

Clean diesel hydrotreating

Design considerations for clean diesel hydrotreating. Critical issues are discussed when designing a hydrotreating facility to produce diesel fuel with very low levels of total sulphur

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For over three decades, refiners worldwide have been implementing various projects in their facilities to accommodate a variety of regulations to improve the quality of transportation fuels in order to reduce vehicle emissions. One of the key areas of interest has been the reduction of sulphur in diesel fuel to very low levels. For example, since mid-2006, the maximum sulphur content of on-road diesel fuel in the US has been limited to 15 wppm. Similar regulations are in place or are in various phases of implementation in many countries.

Design considerations

Middle distillates contain various types of sulphur species, including mercaptans, sulphides, thiophenes and aromatic

sulphur compounds. Sterically hindered dibenzothiophenes are a group of aromatic sulphur compounds that are among the most difficult to remove when hydrotreating to very low sulphur levels. This is particularly true for diesel fuels that contain significant quantities of cracked stocks, such as FCC light cycle oil (LCO), which contains a large concentration of aromatic sulphur compounds. The effective removal of these species requires tailored catalysts and process conditions, as well as consideration of other factors such as feed nitrogen content and aromatics equilibrium.

There are numerous issues to be addressed in the design of a hydrotreater, including:

- Feed characteristics and variability
- Other product quality requirements, especially cetane index
- Catalysts selection
- Optimisation of reactor process variables
- Equipment design requirements
- Reliability
- Minimising product contamination
- Handling of off-spec diesel product.

All of these factors should be carefully considered during the front-end process design.

Process flow

Figure 1 shows a simplified process flow diagram for a diesel hydrotreater. Fresh feed from the surge drum is heated with stripper bottoms, then mixed with

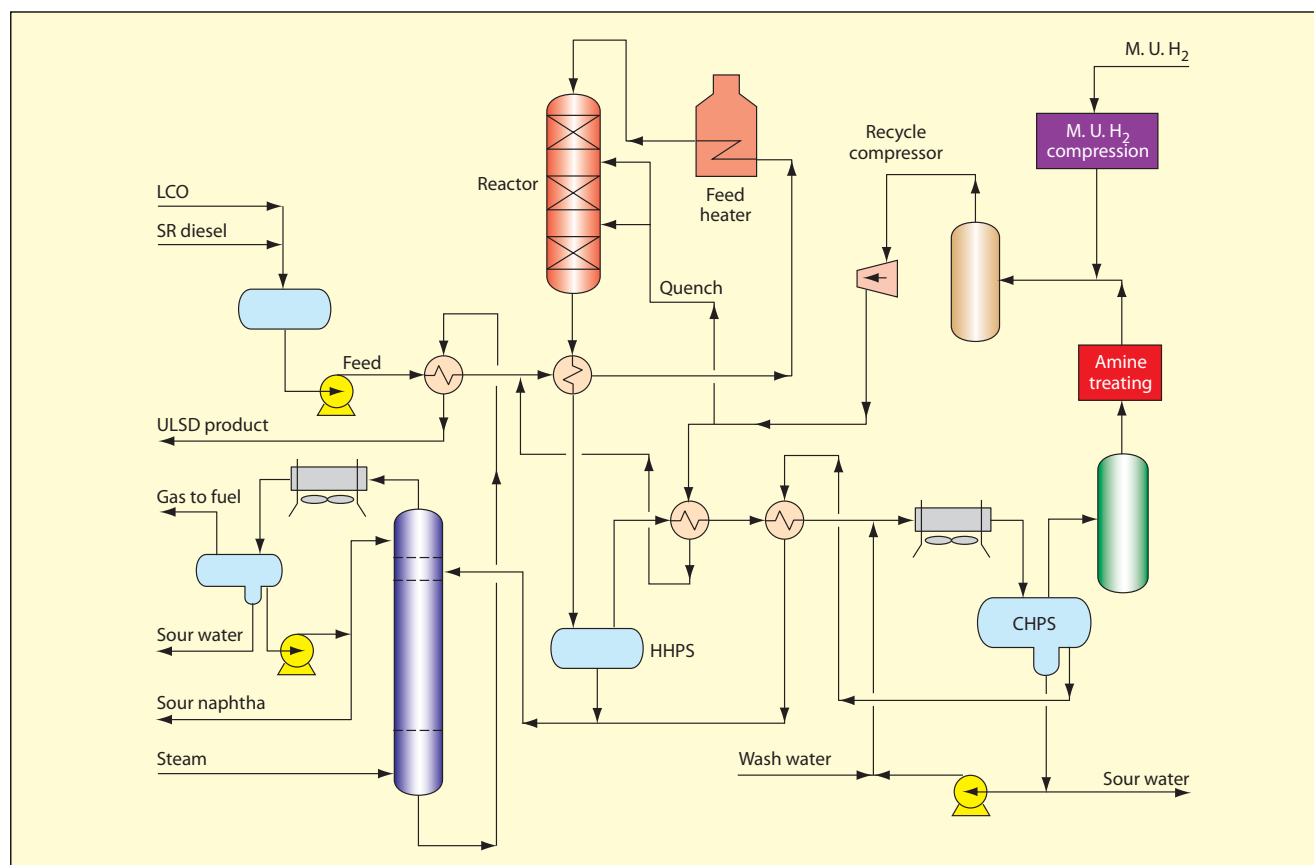


Figure 1 ULSD hydrotreater HHPS/CHPS

Maximum operating pressures for 321 SS and 347 SS stainless steel

	600lb flanges		900lb flanges		1500lb flanges	
	321 SS	347 SS	321 SS	347 SS	321 SS	347 SS
600°F	975	1030	1460	1550	2435	2580
650°F	950	1015	1425	1520	2375	2530
700°F	930	995	1395	1490	2330	2485
750°F	915	985	1375	1475	2290	2460
800°F	900	975	1355	1460	2255	2435
850°F	895	970	1340	1455	2230	2425

Table 1

recycle hydrogen. The combined feed is further heated by reactor effluent, then brought to reactor inlet temperature in the charge heater. Reactor inter bed quench may be required in one or more locations, depending on the volume of cracked stocks (FCC LCO, light coker gas oil) in the feed. This flow diagram shows recycle gas being used as quench. Reactor effluent exchanges heat with the combined feed and flows to the hot, high-pressure separator (HHPS). Vapours from the HHPS are used to heat recycle gas and stripper charge before being cooled in the reactor effluent air cooler (REAC) and entering the cold, high-pressure separator (CHPS). Wash water is injected upstream of the REAC to remove ammonium bisulphide. HHPS liquid is combined with heated CHPS liquid and flow to the product stripper.

Vapours from the CHPS are contacted with amine in a scrubber for hydrogen sulphide removal and flow to the recycle compressor suction drum. Make-up hydrogen is compressed and combines with the recycle gas in the suction drum. Some of the recycle gas may be purged from the compressor suction drum to improve the purity of the recycle gas hydrogen. If needed, the quantity will depend on the reactor hydrogen partial pressure requirements and the purity of the make-up hydrogen.

In the product stripper, superheated steam is introduced into the bottom of the tower to effect the removal of hydrogen sulphide. Stripper overhead vapours are condensed and flow to the stripper accumulator. Accumulator vapour and liquid (wild naphtha) are processed in other off-site facilities. The stripped diesel product is used to heat the feed, it is cooled, then it flows to drying facilities (not shown). This could be a coalescer/salt dryer or a vacuum drying system.

Feed and product characteristics

Sulphur, nitrogen and aromatics content are the most important feed characteristics that impact the process design for diesel hydrotreating facilities. The nitrogen content of the feed has a significant impact on the required operating pressure for a new design.

Nitrogen has to be removed to essentially the same level as sulphur to reach the ultra-low target. This means the catalyst employed and the hydrogen partial pressure selected must be consistent with a high nitrogen removal operation. Normally, the bulk of the feed nitrogen is contained in light coker gas oil and FCC LCO. The aromatics content of the feed will govern the chemical hydrogen consumption at the low space velocities and high hydrogen partial pressures required for very low sulphur diesel production.

Generally, cracked stocks can be included in the feed up to the level limited by the product cetane index or gravity without having a significant impact on hydrotreater design. There is a small increase in the gravity and cetane index during the hydrotreating reaction. If a significant improvement in cetane (say, three to five units or more) or gravity is required, a multi-stage design using aromatics saturation catalysts in the second stage may be the more economical option. The final choice will be driven by the magnitude of the improvement in gravity and cetane required.

Obviously, the design product sulphur target is a key issue, not only from a process design standpoint, but also for off-site storage and transfer. For the ultra-low sulphur diesel (ULSD) programme in the US, most new and revamp facilities have been designed for a product sulphur content of 8–10 wppm to ensure a final product sulphur content of 15 wppm at the point of sale.

Pilot testing of the feed is practically mandatory to confirm reaction process conditions. Testing for variations in feed characteristics, especially FCC LCO and coker light gas oil back-end distillation, should also be considered, because the separation achieved in the products fractionators from these facilities is notoriously poor. This can result in a temporary spike in the content of the most difficult-to-treat sulphur compounds in the hydrotreater feed and requires an increase in reactor temperature. This will increase the catalyst deactivation rate and, hence, directionally reduce the cycle length.

Reaction process variables

The key reaction process variables are:

- Space velocity
- Hydrogen partial pressure
- Make-up hydrogen purity
- Ratio of total hydrogen to reactor/chemical hydrogen consumption
- Cycle length
- Reactor temperature.

For feeds that have a significant aromatics and/or nitrogen content, a nickel-moly (Ni/Mo) or nickel promoted cobalt-moly (Co/Mo) catalyst will be used, along with an appropriate selection of graded catalysts in the top of the bed to mitigate reactor pressure build-up.

For a given cycle length and treating severity reactor space velocity, hydrogen treat gas quantity and hydrogen partial pressure are the variables that are optimised during the process design along with reactor temperature. As the hydrogen partial pressure is increased, the catalyst deactivation rate is reduced. Thus, the space velocity can be increased accordingly for a constant cycle length. However, this is at the expense of higher hydrogen consumption. Another important variable is the ratio of total hydrogen supplied to the reactor bed to the chemical hydrogen consumption for that bed. This value should be between five and six for feeds with significant quantities of cracked stocks. This ratio is normally a “fall-out” number during the process design, but there are instances where it can control the design hydrogen circulation rate, particularly for cases where high purity make-up hydrogen is used exclusively.

From a practical viewpoint, one should be aware of the limiting pressure of the alloy piping flanges in the reactor section when setting the hydrogen partial pressure and total operating pressure. This includes piping from the combined feed exchangers to the feed heater, from the heater to the reactor, and from the reactor to the combined feed exchangers (especially when the feed contains significant quantities of cracked stocks). This piping will be a 300 series stainless steel, and for 600 psi ANSI flanges this corresponds to a maximum operating pressure of around 800 psig at the reactor inlet when using 321 SS (Table 1) and 880 psig when using 347 SS. For 900 psig flanges, the reactor operating pressure can be increased to about 1200 psig when using 321 SS and 1300 psig when using 347 SS. This takes into consideration pressures during relieving conditions. These criteria should not necessarily govern the selection of the operation pressure. However, just barely exceeding these limits and requiring the next higher rated ANSI flange class in the reactor loop will be very costly. Actual pipe wall thickness is calculated for the design conditions and the limiting value used for the piping specification in the reactor

loop alloy sections. The piping material selection between 321 and 347 stainless steel can be optimised, based on design and operating temperature. Type 347 has a higher allowable stress value, as shown in Table 1.

For revamp designs, the hydrogen partial pressure achievable is restricted by the existing equipment and piping mechanical design and hydraulics in the reactor loop. A higher treat gas rate can be used to increase the hydrogen partial pressure, but this is usually limited because of the associated increase in the reactor loop pressure drop and the corresponding maximum operating pressure of the various system components.

Make-up hydrogen purity impacts the hydrogen partial pressure for a fixed reactor operating pressure. Lower purity make-up hydrogen requires higher hydrogen circulation rates to maintain the target hydrogen partial pressure and may even require a purge stream from the cold separator. If the make-up hydrogen purity is too low, there is no combination of recycle rate and purge that can be used to achieve the target reactor outlet partial pressure. For a revamp design, increased make-up hydrogen purity is the most effective means of increasing the hydrogen partial pressure.

For new designs, the cycle lengths have typically been set at 24–36 months. This has been based on the logic that, at some point, the cycle will be limited by factors other than catalyst activity: namely, reactor pressure drop. Hydrogen partial pressure has a major impact on cycle length from a catalyst activity standpoint. For a fixed space velocity, the cycle length increases with hydrogen partial pressure.

Maximum reactor outlet temperature at end-of-cycle catalyst conditions is generally set at 725–750°F to avoid aromatics saturation equilibrium constraints. This is also influenced by the quantity of cracked stocks in the feed and the crude source. Hydrotreating catalyst performance correlations for reactor temperature are usually based on the weighted average bed temperature (WABT). WABT is calculated as the reactor inlet temperature plus two-thirds of the reactor temperature rise. The temperature rise is usually limited to 40–50°F per bed by quenching. Thus, for a 50°F temperature rise and a 725°F maximum reactor outlet temperature, the end-of-run WABT would be:

$$\text{WABT} = (725 - 50) + 2/3(50) = 708^\circ\text{F}$$

The start-of-run WABT has to be sufficient to obtain the required removal of sulphur and nitrogen. Cycle length is determined by the catalyst deactivation rate at the design space velocity and

hydrogen partial pressure. During the cycle, the increase in WABT will be 30–50°F, with lower deactivation rates occurring at higher hydrogen partial pressures.

Reactor design

Initially, the reactor diameter is set to obtain a superficial mass velocity (mass flux) in the range of 2000–5000 lb/hr/ft². A value of 3500 lb/hr/ft² allows for a reasonable turn-down rate and some up side allowance. The mass flow includes all hydrocarbon and hydrogen at the reactor inlet. If the required diameter is larger than the limit for which the reactor can be shop fabricated and shipped to the plant site (12–14ft for over-land shipment), a two-train design may be considered. For diesel hydrotreaters, this is normally in the range of 30 000–40 000 bpsd per reactor train. For plants that have adequate water access and dock facilities, larger reactors and thus larger single-train capacities can be considered.

For diesel hydrotreating, the flow through the reactor is two phase, so an appropriate correlation, such as that by Larkins,⁴ is used for estimating the pressure drop. Inter-bed quench should also be considered for the pressure drop calculation. The catalyst physical properties have a significant impact on the reactor pressure drop. This includes void fraction and equivalent particle diameter. In order to ensure good distribution, most ULSD catalyst beds are dense loaded instead of sock loaded. The calculated clean pressure drop should be in the range of 0.5–1.0 psi per foot of catalyst bed (clean).

According to the required catalyst volume, the bed heights are established based on either the heat of reaction/maximum temperature rise or a bed height of 30–40ft. The clean pressure drop is then estimated for each bed. The maximum overall pressure drop is determined for the design of the bed support and to compare with the crush strength of the catalyst. This value is the sum of the following:

- Fouled bed pressure drop at 175% of calculated clean drop
- Dead weight of catalyst and support material
- Coke deposits at 30% of catalyst dead weight
- Liquid hold-up
- 15 psi allowance for depressurisation.

The pressure drop for the feed distributor/redistributors, quench internals and collector are added to the bed values to arrive at the overall fouled reactor pressure drop. The design of the feed distributor and redistributors is crucial for obtaining the target product sulphur level. The distributors and quench internals are usually catalyst vendors' proprietary designs.

There are several ways to provide internal quench to the reactor to limit the temperature rise. The common method is to utilise hydrogen-rich recycle gas/make-up hydrogen. The quench gas also causes additional vapourisation of some liquid in the reactor, which provides an extra heat sink for reaction heat removal. Another less common approach is to recycle CHPS liquid to the reactor. As noted previously, quench is added to limit the reactor bed temperature rise to 40–50°F.

The reactor metallurgy is typically 11/4 Chrome-1/2 Moly with 1/8in 347 stainless overlay or cladding. The choice often depends on the thickness of the vessel, which will affect the cost. Cladding is usually more economical for vessels with wall thickness up to around 4.0in. For greater thickness, overlay is normally used. For thick-wall reactors with a minimum design metal temperature of less than 0°F, 2 Chrome-1 Moly metallurgy may be the better choice.

Feed exchangers

The feed exchangers are generally specified to optimise heat economy. However, enough duty should be allowed for the charge heater to have a reasonable amount of turndown and overall heat balance control. The feed exchange should not be more than about 80–85% of the total duty needed to heat the feed to reactor temperature at end-of-cycle conditions.

For very low sulphur diesel product, the reactor feed/effluent exchangers' mechanical design requires special attention to minimise leaks. Many refiners prefer a pull-through tube bundle with a floating head because this design is easier to clean and handle. However, the floating head cover is a potential source for leaks. A U-tube bundle avoids this potential problem and is less expensive. Also, tubes should be seal welded to the tube sheet to prevent leaks from the rolled tube joints.

Materials selection for the feed/effluent exchangers is based on predicted corrosion rates for the service conditions. The feed/effluent exchangers have several shells in each train in series with varying metallurgy. The cold shells are all carbon steel. The intermediate temperature units normally have 11/4 chrome shells and 400 series tubes. The hot shells are 11/4 chrome with 300 series stainless cladding and 300 series stainless tubes.

Charge heater

As previously noted, the feed heater design duty should not be any less than approximately 20% of the combined feed duty to the reactor inlet temperature at end-of-cycle conditions to allow for

turndown and heat balance control. A design margin of 10% of the combined feed exchanger duty should be added to accommodate accelerated exchanger fouling and loss of reactor heat of reaction caused by removing highly reactive feeds such as FCC LCO and coker light gas oil. The design duty of the charge heater should be checked to ensure it is sufficient to bring the unit up to operating temperatures from a cold start in a reasonable length of time.

Typically, the number of heater passes matches the number of parallel feed/effluent exchanger trains. The feed and recycle hydrogen can be flow controlled for each train ahead of the exchangers, which will prevent any maldistribution of the two-phase stream entering the heater passes. When there are more heater passes than feed effluent exchanger trains, particular care must be made to ensure equal distribution of vapour and liquid to each heater pass. The limit for one heater pass is in the range of 15 000–18 000 bpsd of fresh feed at typical diesel hydrotreater conditions. This limit is based on 8in diameter maximum heater tubes and an overall pressure drop of 40–50 psi. Heater tube metallurgy is typically 347 stainless steel.

Environmental issues such as NO_x and CO emissions are also a concern in the charge heater design. NO_x emissions limitations are usually addressed by installing low NO_x or ultra-low NO_x burners. CO emissions can be a concern

with heaters with a required turndown to less than 50% of design. This can be the case with a charge heater with a low SOR duty and a high design duty. In this case, the turndown can be 10–20% of design. Maximising energy efficiency can lead to low SOR duty. High design duties are a result of off-design cases such as loss of reactive feed or cold start-up.

Reactor products separation

There are several alternative arrangements for handling the reactor effluent after the combined feed exchangers. In general, the decision is whether or not to use an HHPS in addition to a CHPS, or a CHPS only. The HHPS is usually operated around 500–550°F and, thus, the vapours can be used to heat recycle gas and cold separator liquid. The HHPS improves the overall heat economy and results in a somewhat smaller product stripper and auxiliary equipment. This arrangement will also directionally improve the oil-water separation in the CHPS. For heavy gas oil hydrotreating, a hot separator is practically mandatory to have adequate oil/water separation. This is less of a concern for diesel hydrotreating, unless there are significant quantities of cracked material, which decrease API gravity. Disadvantages of this design are a 5–10% lower recycle gas purity, which requires higher recycle hydrogen compressor horsepower. The addition of a HHPS can be a viable option for debottlenecking an existing hydrotreater that is

hydraulically limited in the reactor loop. This revamp would require, however, a modest increase in inlet flow and power for the recycle compressor.

Other options include the addition of hot and cold low-pressure separators. These designs marginally unload the upper part of the stripper and result in improved LPG recovery. Figure 1 illustrates an ULSD hydrotreater with HHPS/CHPS, Figure 2 an ULSD hydrotreater with CHPS only and Figure 3 a hydrotreater with HHPS/CHPS/CLPS.

The HHPS base material is 11/4 Cr-1/2 Moly normally with clad or overlay. The lining is 300 or 400 series stainless, depending on the actual predicted corrosion rate. For equipment in wet sour service, hydrogen induced cracking (HIC) resistant carbon steel material should be specified. This requirement applies to the CHPS. A 300 series stainless steel mesh pad is installed in the CHPS to assist in coalescing the wash water. The boot of the CHPS should have 1/4in corrosion allowance.

Reactor effluent air cooler

Regardless of the reactor products separators configuration, a final cooling step is eventually used prior to separating liquid and vapour. An air cooler is typically used, sometimes supplemented with a water trim cooler. Metallurgy selection of the piping and equipment in this area requires special attention because of the presence of ammonia and hydrogen sulphide in the reactor

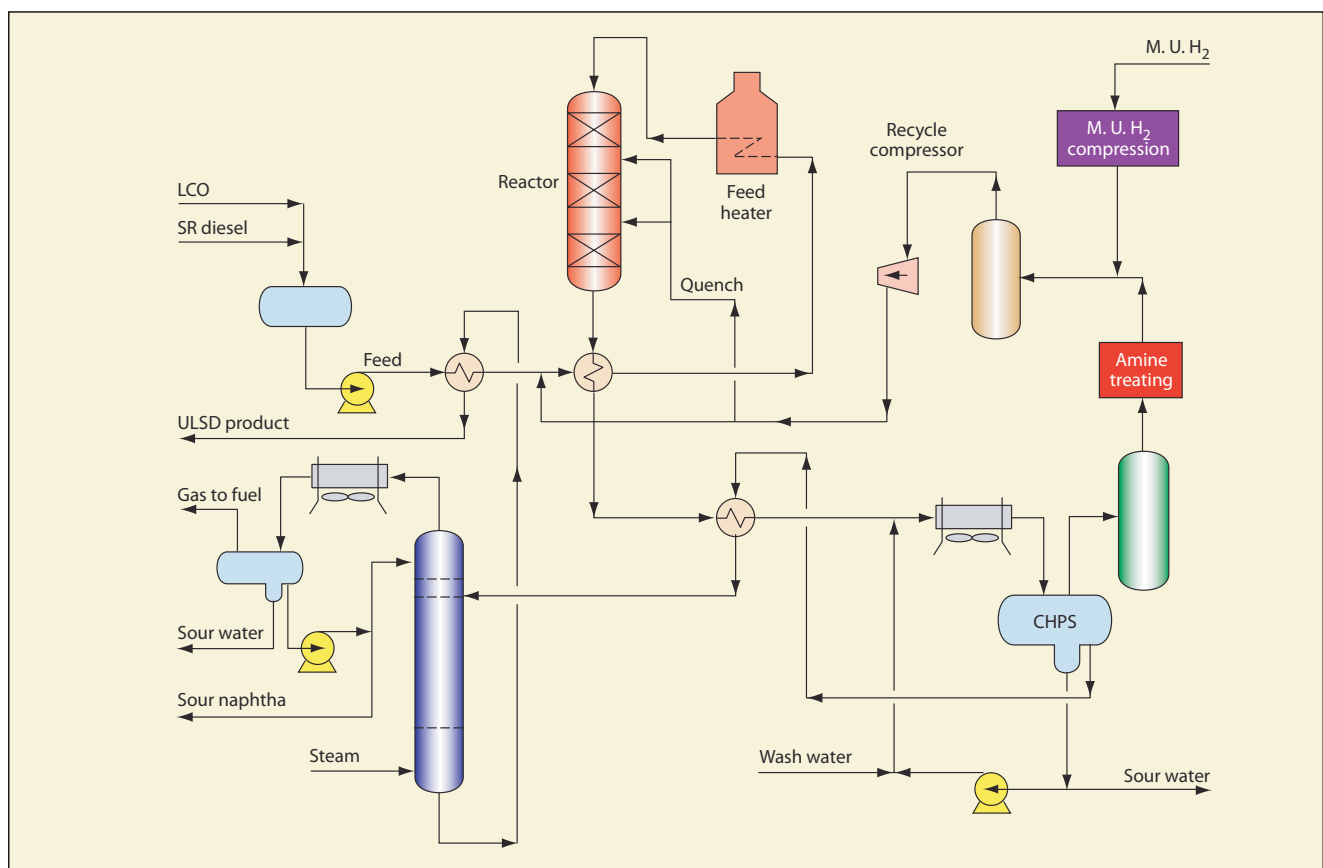


Figure 2 ULSD hydrotreater with CHPS only

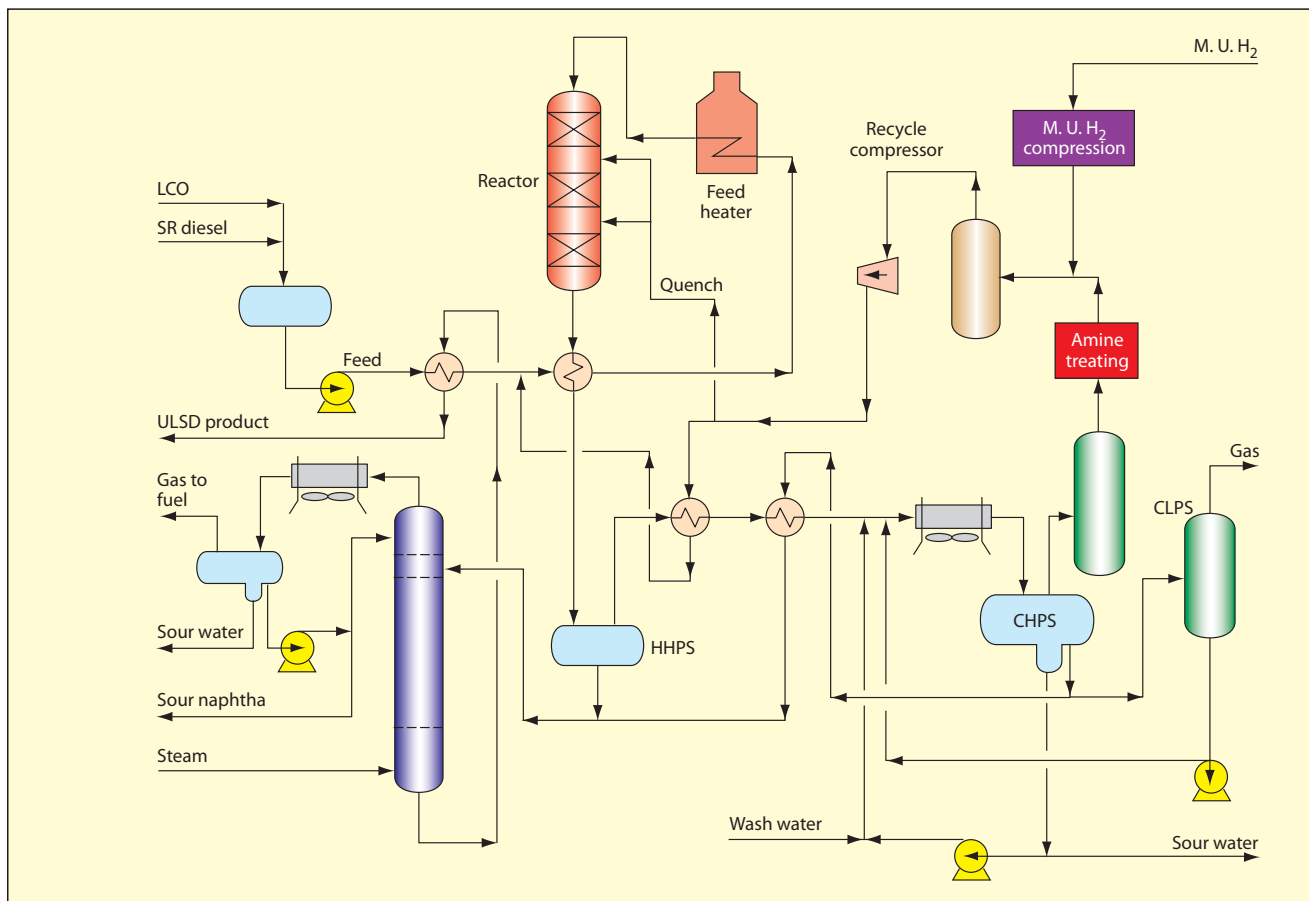


Figure 3 ULSD hydrotreater with HPPS/CHPS/CLPS

effluent. As this stream is cooled, these two compounds combine to form ammonium bisulphide (NH_4HS), which condenses as a solid on the inlet piping and in the cooler tubes. To prevent plugging and under-deposit corrosion, recirculating wash water is injected ahead of the cooler, which introduces a corrosive aqueous solution in the stream.

Tube metallurgy selection is based on the product of the mole percentage of ammonia and hydrogen sulphide in the total stream before water injection. The target maximum value for this product, also known as K_p , is 0.15 (5) before alloy metallurgy is considered:

$$K_p = \text{Mol \% H}_2\text{S} \times \text{Mol \% NH}_3 < 0.15$$

Other criteria include:

- Determine make-up water quantity to limit NH_4HS concentration in the condensed water phase to 4 wt%. Assume equal molar amounts of ammonia and hydrogen sulphide in the CHPS water and that essentially all of the ammonia is dissolved in the water
- Determine circulating water quantity as that needed to maintain a minimum of 20–25% of the total water injected in the liquid phase ahead of the condenser
- Mixed-phase velocities in the piping and tubes should be limited to 20 ft/sec for carbon steel and 30 ft/sec for alloy.

Minimum velocity should be 10 ft/sec — Symmetrical piping in and out of the condenser

— Wash water source can be stripped sour water, but should be from a stripper dedicated to hydrotreating spent wash water service

— Wash water can be injected into the main effluent line upstream of the air cooler or independently into each air cooler nozzle through a distribution device such as a spray nozzle or restriction orifice. Injecting in the piping gives more time for mixing and vapourising the water and is less expensive. Injection into individual air condenser nozzles ensures good distribution in each of the air cooler bays.

If alloy construction is required, typical alloys used are Incoloy 800 or 825 and duplex stainless steel 2205.

High-pressure separator liquid level control valve sizing

In the high-pressure separator, the operating pressure is greater than the critical pressure of the liquid. The high-pressure separator liquid level control valve is in flashing liquid service. Typically, both the vapour pressure and critical pressure are required to determine the effective pressure differential in the sizing of any control valve in flashing liquid service. In this case, the ratio of vapour pressure vs critical pressure is greater than one. This unusual situation

is not acceptable in most control valve sizing programmes. To bypass this problem, one of these two pressure values will need to be artificially manipulated. If the vapour pressure is artificially reduced to below the critical pressure, the effective pressure differential may be over-estimated and the control valve may be undersized. Based on actual experience, the control valve vendors recommend that the critical pressure is raised to equal to the vapour pressure when specifying instrument process data.

Compression

The choice of recycle compressor design is based on inlet flow and head requirements. To use a centrifugal compressor, adequate suction flow is required, but the head should be low enough to limit the number of stages, such that a single body machine can be used (usually maximum of ten stages). In the past, most diesel hydrotreaters did not have an adequate volume of recycle gas for a centrifugal compressor. New, larger ULSD hydrotreaters circulating 3000–5000 SCF of gas per barrel of feed will be candidates for a centrifugal compressor. If a reciprocating compressor is used for recycle, it will normally be combined with the make-up service in a multiple-throw machine. For reciprocating compressors over about 500 HP, two 60% capacity machines are

Typical pressure drop allowance for reactor loop equipment and piping in a hydrotreater with reactor inlet pressure of 1100 psig

Equipment item	Pressure drop, psi
Combined feed exchangers (both sides)	40
Feed heater	50
Reactor (fouled)	75
HHPS vapour/recycle hydrogen (both sides)	20
HHPS vapour/CHPS liquid	10
Reactor effluent air cooler	10
Amine/knock-out drums	5
Piping	23
Total reactor loop pressure drop	233

Table 2

Maximum pressure and typical design temperature for critical piping services in the reactor section

Pipe section	Design pressure, psi	Design T, °F	Max allowable, psi
Exchange to heater	1252	700	1260
Heater to reactor	1200	775	1242
Reactor to exchange	1123	825	1233

Table 3

typically provided. For smaller compressors, two 100% units are used.

The make-up compressor discharge may be routed to either the recycle compressor suction or discharge. Make-up gas flowing to the suction will increase the purity and volume of the recycle gas and decrease the molecular weight. Potentially, this could be a problem for a centrifugal recycle compressor, since the

lower molecular weight directionally requires more head and potentially more stages. This is not as serious an issue with a reciprocating machine. However, higher hydrogen purity will increase the discharge temperature, which is limited per API-618 to 275°F.

Normally, the make-up would be routed to the recycle compressor discharge unless this causes an additional stage of

make-up compression. This approach has the advantage of maintaining some hydrogen flow capability through the reactor loop following a recycle compressor emergency outage.

Reactor loop hydraulics

A hydraulic profile is established for the reactor loop in order to set the recycle compressor head requirements and to establish the design pressure of the piping and equipment.

The reactor inlet pressure is set based on conditions established by the catalyst supplier and, as noted previously, considering the pressure limits of standard piping flanges. Table 2 shows the typical pressure drop allowance for the reactor loop equipment and piping in a hydrotreater with a reactor inlet pressure of 1100 psig.

A point-to-point pressure profile is shown in Figure 4. Data shown are for the normal case (EOR) and for the relieving case when the CHPS is at 10% above operating pressure. Table 3 is a tabulation of the maximum pressure (at relieving conditions) and typical design temperature for critical piping services in the reactor section. This is compared to the maximum allowable pressure for 900 psi, type 321 stainless flanges.

This analysis shows that the assumed pressure profile is consistent with the use of 900 psi flanges in the reactor loop alloy piping. The relieving pressure

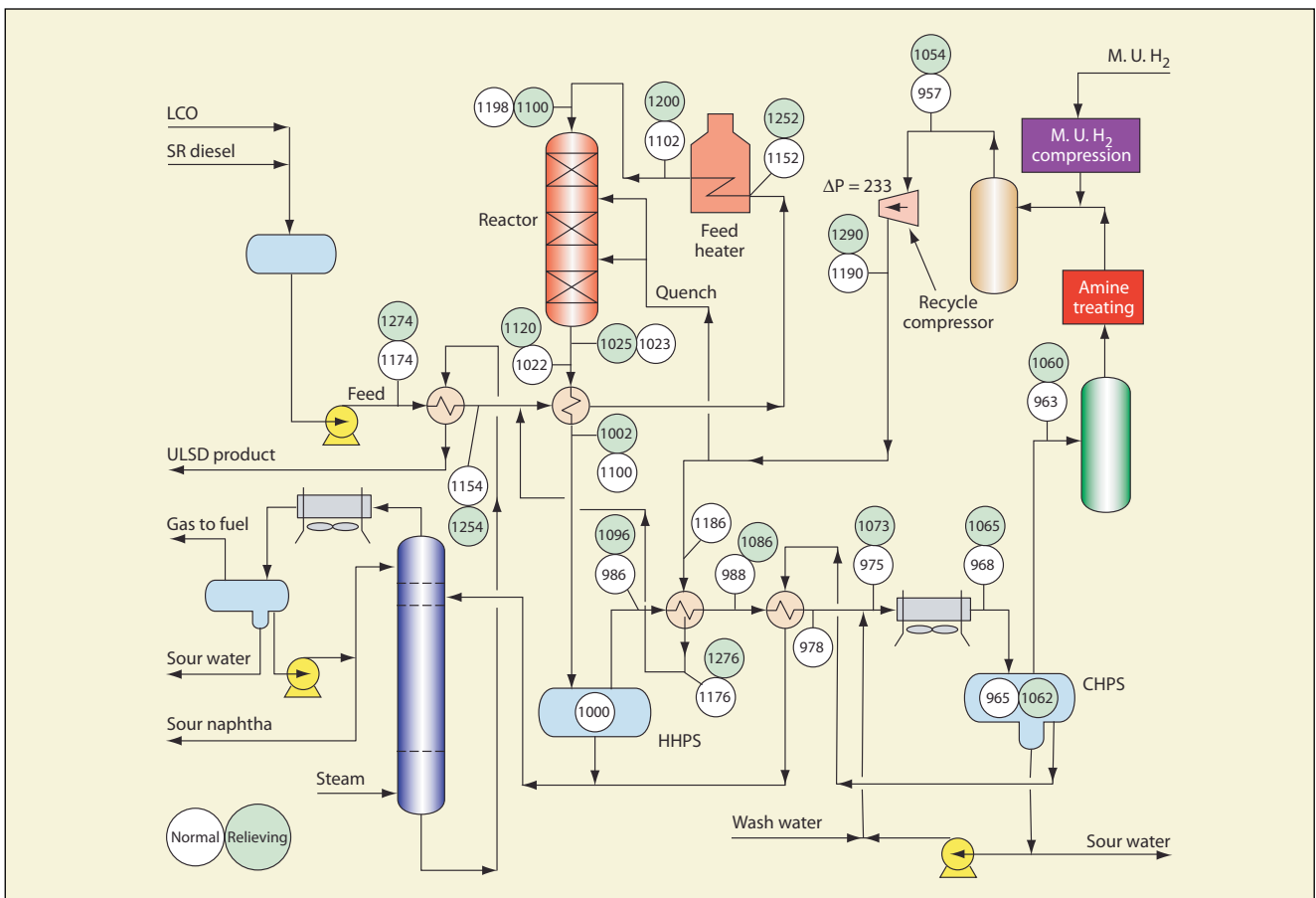


Figure 4 Normal/relieving pressure profile

Estimated equipment/materials delivery matrix

Equipment materials	Current quoted deliveries	Vendor data submittal (current durations)
Compressors (reciprocating)	70–92 weeks ARO	20 plus weeks ARO
(centrifugal)	70–84 weeks ARO	20 plus weeks ARO
Compressors (air)	52–64 weeks ARO	12–16 weeks ARO
Centrifugal pumps (API)	36–52 weeks ARO	16 plus weeks ARO
ANSI pumps	22–30 weeks ARO	10–12 weeks ARO
Agitators/mixers	26–28 weeks ARO	6–8 weeks ARO
Reactors (ULSD/alloy)	18–24 months ARO	6–8 weeks ARO (outline)
Reactors (2 1/4 Cr. - vanadium enhanced)	24–36 months ARO	6–8 weeks ARO (outline)
Large towers (all metallurgies)	54–68 weeks ARO	6–8 weeks ARO (outline)
Pressure vessels (carbon steel)	26–36 weeks ARO	4–6 weeks ARO (outline)
Pressure vessels (alloys)	48–56 weeks ARO	4–6 weeks ARO (outline)
Fired heaters	52–60 weeks ARO	12–14 weeks ARO
(heater tubes - high alloy)	36–48 weeks ARO	(mill delivery)
Shell & tube heat exchangers (carbon steel)	40–52 weeks ARO	6–8 weeks ARO
Shell & tube heat exchangers (alloys)	52–80 weeks ARO	6–8 weeks ARO

(ARO: After receipt of order)

Table 4

profile will be used to set the design pressure of the equipment in the reactor loop. Also, early in the process design, following completion of the heat and weight balance and a preliminary process flow diagram (PFD), a sized equipment list will be developed, which will allow plot plan studies to be completed. Preliminary piping sketches of the reactor loop will be developed and the allowance for piping pressure drop rechecked prior to purchase of the recycle compressor. The reactor loop hydraulics will be checked a final time following receipt of equipment vendor information and completion of piping isometrics for the reactor loop.

Note that the design pressures in the reactor loop were established by assuming the CHPS relief valve set pressure is 10% above normal operating pressure. In fact, this margin can be 5% by using a pilot-operated relief valve, which will lower the design pressure of the reactor loop equipment. Designing for a 10% margin for a grass-roots facility allows room for future capacity increases while staying within equipment design pressure limits. API recommended practice RP 521 appendix G recommends the design pressure to be 105% of the settling-out pressure, which would make the design pressure about 115% of the normal operating pressure. In the beginning of the process design stage when equipment design conditions are being established, there is not sufficient information available to accurately calculate the settling-out pressure. However, when the heat and material balance and PFD are completed, preliminary equipment sizing information and assumptions about pipe quantities along with the reactor loop pressure profile can be used to estimate the settling-out pressure. When actual

equipment and piping design information is available, the settling-out pressure is re-estimated to determine what margin is available compared to the CHPS relief valve set point. If the settling-out pressure exceeds the set pressure, a step-wise calculation or dynamic simulation can be used to estimate the relief rate.

Product stripper

In the product stripper, hydrogen sulphide formed in the reactor is removed from the diesel product along with light hydrocarbons. The removal of light material has to be sufficient to meet a flash point specification of 130°F. This also ensures removal of essentially all of the H₂S. Approximately 30 trays are required in the column. Overhead products include a sour gas, unstabilised naphtha and sour water. Equipment is included to provide intermittent water wash of the condenser.

Steam or vapours from a fired reboiler are used as the stripping medium. Due to temperature constraints (about 700°F maximum), the operating pressure of a reboiled stripper is limited to around 40–50 psig. Unless the refinery has a low-pressure gas recovery system, a small compressor and spare will be required to handle the off-gas. A steam stripper can be operated at over 100 psig, which allows routing into an existing refinery sour fuel gas system.

Diesel product from a steam stripper has to be further treated for water removal. This can be vacuum drying or coalescing, followed by salt drying. If this system is already in place, the steam stripping option would definitely have a lower capital cost compared to a fired reboiler. A steam stripped tower will also be a little smaller than a reboiled tower. If the product drying system is not in

place, the fired reboiler option should be considered. This will also create less sour water to process.

The product stripper material of construction is killed carbon steel with 410 SS trays and caps. Some part of the stripper column may need to be lined if the H₂S concentration is high. The overhead condenser and receiver are in wet sour service and HIC-resistant steel is required. The overhead receiver boot has a 1/4in corrosion allowance. R/F >0.10 is typical for both steam stripper and reboiled stripper.

Feed filtration

Even though distillate stocks are relatively clean, feed filtration is important to reduce exchanger/reactor plugging. A cartridge filter with 25 micron retention is typical for this application. Cracked feeds, if routed directly from the upstream facilities, should not be cooled.

Equipment delivery

Due to high worldwide demand, getting the necessary equipment delivered to the site could be a major challenge for any project. Construction and installation schedules are often governed by the delivery of long-lead items. Table 4 gives an estimate of current equipment and materials delivery.

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