Heat from diesel prod | 0 | 0 | 28 | 38 |
| - E2, Heat from R1 effluent | 68 | 94 | 57 | 57 |
| - E4, Heat from diesel prod | 6 | 0 | 0 | 0 |
| - E5, Heat from HHPS vapour | 0 | 0 | 3 | 4 |
| Total MW = $Q_i$ | 102 | 102 | 102 | 102 |
| Heat ex R-Effluent to reach 50° | | | | |
| - E2, Heat to reactor feed | 68 | 94 | 57 | 57 |
| - E8, Heat to stripper feed | 17 | 7 | – | – |
| - E1, Heat to stripper feed | – | – | 28 | 38 |
| - E3, Heat to HP cooler | 38 | 22 | 16 | 15 |
| - E5, Heat to recycle gas | – | – | 3 | 4 |
| - E6, Heat to St cond'r from oil | – | – | 4 | 4 |
| - E7, Heat to diesel cooler | – | – | 15 | 5 |
| Total MW = $Q_o$ | 123 | 123 | 123 | 123 |

*Table 1*

## Exotherm control

Gasoil HDS reactions release considerable heat, usually in the range 1–3MW per 1000 ton/day gasoil feed. Plant operators must ensure that reactor outlet temperature does not rise above a preset level. For example, if the exotherm goes up by 1MW (with different feed), the most obvious way to compensate is by tuning the reactor charge heater down by 1MW.

The fact that Packinox exchanger technology allows heater duty to be greatly reduced (even to zero) may seem at first to remove the vital degree of control available with less efficient S&T exchangers. However this is simply not the case. A simple flow-controlled, partial bypass of feed (or of effluent) around the exchanger destroys feed pre-heat duty just as effectively as turning down a fired heater. This flow bypass method has been successfully implemented to safely regulate an HDS reactor temperature inlet without the help of a heater.

The manageable exotherm peak ($X_{max}$) is roughly the total of installed cooler duties for HP reactor effluent and stripper bottoms. After all the control steps below have been exhausted, if the actual foreseen worst-case exotherm still exceeds $X_{max}$, only then should the installed HP cooler duty (for PHE or S&T) be increased above its normal operating value. For both HE-types, the normal exotherm control steps are :

—Increase quench gas to the reactor bed
—Reduce heater duty (ultimately to zero)
—Increase bypass around the F/E exchanger
—Line up a colder and/or less-exothermic feed
—Reduce feedrate
—Reduce reactor pressure.

Thus, a large fired heater is definitely not required to control reactor temperature runaways.

## Start-up heater duty

For most HDS units the target warm-up rate is in the range 20–40°C/hr. More rapid heating can cause flange leaks and other problems due to excessive or uneven thermal expansions, while slower rate may unduly delay the start of profitable normal operation. A total warm up rate of 8–16 hours would generally be considered satisfactory.

For the comparative case studies shown in Figures 3 and 4, the heater design size is halved (from 25 to 12MW) when multiple S&T exchangers are replaced with two PHEs. However, transient analysis shows that warm-up times are both in the 11–12 hour range. This is essentially because the Packinox option, with its close-approach design, rejects far less heat via the reactor effluent fin-fan cooler. Instead, it directs effluent heat efficiently back into feed so that more net heat from the heater goes into reactor warm-up.

The conclusion is clear; while a certain minimum HDS charge heater size is required for economic warm-up rate, this minimum can be relatively small when a close approach exchanger system is installed. Big heaters are not needed for warm-up or steady operation, they just increase emissions and risk levels, and cost more to build, operate and maintain.

When the exotherm decreases, the HDS charge heater's ability to maintain full design feedrate depends on its design margin (above normal duty), not its total megawatt size. Thus a modern HDS with a relatively small furnace (12MW, including 7MW margin as in Figure 4) would cope better with low exotherms than an older unit with a charge heater twice the size but with less design margin (25MW, but only 2MW margin as in Figure 3).

## Formosa HDS

In the mid 1990s, Packinox commissioned a large number of HDS studies for interested refiners in Asia, Europe and the Americas. For the average study-capacity of 35 000bpd, the average five-year benefit was US$10 million for selecting PHE rather than S&T exchangers. This equates to just under $3 million per 10 000bpd, for a mix of hot and cold HP separator designs.

In 1996 a major Taiwanese refiner (FPC) was planning two new 65 000 bpd (8600t/d) cold-separator HDS units. For all reactor and stripper F/E duties, FPC selected PHE (rather than S&T exchangers) after a careful re-design by the European licensor IFP confirmed this would save US$48 million in the refiner's five year business plan. In effect, this 1996 design study compared IFP's alternatives via Figures 3 and 4 to derive the following cost savings:

For each plant, a 20MW energy saving (direct fuel and electricity) was valued at US$3.3 million/year, or $16 million for

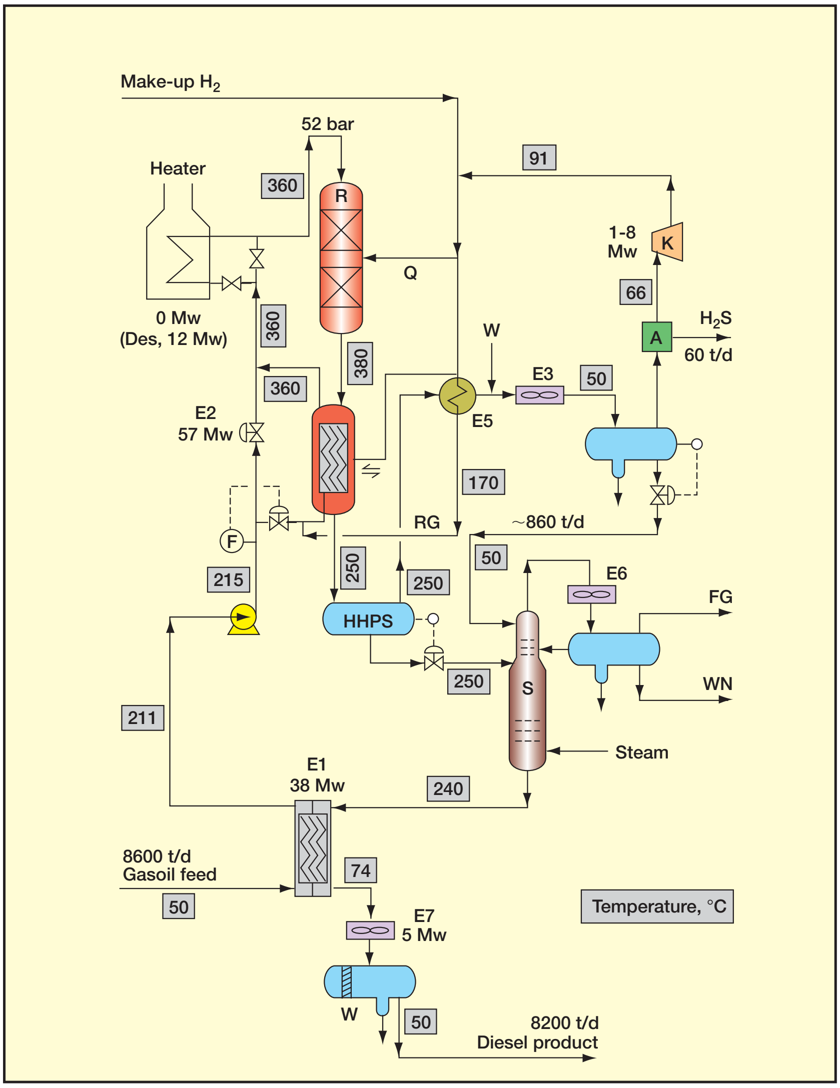


***Figure 6*** *Hot separator HDS – Packinox PHE*

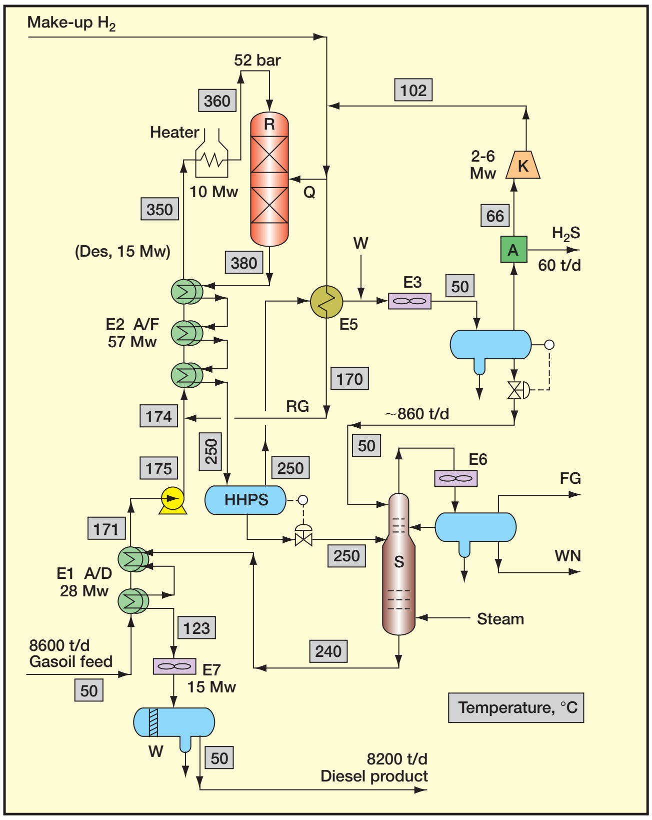


***Figure 5*** *Hot separator HDS – shell and tube HE*

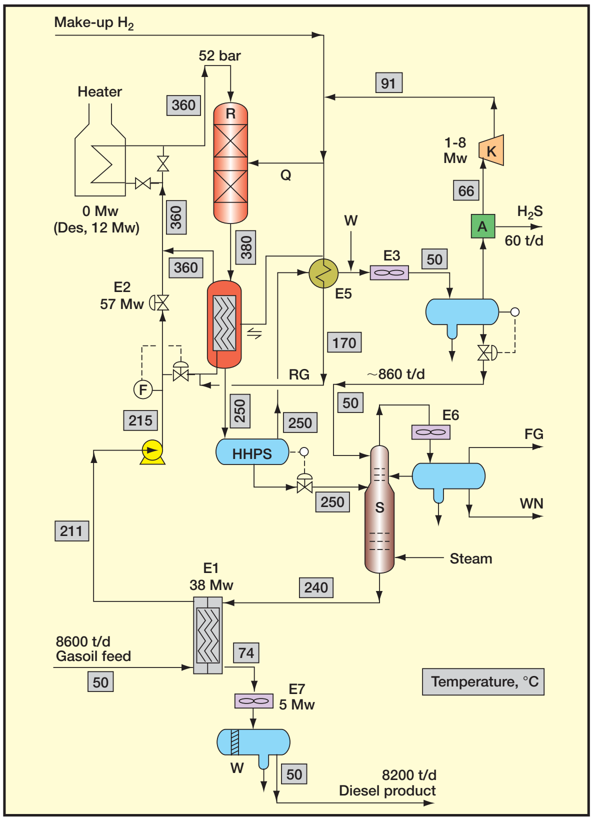


***Figure 6*** *Hot separator HDS – Packinox PHE*

the first five years of operation. In addition, fewer and smaller equipment items would cut installed capital cost by $8million. Thus, total five-year savings = (16 + 8) x 2 = $48 million for the two units. This equates to US$3.7 million per 10 000bpd (close to the above-mentioned study average). Today, FPC's project operates successfully as per Figure 4.

Of the $8million capex saved per unit, only $1 million was directly due to lower F/E exchanger purchase costs cf S&T costs. The other $7 million was due to spin-off effects, such as smaller fired heaters, coolers and compressors, plus lower civil and piping installation costs. This project highlights a common issue found on many projects, where the price difference between PHE and S&T equipment is relatively trivial but hides much larger savings elsewhere.

## Basic PFD types

Figures 3 through 6 now show four HDS cases for the same feed and reactor conditions. All exotherms are 22MW. As in Table 1, all temperatures are in °C. S&T vs PHE are compared – CHPS-designs in Figures 3 and 4, HHPS-designs in Figures 5 and 6.

Equipment tags used in Figures 3 through 6 are as follows:

H Charge heater
R Reactor
A Amine $H_2S$ absorber column
K Recycle gas compressor
S Stripper column
W Wash water injection
E1 Stripper bottoms main exchanger
E2 Reactor feed/effluent exchanger.

The table compares overall plant heat balances, including direct fuel and power consumption. These values lead to capital and operating costs for each of the four PFDs.

Table 1 also gives the heat input to raise feed mix (cold gasoil plus cold recycle gas) to reactor inlet temperature. This is a useful cross-check, since heat input must be the same for all four design cases. Similarly, the table also shows the heat removal to lower reactor effluent from 380°C down to 50°C. All energy data is given to the nearest megawatt. (1MW = 3.41 million Btu/hr).

For the Formosa CHPS project, revising the design from Figure 3 to Figure 4 saved $8 million in capex due to :

*Fewer* large heat exchangers
8 S&T → 2 PHE
*Smaller* charge heater design duty
25MW → 12MW
*Smaller* total fin-fan cooler design duty
50MW → 37MW
*Smaller* RG compressor
2 casings → 1 casing
*Smaller* electric motor drive for K1
4MW → 2.5MW
*Less* overall civil work, piping and plot space
920m$^2$ → 820 m$^2$

## HHPS design.

Changing from CHPS to HHPS design-type greatly lowers charge heater duty for both HE-types. Direct fuel use falls by over 50% for a S&T (compare Figures 3 and 5), and by over 90% for PHE (compare Figures 4 and 6). In actual megawatts, S&T gets closer to PHE whose lead erodes from 20MW to 11MW delta energy. Likewise, the installed capital cost difference is halved, changing from $8 million to $4 million.

Thus for the 65 000bpsd example HDS units of Figures 3 through 6, overall five-year cost saving (on moving from S&T to PHE) is $24 million (for CHPS units) or $12 million (for HHPS units) in this 1996 study.

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