CHAPTER 3

PROCESS DESCRIPTION

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1. PROCESS THEORY

Unicracking is carried out at elevated temperatures and pressures in a hydrogen atmosphere. Hydrogen partial pressure at reactor R-02 outlet is about 140 bar g and reactors R-01 and R-02 operating temperatures range from 380 - 450°C.

The Unicracking catalysts are formulated by compositing varying amounts of Group VIB and VIII metals on amorphous or zeolitic silica-alumina base.

Amorphous silica-alumina is viewed as a polymer of alumina on a backbone of silica. Crystalline synthetic silica-alumina, known as zeolite, is regarded as a copolymer of alumina and silica with ion exchange capacity.

1.1 HYDROCRACKING CHEMISTRY

Hydrocracking is essentially the acid catalyzed carbonium ion reactions of catalytic cracking coupled with hydrogenation reactions. C_4 fractions, rich in isobutane, are formed because of a great tendency to form tert-butyl-carbonium ions. Reaction of paraffins starts with formation of olefins at the metallic sites and the formation of carbonium ions from this olefin at the acidic sites. Extensive catalytic cracking followed by hydrogenation to form isoparaffins are the primary reactions. The rate of hydrocracking increases with the molecular weight of the paraffin.

$$C_{22}H_{46} \quad + \quad H_2 \quad \longrightarrow \quad C_{16}H_{34} \quad + \quad C_6H_{14}$$

A typical hydrocracking reaction of a cycloparaffin is known as a paring reaction, in which methyl groups are selectively removed from the cycloparaffin without severely affecting the ring itself. Normally the main non-cyclic product is isobutane. The hydrocracking of multiple ring naphthene such as decalin is more rapid than that of a corresponding paraffin. Naphthenes found in the product contain high ratio of methyl cyclopentane to methyl cyclohexane, far in excess of thermodynamic equilibria.

During hydrocracking of alkyl aromatics, isomerization, dealkylation, paring, hydrogenation and cyclization are the reactions that take place. In case of alkybenzenes, ring cleavage is almost absent and methane formation is at a minimum.

1.2 HYDROTREATING CHEMISTRY

1.2.1 SULFUR REMOVAL

a. (Mercaptan)

$$C - C - C - C - SH \quad + \quad H_2 \quad \longrightarrow \quad C - C - C - C \quad + \quad H_2S$$

b. (Sulfide)

$$C - C - S - C - C \quad + \quad 2H_2 \quad \longrightarrow \quad 2 C - C \quad + \quad H_2S$$

c. (Disulfide)

$$C - C - S - S - C - C + \quad 3 H_2 \quad \longrightarrow \quad 2 C - C \quad + \quad 2 H_2S$$

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d. (Cyclic Sulfide)

$$\text{(thiolane ring)} + 2H_2 \rightarrow \begin{cases} C-C-C-C + H_2S \\ C-C-C \ (\text{with } C \text{ branch}) + H_2S \end{cases}$$

e. (Thiophenic)

$$\text{(thiophene ring)} + 4H_2 \rightarrow \begin{cases} C-C-C-C + H_2S \\ C-C-C \ (\text{with } C \text{ branch}) + H_2S \end{cases}$$
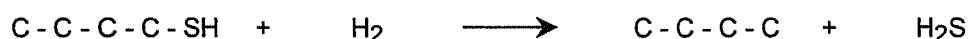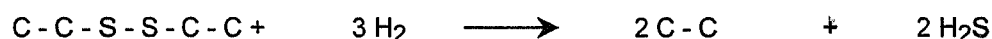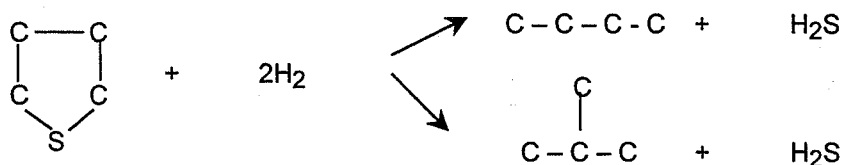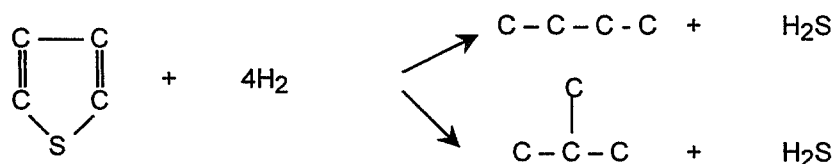
1.2.2   NITROGEN REMOVAL

a. (Pyridine)

$$\text{(pyridine ring)} + 5H_2 \rightarrow \begin{cases} C-C-C-C-C + NH_3 \\ C-C-C-C \ (\text{with } C \text{ branch}) + NH_3 \end{cases}$$

b. (Quinoline)

$$\text{(quinoline ring)} + 4H_2 \longrightarrow \text{(ring)}-C-C-C + NH_3$$

c. (Pyrrole)

$$\text{(pyrrole ring, N-H)} + 4H_2 \rightarrow \begin{cases} C-C-C-C + NH_3 \\ C-C-C \ (\text{with } C \text{ branch}) + NH_3 \end{cases}$$

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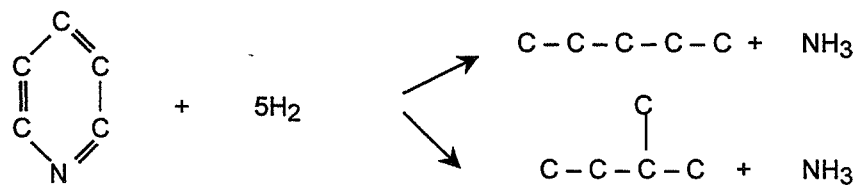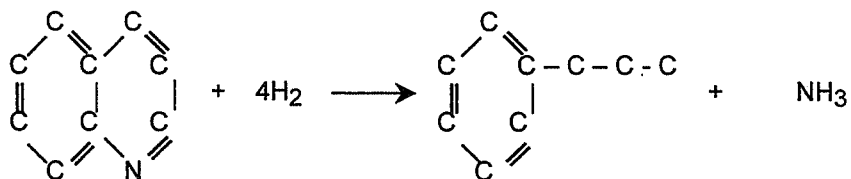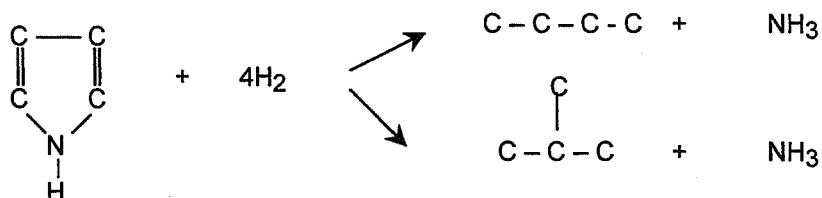
1.2.3   OXYGEN REMOVAL

a. (Phenols)
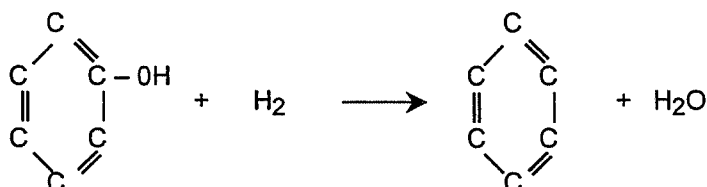
1.2.4   OLEFIN SATURATION

a. (Linear Olefin)

$$C - C = C - C - C - C \quad + \quad H_2 \quad \longrightarrow \quad C - C - C - C - C - C$$
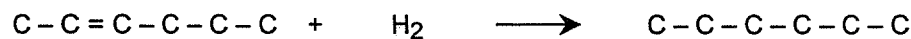
b. (Cyclic Olefins)
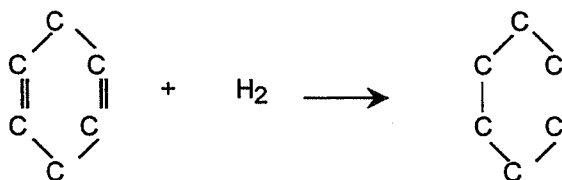
1.2.5   METALS REMOVAL

Almost all the metals removal occurs at the top bed of the first reactor. The mechanism of the decomposition of organo-metallic compounds is not well understood; however, it is known that metals are retained on the catalyst by a combination of adsorption and chemical reaction. The catalyst has a certain maximum tolerance for retaining metals. The useful life of the catalyst will be determined, to a large extent, by the amount of metals that are accumulated on it during the course of the operation.

Removal of metals from the feed normally occurs in plug flow fashion with respect to the catalyst bed. Typical organic metals native to most crude oils are nickel and vanadium. Iron can be found concentrated at the top of catalyst beds as iron sulfides which are corrosion products. Sodium, calcium and magnesium are due to contact of the feed with salt water or additives. Improper use of additives to protect fractionator overhead systems from corrosion or to control foaming account for the presence of phosphorus and silicon.

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1.2.6   HALIDES REMOVAL

Organic halides, such as chlorides and bromides, are decomposed in the reactor. The inorganic ammonium halide salts which are produced when the reactants are cooled are then dissolved by injecting water into the reactor effluent, as shown below.

$$HCl \;+\; NH_3 \;\longrightarrow\; NH_4Cl$$
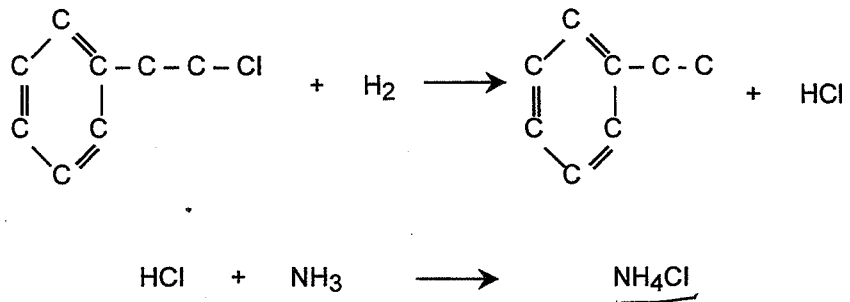
1.2.7   REACTION RATES

The approximate relative reaction rates for the three major hydrotreating reactions are:

Desulfurization	100
Olefin Saturation	80
Denitrification	20

The relative heats of reaction per unit of hydrogen consumption for a typical gas oil feed for these reactions are:

Desulfurization	10
Olefin Saturation	50
Denitrification	1

As can be seen from the above hydrotreating summary, desulfurization is the most rapid reaction taking place, but it is the saturation of olefins which generates the greatest amount of heat.

2.   DESCRIPTION OF FLOW

For reference, simplified schematic flow diagrams as well as PFDs of the HCK Unit are included at the end of this section.

2.1   REACTOR CIRCUIT

2.1.1   FRESH FEED AND LIQUID RECYCLE

The fresh feed is brought into the unit from:

♦ either the gas blanketed storage tank 51-TK-11 (this operation is called the "Cold Feed" operation),

♦ or directly from the Vacuum Unit (Unit 01) or the Delayed Coker Unit (Unit 11) or a combination of each (this operation is called "Hot Feed" operation).

"Hot Feed" and "Cold Feed" operations cannot happen simultaneously. In addition to these streams, a small quantity of Disulfide Oil is coming from the LPG Merox Unit (Unit 16).

The fresh feed is joined by the liquid recycle stream from the fractionation circuit (fractionator T-53 bottoms), and the combined feed (fresh feed plus liquid recycle) passes through filter F-01 to remove solid particles. The combined feed then enters the gas blanketed feed surge drum (V-02).

a. Feed Filter (F-01)

The feed filter is an automatic backwash type filter, specified to remove particulate matter of 20 microns and larger. The backwash sequence of the filter is started upon pressure drop switch signal (PDAH-005).

The backwash material is collected in the Backwash Surge Drum (V-11), pumped by P-05 A/B and cooled in the trim cooler E-13 before being sent to the battery limit on temperature control.

Filtering this material effectively removes the majority of the particulate matter that would accumulate in the reactor and cause pressure drop built-up.

b. Feed Surge Drum (V-02)

During "Hot Feed" operation, there is no control of the Feed Surge Drum level. Vacuum and coker gasoils are brought into the HCK Unit under flow control located inside their originating units. Only the V-02 level indication is sent to Units 01 and 11. In case of high level alarm in V-02, gasoil products from Units 01 and 11 have to be diverted in the storage area.

During "Cold Feed" operation, the flow of gasoil from storage is controlled by FIC-002 the setpoint of which is reset by drum V-02 level control LIC-004.

Switch LHS-004C allows to switch from "Hot Feed" to "Cold Feed" operations. In both operations, the liquid recycle is flow controlled from the fractionation to the Feed Surge Drum.

The Feed Surge Drum pressure is controlled by PIC-004 through split range action: either pressurization with hydrogen make-up or gas release to the flare system.

c. Charge Pumps (P-01 A/B)

From the bottom of the Feed Surge Drum, the combined feed flows to the suction of the reactor charge pump P-01. Minimum flow through this pump is ensured by controller FIC-005A acting on control valve FV-005 on spillback line from pumps discharge line back to the Feed Surge Drum.

The liquid feed is pumped by P-01 A/B at a controlled flowrate (FIC-008A) to the reaction section.

Because of the high pressure differential from discharge to suction, whenever the Charge Pump is shut down the feed control valves FV-008 and TV-019 as well as the on/off valve UV-026 are closed automatically in order to assure that a flow reversal from the reaction section back to the Feed Surge Drum do not occur via the pumps P-01 A/B miniflow (in case the two discharge check valves are not seated properly).

d. Liquid Feed to Reactors

After the feed control valve (FV-008), the liquid feed passes through feed-effluent exchangers (E-04 A/B and E-02). The feed stream passes through the shell side of the feed-effluent exchangers where they pick-up heat from the material heaving the reactor R-02. Exchanger E-02 outlet temperature is controlled by a line bypassing the exchangers E-04 A/B, via control valve TV-019.

A portion of the recycle gas is injected in the liquid feed in order to minimize the possibility of excessive fouling of the feed-effluent exchangers (the liquid feed should not be heated up above 370 °C prior to mixing with recycle gas). Also, the higher velocities in the exchangers provide better heat transfer.

2.1.2 HYDROGEN RECYCLE TO REACTOR

From the Recycle Gas Compressor (K-01), the recycle gas also passes through exchangers E-05, E-03 and E-01 to pick-up heat from the material leaving the reactor R-02. The exchanger E-01 outlet temperature is controlled by a line bypassing the exchanger E-03, via temperature control valve TV-009. The recycle gas then proceeds to the Recycle Gas Heater H-01 where it is further heated. The recycle gas is distributed through each of the ten passes of H-01 through flow control valves FV-009A thru J. The object is to maintain sufficient gas flow to each Recycle Gas Heater pass to avoid overheating the tubes. After the Recycle Gas Heater, the recycle gas joins the liquid feed upstream of the first reactor R-01.

The desired reactor R-01 inlet temperature of the combined recycle gas and feed is achieved via a cascade of the reactor inlet temperature (TIC-025) resetting the Recycle Gas Heater outlet temperature (TIC-024) which controls the firing of heater H-01.

2.1.3 REACTORS R-01 AND R-02

Once it has been heated to the desired reactor inlet temperature, the feed enters the top of the first reactor through an inlet distributor located in the top inlet flange of the reactor. Below the inlet distributor will be a vapor/liquid distributor tray. This tray is designed to evenly distribute both the gas and liquid over the cross-sectional area of catalyst below.

As the reactants flow downward through the catalyst bed, the chemical reactions described earlier (refer to § 1 Process Theory) occur and the temperature increases. In order to have close control of how much reaction is taking place and therefore how much heat is being generated, thermocouples are located in the radial center of the bed at the very top of the catalyst and then again at periodic levels down through the catalyst.

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Each level normally has three thermocouples with the exception of the bottom of the bed which has a multipoint thermocouple assembly (24 points). In addition to these bed temperature measurements skin thermocouples are provided on each catalyst beds.

The reactor R-01 is divided into three individual catalyst beds supported on a beam and grid support system. The support system is separated from the next bed of catalyst by a quench gas distributor, a reactant mixing chamber and a vapor / liquid redistributor tray. The reasons for separating a reactor into separate beds are the following:

a. If the gas and liquid flows become poorly distributed part way through the reactor, the catalyst will not be effectively utilized. By separating a reactor into multiple beds with redistribution trays in between, the reactants exiting one bed are redistributed evenly across the cross-sectional area of the next catalyst bed. In this way, should there be a problem with distribution in a bed, the catalyst in the lower beds will still be effectively utilized.

b. If the radial temperature profile becomes skewed part way through a reactor, the reaction rates will be different in different parts of the catalyst bed. This overworks the part of the catalyst that is hotter and under-utilizes the part of the catalyst bed that is cooler. It is also potentially hazardous if the hotter portion becomes significantly higher than the bulk temperature and forms a hot spot, especially if the hot spot is next to the reactor wall. Mixing of the reactants re-establishes thermal equilibrium and thus, after every quench zone, assures that all the catalyst is efficiently utilized.

c. In certain situations the heat of reaction will be large enough that the temperature increase across a reactor will be greater than the design. If this were allowed to happen, a reaction could become unstable and result in a temperature runaway. Therefore, cold recycle gas about 75°C is brought into the reactor at the interbed quench points in order to cool the reactants and thus control the reaction rate.

For this, the second bed and the third bed inlet temperatures are controlled by a TC → FC cascade on the flowrate of the recycle gas quench.

In order for the quench gas to be effective in cooling both the liquid and vapor reactants and to bring all reactants to thermal equilibrium, there must be good mixing in the quench zone. To do this, the quench gas is introduced through a spider-type distributor pipe to provide bulk cooling around the entire cross-sectional area of the reactor. The reactants then are swirled together in a mixing chamber to achieve thermal equilibrium of the liquid reactants. Finally, the reactants flow through a redistribution tray where thermal equilibrium is achieved between the vapor and liquids.

At the outlet of the first reactor, the reactants combine with quench gas to control the inlet temperature of the second reactor R-02 (cascade TIC-130 → FIC-013). At inlet of the second reactor, the reactants pass through an inlet distributor and a vapor/liquid distributor tray same as in the first reactor. The second reactor has two catalyst beds with one interbed recycle gas quench.

2.1.4 REACTOR EFFLUENTS COOLING AND SEPARATION

As the reactants leave the bottom of the second reactor R-02, they flow through the tube side of the feed-effluent and recycle gas exchangers E-01, E-02, E-03, E-04 A/B. Heat is given up to the feed and recycle gas as described earlier, before the reactants enter the Hot Separator V-03. A large portion of the heat generated in the reactors is recovered in the combined feed exchangers, thereby reducing the utilities cost of heating the feed up to the initial reaction temperature.

The hot separator vapor is cooled by exchange with the recycle gas stream (in E-05 A and B) and with the liquid from the Cold Flash Drum (V-06) in exchanger E-06. The vapor is then further cooled to about 60°C in the air cooled reactor effluent condenser E-07. As the reactor effluent vapors are condensed, ammonium bisulfide may be formed due to the presence of ammonia and hydrogen sulfide. Likewise, ammonium chloride may be formed due to the presence of ammonia and hydrogen chloride. To avoid risk of corrosion, wash water is injected before entering the air cooler E-07 to dissolve these salts and remove them from the system.

The wash water is a mixture of clean condensate (from battery limit) and water recovered in the Product Fractionator Receiver V-53. This wash water is injected into the reactor effluents by means of reciprocating pumps P-59 A/B. Clean condensate make-up is done under interface level control of the drum V-53 boot (LIC-037).

Air cooled exchanger E-07 is made up of eight banks of tube bundles, each with an inlet and outlet header to distribute flow to the individual tubes. Symmetrical inlet piping is used to obtain good distribution of water and effluent in each bank of tube bundles. In order to minimize the space requirements, the exchanger is multipass. At the end of each pass the flow is reversed to enter the next pass. This is done with a header box to facilitate cleaning.

The combined stream from the Hot Separator Vapor Condenser, now cooled to about 60°C, flows into the Cold Separator V-05 where sour water, hydrocarbon liquid and vapor fraction are separated and removed individually.

The Cold Separator V-05 is a horizontal vessel with a boot to collect the water phase. The liquid hydrocarbon and water are allowed to settle in the separator. A full diameter mesh blanket is installed to coalesce any water droplets and knock them out of the hydrocarbon phase. The water is removed from the separator boot on interface level control (LIC-011) and sent to the Cold Flash Drum V-06.

The Cold Separator V-05 is equipped with a contactor LSHH-045 that shuts down the Recycle Gas Compressor in case of high high level in the drum.

The pressure of the whole reaction loop is controlled at the level of the V-05 off-gas line. This pressure control is performed by PIC-013 acting on split range which enables:

♦ first to adapt the hydrogen make-up flowrate to the reaction loop by a control of the spillback valves PV-020 C/D, from the discharge line to the suction drum of the third stage of the Makeup Gas Compressors K-02 A/B/C,

♦ when spillback valves are fully open, to reset the flowrate set point (FIC-015) of the purge gas from the V-05 off-gas to the battery limit.

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2.1.5   **RECYCLE GAS**

After separation of the gas and liquid phases in the Cold Separator V-05, the gas leaves from the top of the separator and flows to the suction of the Recycle Gas Compressor K-01.

The Recycle Gas Compressor Suction Drum V-12 located at K-01 suction is isolated (blinded) in normal operation. It is lined up and operated only during catalyst regeneration.

This recycle compressor is driven by a high speed condensing steam turbine coupled directly to the compressor. Maximum energy is recovered from the steam by condensing the exhaust at an absolute pressure of about 100 mm Hg in surface condenser E-14. Non-condensable gases are removed from the condensate receiver by means of steam jet ejectors. Condensate is removed from the receiver and pumped by P-06 A/B to the refinery condensate collection system.

After the recycle compressor discharge, some recycle gas will be split off the main stream for use as quench gas between catalyst beds and reactors. Separate quench gas streams are used to reduce reactor interbed temperatures before each catalyst bed. Quench flows are regulated by flow controller cascaded from a temperature controller.

From after the Recycle Gas Heater H-01 until it returns to the Cold Separator, the gas flows along with the liquid through the reactor circuit in the same manner previously described.

2.1.6   **HOT AND COLD FLASH DRUMS CIRCUIT**

The Hot Separator V-03 liquid phase is routed to the Hot Flash Drum V-04 under level control LIC-006 or 007:

♦ valve LV-006A feeds the Power Recovery Turbine HT-01 which partially drives the Charge Pump P-01 A by means of a clutch. The additional power required for pump P-01 A is given by a motor which supplies the full power of the pump during the unit start-up or turbine HT-01 shutdown.

  Around 90% of the liquid hydrocarbon flowrate is expected to feed the turbine in normal operation. Valve LV-006A has a low pressure drop and is expected to be almost fully open during normal operation.

♦ The remaining liquid from V-03 flows through valve LV-006B (about 10%). A split range between valves LV-006A and LV-006B dedicates to valve LV-006B the control of the small liquid level variations to ensure, as far as possible, a constant flowrate through the turbine. This valve is an angle type valve due to the important pressure drop.

♦ In case of failure of the Power Recovery Turbine, all the liquid flowrate from V-03 is flowing through the angle type valve LV-006C (designed for 110 % of the total normal liquid flow). In this case, the level of V-03 is controlled by a split range on LV-006B and LV-006C, while LV-006A is closed.

The hydrocarbon liquid from V-03 is flashed into the Hot Flash Drum V-04 where dissolved hydrogen, H_2S and light hydrocarbon are flashed off:

♦ the flashed vapor from the Hot Flash Drum is cooled in the Hot Flash Vapor Condenser E-08 (air fin condenser) and then joins with the liquid hydrocarbon and sour water from the Cold Separator before entering the Cold Flash Drum V-06,

♦ the liquid from the Hot Flash Drum is routed to the fractionation section under level control LIC-008 or 009.

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The hydrocarbon liquid leaves the Cold Separator on level control LIC-012 and flows into the Cold Flash Drum V-06. The liquid is depressurized from Cold Separator pressure down to about 31 barg in the Cold Flash Drum. As it flows through the level control valve LV-012 A or B dissolved hydrogen, H_2S, and light hydrocarbons are flashed off:

♦ vapor is removed from the flash drum on pressure control (PIC-078) and sent to the battery limit (to Unit 12),

♦ the hydrocarbon liquid leaves the Cold Flash Drum on level control (LIC-015) and is heated up in exchanger E-06 before being sent to the fractionation section,

♦ sour water is removed from the flash drum boot on interface level control LIC-013 or 014 and sent to the battery limit (and further to sour water treating facilities in Unit 15).

2.1.7   MAKE-UP GAS

The makeup gas for the unit is a hydrogen rich gas (99.9% mole H_2) coming from the Hydrogen Plant (Unit 09). The hydrogen compressors K-02 A/B/C will then compress the gas (3 stages of compression) from the H_2 supply pressure (19.0 barg) up to the reactor circuit pressure (about 165 barg). From the discharge of the last stage of compression, the make-up gas joins the reactor effluents upstream the Hot Separator Vapor Condenser E-07.

The makeup compressors are reciprocating machines driven by motors. A suction drum is provided on the suction line of each stage of compression to knock out condensed liquids (drums V-07, V-08 and V-09). Trim cooler and air fin cooler are provided on the discharge of each stage to remove the heat generated by compression:

♦ air fin cooler E-09 and trim cooler E-10 at first stage discharge,

♦ air fin cooler E-11 and trim cooler E-12 at second stage discharge.

The flow of makeup gas through the compressor and into the unit is controlled by a complex system of pressure controllers on the Cold Separator and the interstage suction drums. The basic philosophy of the control scheme is to control the flow of gas as the demand for hydrogen dictates (as determined by the pressure in the Cold Separator). As hydrogen is consumed in the reactors, the pressure in the Cold Separator V-05 will start to decrease. The pressure controller PIC-013 will in turn call for more makeup gas by closing the control valves in the spillback line from the final stage discharge to suction (valves PV-020C and 020D). This will cause the suction pressure of the final stage to decrease, which then closes the spillback around the second stage of compression (valves PV-020A and 020B). This works in the same way back through each stage, with the net result that less gas is spilled back through each stage and more gas flows forward into the process.

In the event that the unit is calling for more makeup gas than is available (hydrogen consumption is too high), the pressure control system essentially works in reverse. The first stage suction drum pressure will start falling because there is not sufficient gas available from the Hydrogen Plant to replace what is being pumped out of the suction drum. When this happens, the pressure controller on the first stage suction drum senses the decrease in pressure and, in order to protect the first stage of the compressor from excessive compression ratio, overrides the second stage suction drum pressure controller and forces the first stage spillback valves to open (valves PV-019A and PV-019B). This spills back more gas back to the first stage suction drum to build the pressure back up. When this happens, the second stage suction drum pressure starts falling and its pressure controller opens the second stage spillback. The same thing happens to all stages of compression and the final result is that the Cold Separator pressure will start decreasing. At this point the operator will recognize that hydrogen consumption is exceeding supply and must be reduced either by decreasing reactor temperatures or feed rate or both. Alternatively, the hydrogen supply may be increased.

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High pressure in the First Stage Suction Drum V-07 will result in makeup hydrogen release, first to the fuel gas network via PV-017 A and last to the HP flare network via PV-017B (split range control).

Each compressor stage is equipped with double spillback valves with the following purpose:

♦ small spillback valves PV-019A, PV-020A and PV-020C, for normal pressure control of each discharge stage,

♦ spillback valves PV-019B, PV-020B and PV-020D, each being able to handle 100% of the compressed gas of the corresponding stage.

2.2 FRACTIONATION SECTION

The feed to the fractionation section, made of hydrocarbon liquids from the Hot Flash Drum and from the Cold Flash Drum, will contain everything from hydrogen and hydrogen sulfide through the heaviest components, and it is necessary to separate these streams into the desired products as light gases, LPG, gasoline, kerosene and diesel oil.

This feed is first sent to a steam Stripper T-51, located upstream of the Product Fractionator, whose purpose is to remove all the H_2S from the Product Fractionator feed:

♦ the stripper overhead liquid is sent to the Debutanizer T-56 where LPG and light gases are recovered in the overhead,

♦ the stripper bottoms is heated in the fired heater H-51 and then routed to the Product Fractionator T-53.

Sour LPG from the Debutanizer Receiver V-55 is sent to the Amine Absorber column T-101 in order to remove H_2S before sending it to the LPG Caustic Washing in Unit 12.

The naphtha product from the fractionator overhead joins with part of the stabilized naphtha from Debutanizer bottom as sponge oil into the Sponge Absorber in order to maximize C_3 and C_4 recovery on Stripper and Debutanizer overhead gases.

The product Fractionator bottoms is split into two streams, one being recycle to the Reaction Section, the other one (unconverted oil) being sent to storage as fuel oil.

2.2.1 STRIPPER COLUMN (T-51)

The Stripper T-51 feeds, Hot and Cold Flash Drum liquids are flowing to the column under level control respectively LIC-008 or 009 and LIC-005 on these drums.

The stripping is achieved by medium pressure steam injection under flow control (FIC-025) below the bottom tray N° 29, to remove all H_2S from the Product Fractionator feed. The overhead from the Stripper is partially condensed in the air cooled fin tube condenser (E-51) and the trim condenser (E-52). The effluent the condenser E-52 is separated into vapor and liquid phases in the overhead receiver V-51:

♦ Vapor phase (butanes and lighter) is sent to the Sponge Absorber T-52.

♦ Liquid hydrocarbon phase is pumped by Stripper Overhead Pumps P-51 A/B. Main part of this liquid is used as reflux for the stripper on flow control (FIC-027) reset by the stripper overhead vapor temperature controller (TIC-255). The remaining part is sent to the Debutanizer Column T-56 on flow control (FIC-028) reset by the control of the level of the Stripper Receiver V-51 (LIC-024).

♦ The sour water is removed from the receiver boot on interface level control (LIC-023) and sent to the battery limit (and further to the Sour Water Stripper - Unit 15).

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Pressure control of the Stripper (PIC-045) is performed by degassing the Sponge Asborber Receiver via control valve PV-045.

To prevent from corrosion in the overhead line of the Stripper, corrosion inhibitor is injected by means of dosing pumps P-62 A/B. This inhibitor is mixed with hydrocarbon liquid from the reflux pumps and sent to the stripper overhead line.

The stripper bottoms material leaves the bottom of the stripper and is divided into the four passes of the Product Fractionator Feed Heater H-51 which are flow controlled (FIC-026 A thru D) reset by the stripper bottom level (LIC-022).

2.2.2　SPONGE ABSORBER (T-52)

Vapor phases (butanes and lighter) from the overhead receivers of the Stripper and the Debutanizer (respectively V-51 and V-55) are feeding the Sponge Absorber below the bottom tray N° 15.

Lean oil, a mixture of the stabilized naphtha from T-56 and the Product Fractionator overhead liquid, is cooled in the Lean Oil Cooler E-54 before entering the Sponge Absorber above the top tray. The temperature of the lean oil is controlled by a bypass of the exchanger E-54 via valve TV-372.

The overhead vapor from the Absorber is partially condensed in the Sponge Absorber trim condenser E-53. Effluent from this condenser is separated into vapor and liquid phases in the overhead receiver V-52:

♦ the vapor phase is removed from the receiver on pressure control (PIC-045), and sent to the unit battery limit (to unit 12),

♦ the hydrocarbon liquid phase is pumped by the Sponge Absorber Reflux Pumps P-53 A/B and integrally returned to the top of the Sponge Absorber (total reflux) on receiver V-52 level control LIC-027,

♦ the sour water is slowly accumulating in the receiver boot and removed periodically by means of a manual valve.

The Sponge Absorber bottoms material is pumped by P-52 A/B under level control (LIC-025) and mixed with the Stripper overhead liquid as feed to the Debutanizer T-56.

2.2.3　DEBUTANIZER (T-56)

The debutanizer feed is preheated in the Debutanizer Feed-Bottoms Exchanger E-65 A/B/C before entering the Debutanizer above tray N° 21.

The overhead from the Debutanizer is partially condensed in the water trim cooler E-67. The effluent of the condenser is separated between vapor and liquid phases in the overhead receiver V-55:

♦ the vapor phase (butanes and lighter) leaves the receiver as feed to the Sponge Absorber on pressure control to regulate the Debutanizer overhead pressure (PIC-059),

♦ the liquid hydrocarbon phase from the receiver is split in two streams:

- one stream is pumped by the Debutanizer Overhead Pump P-61 A/B and refluxed to the Debutanizer top tray under flow control (FIC-056) reset by the overhead receiver level control (LIC-040),

- the other stream, the net LPG production, is sent to the LPG Amine Absorber T-101 on flow control (FIC-057) reset by the sensitive tray N° 5 temperature controller (TIC-295).

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The control of the cut point between LPG and stabilized naphtha is performed by this temperature control in the upper part of the column,

♦ the sour water is slowly accumulating in the receiver boot and removed periodically by means of a manual valve.

To prevent from corrosion in the overhead line of the Debutanizer, corrosion inhibitor is injected by means of dosing pumps P-63 A/B. This inhibitor is mixed with hydrocarbon liquid from the reflux pumps and sent to the Debutanizer overhead line.

The stripping vapors in the lower portion of the column are provided by process heat exchange in the Debutanizer Reboiler E-68. The heating medium (diesel oil pumparound) is flow controlled (FIC-072) to achieve a steady duty of the reboiler E-68. The stabilized naphtha leaves the bottom of the Debutanizer on level control (LIC-039), is partially cooled in the feed-bottoms exchanger and can be sent to two possible destinations:

♦ to the Naphtha Hydrotreater (Unit 02) when it is possible. The following signals give information concerning the availability of the Unit 02:

  ■ 02-XL-001 A/.B: status of the Naphtha Hydrotreater feed pumps 02-P-01 A/B,

  ■ 02-LIC-001: level of the Naphtha Hydrotreater Feed Surge Drum 02-V-01.

♦ to the storage area, after further cooling in the Debutanizer Bottoms Cooler E-69 A/B, when Unit 02 is not available.

2.2.4   LPG AMINE ABSORBER (T-101)

The sour LPG coming from the debutanizer receiver V-55 is sent to the amine washing section under flow control FV-057, in order to decrease the H_2S content from 26.6 mol percent down to 500 ppm mol.

Sour LPG in mixed phase at 9 barg is vaporized in the LPG Vaporizer E-101. The LPG vaporization rate is controlled by a level control in the kettle (LIC-081) cascading the LP steam flow rate controller (FIC-084).

Vaporized LPG is then superheated up to 60°C in the LPG Heater E-102 to avoid LPG condensation in the absorber. The temperature at the heat exchanger outlet is controlled (TIC-375) by adjusting the vapor flow rate in the exchanger.

Vaporized LPG is fed to a knock-out drum located at the column T-101 bottom and then sent to the column to remove H_2S.

The amine absorber T-101 operates at a pressure of 8.3 barg. Lean amine cooled down to 65°C in the Amine Cooler E-103, is fed above the column top tray on flow control located on the lean amine supply line (FIC-086).

Rich amine withdrawn from the column bottom is sent to the Amine Regeneration Unit (Unit 14) on level control (LIC-083).

Washed gas are sent to the Washed Gas Knock-out Drum V-101 integrated to the absorber T-101 top in order to settle amine droplets that may have been carried away in the gas phase. The level of this drum, which increases slowly, is ON/OFF controlled by valve LV-084.

The absorber overhead gas are condensed in the trim condenser E-104, and collected in the Washed LPG Surge Drum V-102 equipped with a boot for separated water removal:

♦ The pressure of this reciever V-102 is controlled by means of split range action on the hot by-pass of the condenser E-104 (valve PV-095B) and on a gas relief to the Unit 12 (via valve PV-095 A).

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♦ Washed LPG are pumped under level control (LIC-086) with LPG Feed Pump P-101 A/B to Unit 12 for caustic washing and LPG recovery.

2.2.5   PRODUCT FRACTIONATOR (T-53 / 54 AND 55)

The fractionator feed is heated up in the heater H-51 before it enters the lower portion of the column T-53, under the tray N° 39. The heater is operated to provide the desired flash zone temperature (around 377°C) to achieve the lift required with the column operating at about 1 barg in the flash zone.

The Product Fractionator is a steam stripped column with two sidecut drawoffs (kerosene and diesel).

a. Overhead Section of T-53

The overhead vapor from the Fractionator is completely condensed in the air cooled fin tube condenser E-64. The effluent of this condenser E-64 is separated in the Product Fractionator Receiver V-53.

The overhead receiver V-53 pressure is controlled by PIC-055 through a split range action, either venting a small amount of gas to the flare via PV-055B or keeping a nominal 0.35 barg pressure of fuel gas on the drum.

■ The overhead temperature from the column controls the amount of liquid being refluxed from the receiver V-53 back to the column (cascade TIC-270 → FIC-031) via reflux pumps P-64 A/B, and thus controls the end point of the overhead product.

■ The liquid overhead material is pumped by the Product Fractionator Net Overhead Pump P-58 A/B on flow control (FIC-033) reset by receiver V-53 level controller (LIC-038). This product is used as sponge oil in the Sponge Absorber T-52.

■ The sour water (condensed stripping steam) of the receiver boot is pumped by P-59 A/B and sent to the reaction section as wash water (upstream of air cooler E-07). The receiver boot interface level is controlled (LIC-037) by Cold Condensate makeup.

b. Kerosene Circuit

The Fractionator kerosene sidecut flows to the Kerosene Stripper T-55 top tray on level control (LIC-035). Here, the kerosene flash point is adjusted by stripping off the light ends with the vapor from thermosiphon reboiler E-59, which is heated by the Product Fractionator bottoms product. The heating medium is flow controlled (FIC-071) in order to achieve a steady duty of the reboiler E-59. The stripped light ends from the top of the Kerosene Stripper are returned to the main column.

The stripped kerosene is pumped under flow control (FIC-049) by the Kerosene Product Pumps P-57 A/B, cooled in exchangers E-58 A/B/C, in the air cooled fin tube cooler E-60 and then in the trim cooler E-61 before being sent outside the unit battery limit to the storage area.

c. Gasoil Circuit

The Fractionator diesel oil sidecut flows to the Diesel Stripper T-54 top tray on level control (LIC-032). Here, the diesel oil flash point is adjusted by stripping off the light ends with the vapor from low pressure stripping steam. This stripping steam injection is flow controlled in order to achieve a steady operation of the Stripper (FIC-045). The stripped light ends from the top of the Diesel Stripper are returned to the main column.

Stripped diesel oil is pumped under flow control (FIC-048) by Diesel Product Pumps P-56 A/B. It is first cooled in the LP Steam Generator E-70, then in the air cooled fin tube cooler E-62 and final cooling is performed in the Diesel Product Trim Cooler E-63. Diesel oil flows through a Diesel Coalescer V-56 and a Salt Drier V-57, in order to remove all its free water, before being sent outside the unit battery limit to the storage area.

d. Gasoil Pumparound

Part of the diesel oil withdrawn from the chimney tray of the fractionator is pumped by the Diesel Pumparound Pump P-55 A/B.

The pump P-55 A/B discharge is split in two flows:

- the first flow is returned to the column as internal reflux under flow control (FIC-035) reset by the chimney tray level controller (LIC-028),

- the other stream is flow controlled to heat exchangers E-68 and E-55 to be cooled and returned to the column T-53. Heat released by diesel pumparound is used for the Debutanizer reboiling in E-68 and for generation of MP steam in E-55.

e. Fractionator Bottoms Product

The main fractionator stripping steam is injected into the column just below the bottom tray (N° 49). This stripping is flow controlled (FIC-037) to adjust the flash point of the Fractionator bottoms product.

The nominal 368°C plus material is removed from the bottom of the Fractionator and pumped by the Product Fractionator Bottoms Pumps P-54 A/B.

This flow controlled stream (FIC-043 and FIC-044) is used to provide process heat for the Kerosene Stripper Reboiler E-59 and to generate MP steam in exchanger E-56 before being split in two streams:

- one stream, called "unconverted oil" is cooled in the air cooled fin tube cooler E-57 before being sent, under flow control (FIC-043), to the storage area,

- the other stream is returned to the reactor section on flow control (FIC-044) as liquid recycle. This liquid recycle temperature is controlled (TIC-278) by injection of cold unconverted oil via control valve TV-278.

2.2.6   **STEAM GENERATION SYSTEMS**

The boiler feed water, delivered at about 22 barg and 116°C, is first preheated in the Kerosene Product / BFW exchanger E-58 and then split in two streams for low pressure and medium pressure steam generation.

Low pressure steam is generated in the LP Steam Generator E-70. The boiler feed water is injected into the kettle on flow control (FIC-047) reset by the output signal of the summing function FY-046 (using the kettle level controller (LIC-033) and indication of the flow of LP steam generated (FI-046)). This LP steam is superheated in the convection zone of the Product Fractionator Feed Heater H-51 before being split in two streams:

♦ one stream is directly used as stripping steam in columns T-53 and T-54,

♦ the excess is partially desuperheated in DS-51 on temperature control (TIC-260) before being routed to the LP steam network of the unit.

Medium pressure steam is generated utilizing the heat withdrawn from the diesel pumparound (in E-55) and from the fractionator bottoms product (in E-56). The boiler feed water is injected into the Steam Disengaging Drum on flow control FIC-038 reset by the output signal of the summing function FY-039 (using the drum level controller (LIC-031) and the indication of the flow of MP steam generated (FI-039)). The BFW is pumped from the steam drum by the pumps P-65 A/B and circulated through exchangers E-55 and E-56 on flow control (respectively FIC-041 and FIC-042). This MP steam is superheated in the convection zone of the Product Fractionator Feed Heater H-51 before being split in two streams:

♦ one stream is directly used as stripping steam in the Stripper Column (T-51),

♦ the excess is partially desuperheated in DS-52 on temperature control (TIC-263) before being routed to the MP steam network of the unit.

Both Steam Generator E-70 and Steam Disengaging Drum V-58 are equipped with one continuous and two discontinuous boiler blowdown connections. The continuous blowdown water is collected in the Continuous Blowdown Drum V-60. Flashed steam is routed to the LP steam header and the water is flowing on level control (LIC-030) to the Intermittent Blowdown Drum V-59 which also collects the intermittent blowdown of E-70 and V-58.

Phosphate is injected by means of dosing pumps (PK-01 / P-01/P-02) in the Steam Generator E-70 and in the Steam Disengaging Drum V-58. The purpose of the phosphate injection is to form a protective layer on the internal surface of the boilers (chemical precipitate).

The phosphate concentration (Na_3PO_4) required in the boiler is typically ranging from 30 to 50 mg/l, depending on the operating conditions of the steam generation (pressure and temperature), and is measured in the continuous blowdown stream.

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2.3 **ATTACHED UNIT PFDS FOR REFERENCE**

Enclosed schematic flow diagrams of the reaction section and the fractionation section of the HCK unit.

♦ Figure 4.1 - Unicracking Process Single Stage Unit - Reactor Circuit

♦ Figure 4.3 - Unicracking Process - Fractionator circuit (fractionator first)

As well as UOP PFDs of the unit showing the normal operating conditions :

♦ UOP PFDs 555027-110-01 thru 05

♦ UOP PFDs 555028-110-01 thru 05

Figure 4.1

Unicracking Process Single Stage Unit
Schematic Flow Diagram
Reactor Circuit

THIS SCHEME IS VALID
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OF MIDOR REFINERY ONLY

Flash Gas
to Treating
and
Hydrogen
Plant

To Product
Fractionation

To Product
Fractionation

CFD

SW

CS

WW

H$_2$

HFD

HS

Charge
Stock

Recycle
Liquid

Legend
HS = Hot Separator
CS = Cold Separator
HFD = Hot Flash Drum

CFD = Cold Flash Drum
WW = Wash Water
SW = Sour Water
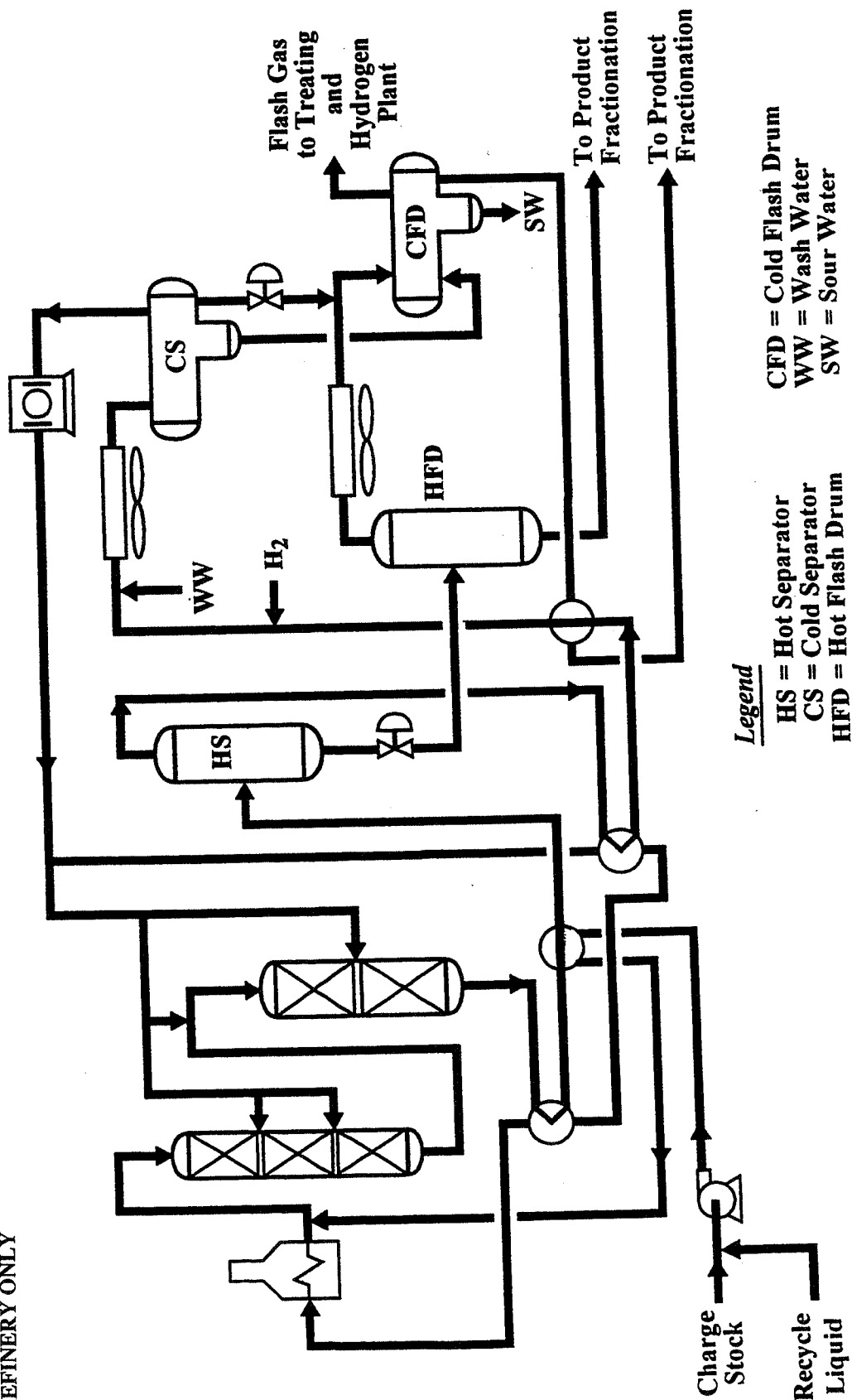
Figure 4.3

Unicracking Process
Schematic Flow Diagram – Fractionation Circuit
(Fractionator First)

THIS SCHEME IS VALID
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Sponge
Absorber

Off
Gas

Debutanizer

LPG
Product

Lean
Amine

UOP 2032-3
UOP 2956-2

Rich
Amine

LPG
Amine
Absorber

Heat

Naphtha
Product

Product
Fractionator

Heat

Kerosene
Product

Stripping Steam

Diesel
Product

Stripping Steam

Unconverted
Bottoms
to Storage

Heat

Stripping Steam

Stripper

Stripping Steam

Fractionator
Charge Heater

Liquid Recycle
to Reactor Circuit

Flash
Drum
Liquid
from
Reactor
Circuit
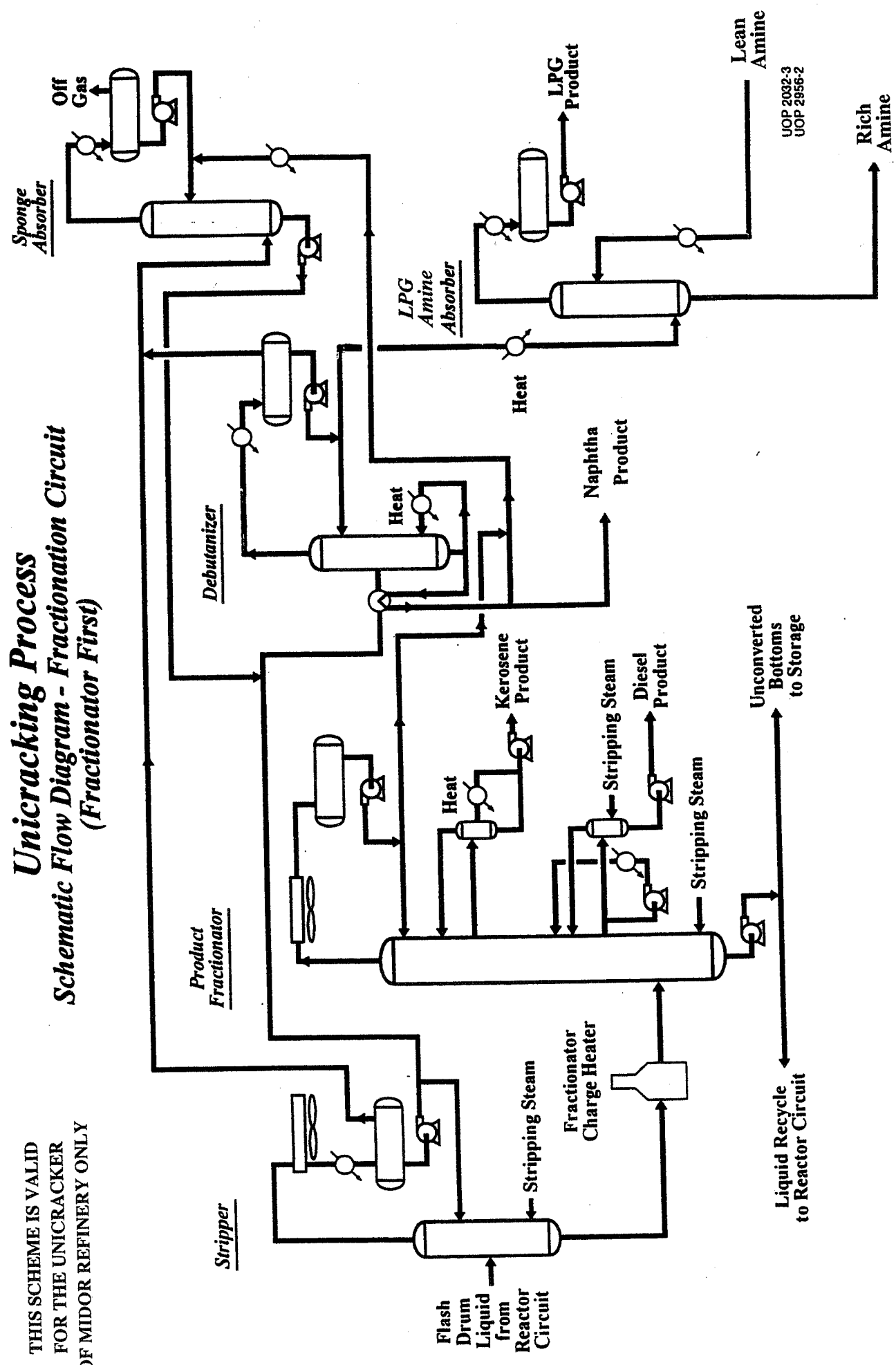
3. PROCESS VARIABLES

The major operating variables and their effect on product quality and life of the catalyst are presented in the following paragraphs.

The most important part of the Unicracking Process is the catalyst system. The amount of conversion which takes place in the reactors is going to be determined by several variables: the type of feedstock, the amount of time the feed is in the presence of catalyst, the partial pressure of hydrogen in the catalyst bed, and most important, the temperature of the catalyst and reactants. The obvious generalization about temperature is that the higher the temperature, the faster the rate of reaction and therefore, the higher the conversion.

3.1 REACTORS

3.1.1 REACTORS TEMPERATURE

The reactor inlet temperature is controlled by the operator to adjust the amount of conversion. The reactor outlet temperature is a function of the feed quality and cannot easily be varied except by changing the reactor inlet temperature. The inlet temperature must always be controlled at the minimum required to achieve the desired conversion.

During the course of an operating cycle, the catalyst should deactivate continuously and gradually due to coke formation. This loss of catalyst activity can be compensated by increasing reactors inlet temperatures up to a limit of about 420°C, the maximum bed temperature, above which coke formation becomes very rapid and little improvement in performance is expected. The design temperature of the reactors and the Recycle Gas Heater will also determine the maximum allowable operating values. The temperature rise across the reactors must be monitored continuously in order to ensure that the design limitation of the unit is not exceeded.

The close relationship between LHSV and operating temperature makes it necessary to reduce temperature before lowering the feed rate (LHSV), and to increase the feed rate before raising the temperature.

3.1.2 CATALYST DEACTIVATION

Loss of catalyst activity can be caused by several mechanisms. Under normal circumstances, the catalyst should deactivate continuously and gradually throughout the entire run period due to coke formation. However, several mechanisms exist for deactivation which can cause unusually high deactivation rates.

a. Coke Formation

Coke formation can occur from the high molecular weight compounds contained in the feed or from the condensation reaction of polynuclear aromatics into high molecular weight, low hydrogen content coke. During normal operation, the high partial pressure of hydrogen and the hydrogenation activity of the catalyst inhibit coke formation due to condensation reactions. The formation of coke from high molecular weight materials in the feed is limited by controlling the quality of the fresh feed.

The heptane insolubles test is used to measure undesirable high molecular weight hydrocarbons. This test reports the weight percent of feed which is insoluble in normal heptane (0.05 wt% for the vacuum and coker gasoils). In general, this material has an average molecular weight of about 5,000, and a major portion of it is asphaltenes with molecular weights above 5,000.

b. Metals Poisoning

Organo-metallic compounds will be decomposed and retained on the catalyst. Calcium, magnesium and alkali metals, such as sodium, can accumulate on the catalyst from inadequately desalted feedstock or contact of feedstock with salt water and additives. Other inorganic metals, such as silicon and phosphorus due to improper use of refining additives can also attribute to metal poisoning. These metals are **non-regenerable** poisons to the catalyst. By limiting the total metals in the feed the amount of metals deposited on the catalyst can be controlled.

c. Ammonia

Organic nitrogen compounds in the feed are converted to ammonia. Since ammonia is a basic compound, it competes with the normal reactants for the acid sites on the catalyst and inhibits the activity of the catalyst. However, the majority of the ammonia is removed from the reactor circuit by the reactor effluent wash water and, therefore, does not appreciably affect catalyst deactivation. In the event that wash water injection should be stopped for any period of time, the ammonia will concentrate in the recycle gas returning to the reactors. This in turn will result in a sudden, sharp loss in catalyst activity. The activity will return to normal when the wash water is reinstated. The catalyst temperatures should not be increased while water was stopped to compensate for loss in activity.

d. Severity of Operation

Any increase in severity (i.e., LHSV, conversion level, feedstock quality, etc.) above that for which the unit was designed will result in an increased rate of catalyst deactivation - mainly due to a higher rate of coke formation. For that reason, as well as possible heat balance and mechanical limitations, it is not recommended to operate above design conditions unless a reduced catalyst life can be tolerated.

3.2　FRESH FEED QUALITY

The quality of the raw oil charged to the Unicracker will affect the temperature required in the catalyst bed to reach the desired conversion, the amount of hydrogen consumed in the process, the length of time between catalyst regenerations, and the quality of some products. The effect of the feedstock quality on the performance of the unit is important and should be well understood especially with regard to contaminants which can greatly reduce the life of the catalyst.

3.2.1　SULFUR AND NITROGEN COMPOUNDS

In general, increasing the amount of organic nitrogen and sulfur compounds contained in the feed results in an increase in severity of the operation. The sulfur content of the feed for a normal fresh feed (vacuum + coker gasoils) can vary up to as high as 2.5 to 3.0 wt percent.

The organic nitrogen compounds are converted to ammonia which, if allowed to build up in the recycle gas, competes with the hydrocarbon for the active catalyst sites. This results in a lower apparent activity of the catalyst as the ammonia concentration increases. Because of this, feedstocks with high organic nitrogen contents are more difficult to process and require higher catalyst temperatures.

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3.2.2   UNSATURATED COMPOUNDS

The amount of unsaturated compounds (such as olefins and aromatics) contained in the feed will have an effect on the heat release during the reaction and on the total hydrogen consumption on the unit. In general, for a given boiling range feedstock, a reduction in API gravity indicates an increase in the amount of unsaturated compounds (and, therefore, higher heat of reaction and higher hydrogen consumption).

Large amounts of unsaturated hydrocarbons can also cause a heat balance problem if the unit has not been designed to process this type of feed.

3.2.3   BOILING RANGE

Increasing the boiling range usually makes the feed more difficult to process which means higher catalyst temperatures and shorter catalyst life. This is especially true if the feed quality is allowed to decrease significantly due to entrainment of catalyst poisons in the feed. Higher endpoint feeds also usually have higher sulfur and nitrogen contents which again make it more difficult to process.

3.2.4   CRACKED FEED COMPONENTS

Cracked feedstocks derived from the Delayed Coker Unit tend to have higher contaminants such as sulfur, nitrogen, and particulates. They are also more refractory, with high aromatics content and polynuclear aromatic precursors. These compounds make cracked stocks harder to process to produce quality distillates and generally show an increase in HPNA formation. The hydrogen makeup requirement increases since hydrogen consumption is higher when cracked feedstock is processed. Increase in catalyst deactivation rate is expected from inclusion of cracked feedstocks.

3.2.5   PERMANENT CATALYST POISONS

Organo-metallic compounds contained in the feed will be decomposed and the metals will be retained on the catalyst, thus decreasing its activity. Since metals are normally not removable by oxidative regeneration, once a catalyst has been poisoned by metals, its activity **cannot be restored**. Therefore, metals content of the feedstock is a critical variable which must be carefully controlled. The particular metals which exist in Vacuum gas oil and Coker gas oil feeds are naturally occurring nickel, vanadium and arsenic as well as some metals which are introduced by upstream processing or contamination such as lead, sodium, silicon and phosphorous. Iron naphthenates are soluble in oil and will be a poison to the catalyst. Iron sulfide as corrosion product is normally not considered a poison to the catalyst and is usually omitted when referring to total metals.

The tolerance of the catalyst to metals is difficult to quantify and is somewhat dependent upon the type of catalyst being employed and the severity of the operation, i.e., the higher the severity, the lower will be the metals' tolerance since any impairment of activity will affect the ability to make the desired conversion. In any event, it is recommended to keep the total metals in the feedstock as low as possible and never over the design maximum of 2 wt-ppm. It should also be pointed out that feed metals analyses should not be averaged (i.e., 10 wt-ppm metals for one day is not equal to 1 wt-ppm metals for ten days). The reasons are:

a. High metals analyses usually indicates entertainment of asphaltenic hydrocarbons in the vacuum gas oil since metals are not normally volatile. This means that along with the metals, the feed is bringing in hydrocarbons which are very difficult to hydrocrack, which can contain high molecular weight organic nitrogen and sulfur compounds, and which have a high coke forming tendency, thus leading to rapid coke deactivation of the catalyst.

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b. Processing high metals content (and highly asphaltenic) feeds for a short period of time causes the catalyst to deactivate to some degree (depending on the amount of contaminants). Higher catalyst temperatures are then required to make the desired conversion. This higher temperature results in faster coke formation for the remainder of the run even though the metals are reduced back to an acceptable level.

3.2.6   REGENERABLE CATALYST CONTAMINANTS

Even though the formation of coke is the normal deactivation mechanism and is expected, the quality of the feed can affect the rate of formation and, therefore, cause unacceptably short catalyst life between regenerations if not controlled.

Large quantities of asphaltenic hydrocarbons brought in with the feed can cause extremely rapid coke deactivation. Therefore, one of the major control variables of the Unicrackers is the amount of asphaltenes in the feed as measured by the <u>heptane insolubles test</u>. The heptane insolubles should not be allowed to exceed 0.05 wt percent and should be kept as low as possible. Another test which is valuable in controlling feedstock quality is the Conradson carbon residue test (ASTM D-189 or D-4530) which measures the carbon forming properties of oil under controlled conditions. The Conradson carbon analysis should not be allowed to exceed design values and should be kept as low as possible.

It is advisable to continuously monitor both the feed for heptane insolubles and Conradson carbon to ensure normal deactivation rates.

If it is found that the feed is high in asphaltenes or metals, the source of contamination should be located and eliminated.

Therefore, if the vacuum unit is being run significantly over the design capacity or on a substantially different crude, problems can be encountered with the ability to produce both a clean VGO and specification asphalt at the same time.

3.2.7   PARTICULATE MATTER

Particulate matter in the feed, such as corrosion products from upstream units, would limit the length of the operation by fouling the top of the catalyst bed and increasing the reactors pressure drop. A feed filter is provided to remove most of the solid particles of 20 microns and larger that may be brought in with the feed.

3.3   FRESH FEED RATE (LHSV)

The amount of catalyst loaded into the reactors is based upon the quantity and quality of design feedstock and the desired conversion level. The variable which is normally used to relate the amount of catalyst to the amount of feed is termed **liquid hourly space velocity** (LHSV). LHSV is the ratio of volumetric feed rate per hour to the catalyst volume.

$$LHSV = \frac{Volume\ of\ Feed\ per\ Hour}{Volume\ of\ Catalyst\ in\ Reactors}$$

The LHSV is measured in h^{-1}, and the volume of fresh feed is based on the specific gravity of the feed at 15°C.

The Unicracking Unit is designed for a LHSV which depends on the severity of the operation. Increasing the fresh feed rate with a constant catalyst volume increases the LHSV and a corresponding increase in catalyst temperature will be required to maintain a constant conversion. The increased catalyst temperature will lead to a faster rate of coke formation and, therefore, reduce the length of time between regenerations. If the LHSV is run significantly higher than the design of the unit, the rate of catalyst deactivation may become unacceptable.

3.4 LIQUID RECYCLE

The Unicracking Unit is designed to recycle unconverted feed from the product fractionator bottoms back to the reactors. This stream, normally a TBP 368°C plus material, is termed liquid recycle. The liquid recycle rate is normally adjusted as a ratio with fresh feed.

This variable is called combined feed ratio (CFR), and is defined as follows:

$$CFR = \frac{Fresh\,Feed\,Rate \;+\; Liquid\,Recycle\,Rate}{Fresh\,Feed\,Rate}$$

For this unit the CFR is about 1.5.

The reasons for recycling unconverted material back to the reactors are as follows:

1. Even though the liquid recycle is unconverted feed, it is a very stable reactant to put back to the catalyst because it will have been totally desulfurized and almost completely saturated. Therefore, it will only generate the amount of heat liberated from hydrocracking. For this reason, it is used as a heat sink, that is increased mass flow through the catalyst bed to carry heat away, thus limiting the temperature rise across the catalyst. Without liquid recycle in some cases the temperature rise across the catalyst would be excessive to the point where reaction would be difficult to control.

2. A second important function of liquid recycle is to reduce the severity of the operation. This can be shown by considering conversion per pass which is defined as follows:

$$Conversion\,per\,pass = \left(\frac{Fresh\,Feed\,Rate \;-\; Frac\,Bottoms\,Rate\,to\,Storage}{Fresh\,Feed\,Rate \;+\; Liquid\,Recycle\,Rate}\right) \times 100$$

For this unit, the conversion per pass is about 66%.

As the CFR increases, the conversion per pass decreases. This is also seen that the catalyst temperature requirement is reduced as the CFR is increased (at a constant fresh feed conversion level). Therefore, reducing the CFR below the design value can lead to higher catalyst temperatures and shorter life between regenerations. Increasing the CFR above design can be helpful when operating at low fresh feed rates since it does not allow the total mass flow through the catalyst bed to reach such a low value that poor distribution patterns are established.

3. The rate of liquid recycle also has a small effect on the distribution of products from a given feedstock. The directional effect of increasing CFR will be:

   a. Less yield of naphtha and increased yield of total 150°C plus material.

   b. Of the total 150°C plus material, a greater percentage will be 260 - 370°C diesel.

Normally the range of combined feed ratio which can be run on the HCK unit will be so small that only marginal changes in yield structure will be possible by varying CFR, and, therefore, it is not used as a control variable.

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3.5   **HYDROGEN PARTIAL PRESSURE**

The reactor section operating pressure is controlled by the pressure that is maintained at the Cold Separator V-05. This pressure, multiplied by the hydrogen purity of the recycle gas, determines the partial pressure of hydrogen at the separator. The hydrogen partial pressure required for the operation of the unit is chosen based on the type of feedstock to be processed and the amount of conversion desired. The function of hydrogen is to promote the saturation of olefins and aromatics and saturate the cracked hydrocarbons. It is also necessary to prevent excessive condensation reactions from forming coke.

1. Running the unit for extended periods of time at lower than design partial pressure of hydrogen will result in increased catalyst deactivation rate and shorter time between regenerations. This may also affect products purity, especially those properties associated with aromatics reduction, such as kerosene smoke point or diesel cetane index.

2. Operating at higher than design pressure would favour catalyst life but may not be possible. There is a practical equipment limitation which will not allow significantly higher pressure than design, such as the pressure rating of the heaters, exchangers, and vessels. Also, the Recycle Gas Compressor usually will not be able to put up the required head on a significantly lighter, higher purity recycle gas.

Consequently the major control variable for hydrogen partial pressure is the recycle gas purity which should be monitored closely to assure it is always maintained above the minimum value. The hydrogen purity can be improved by either increasing the hydrogen purity of the makeup hydrogen, venting gas off the Cold Separator, or reducing the temperature at the Cold Separator.

3.6   **RECYCLE GAS RATE**

In addition to maintaining a prescribed partial pressure of hydrogen in the reactor section, it is equally important to maintain the physical contact of the hydrogen with the catalyst and hydrocarbon so that the hydrogen is available at the sites where the reaction is taking place. This is accomplished by circulating the recycle gas throughout the reactor circuit continuously with the Recycle Gas Compressor. The amount of gas which must be recycled is a design variable again set by the design severity of the operation. The standard measure of the amount of gas required is the ratio of the gas being recycled to the rate of fresh feed being charged to the catalyst.

As with hydrogen partial pressure, the recycle gas/feed ratio should be maintained at the design ratio calculated as follows:

$$H_2/HC = \frac{\text{Total Gas to Reactors (Nm}^3\text{/h)} \times H_2 \text{ Purity (Mole Fraction)}}{\text{Fresh Feed (m}^3\text{/h)}}$$

where the Fresh Feed rate is based on the specific gravity of the feed at 15°C. For this Unit, the H_2/HC ratio is on the order of 2000 - 2100 Nm^3/m^3.

A gradual increase in pressure drop of the reactors will tend to reduce gradually the recycle gas flowrate during the course of an operating period. The Recycle Gas Compressor of the HCK Unit is speed controlled and the reactors pressure drop increase can be compensated by increasing the compressor rotation speed in order to operate the Unit at constant H_2/HC ratio.

The recycle gas hydrogen content should be 89 mole % in normal operation (with a make-up hydrogen purity of 99.9 mole %). Typically, 80 mole % H_2 in recycle gas is a minimum for continuous operation.

3.7   MAKEUP HYDROGEN

3.7.1   QUALITY

The quality of the hydrogen-rich gas from the Hydrogen Plant is an important variable to the performance of the Unicracker, since it can affect the hydrogen partial pressure and recycle gas/feed ratio and thereby influence the catalyst stability (deactivation rate). The following guidelines should be used in operating the Hydrogen Plant to produce acceptable feed gas to the Unicracker:

a. Hydrogen Purity

The purity of hydrogen in the makeup gas to the Unicracker will have a major influence on the hydrogen partial pressure and recycle gas/feed ratio. Therefore, the minimum purity on the makeup gas should be set to provide the minimum recycle gas purity allowed (i.e. 80 mole % for continuous operation). If the Hydrogen Plant is unable for some reason to produce minimum hydrogen purity product, it may be possible to purge sufficient recycle gas off the Cold Separator to maintain the recycle gas purity requirements.

b. Nitrogen and Methane Content

The total of the nitrogen and methane contained in the makeup gas is only harmful as a diluent, i.e., it will reduce the hydrogen partial pressure and as long as the minimum hydrogen purity is maintained, it will not affect the unit. However, it should be noted that excessive quantities of molecular nitrogen entering the Unicracker in the makeup gas stream can cause a build-up of nitrogen in the recycle gas since the nitrogen is noncondensable. If this is the case, the nitrogen will have to be removed from the reactor circuit by a small, continuous purge of recycle gas off the Cold Separator.

c. CO + CO_2 Content

The normal specification for CO plus CO_2 in the makeup gas stream to the Unicracker is about 20 mol-ppm maximum. Larger quantities can have a harmful effect on catalyst activity. CO is considered the worst impurity due to the fact that it has a limited solubility in both hydrocarbon and water and will, therefore, build up in the recycle gas. CO_2, on the other hand, is much more soluble and is readily removed from the system in the Cold Separator liquids.

Both CO and CO_2 have similar effects on the Unicracker catalyst; they are converted on the active sites of the catalyst in the presence of hydrogen to methane and water. This methanation of CO and CO_2 competes with the normal hydrocarbon reactants for the active sites on the catalyst. Therefore, if CO + CO_2 is allowed to build up, higher catalyst temperatures will be required. In an extreme case where a large quantity of CO or CO_2 would be introduced to the Unicracker unit in a short period of time, it is theoretically possible that a temperature excursion would result since the methanation reaction is highly exothermic.

It is recommended practice that if the CO + CO_2 content exceeds the maximum design limit, the catalyst temperature should not be increased to compensate for a resulting decrease in conversion. Catalyst temperature should be maintained at the same level or reduced until the problem causing the high CO + CO_2 is eliminated. In this way the catalyst will not be harmed by increased deactivation at a higher temperature and it will also eliminate the possibility of a temperature runaway due to methanation.

3.7.2   QUANTITY

The Unicracker Process consumes hydrogen. Therefore, hydrogen must be added to the reactor circuit continuously in order to maintain the hydrogen pressure. The amount of hydrogen which is required is determined by:

a. Chemical Hydrogen Consumption

The hydrogen which is consumed in the saturation, desulfurization, denitrification and hydrocracking reactions.

b. Solution Losses

The hydrogen which is removed from the reactor circuit dissolved in the liquid hydrocarbon leaving the Cold Separator.

c. Mechanical Losses

A certain amount of hydrogen is lost through the Makeup and Recycle Gas Compressors' packing vents and seals.

d. Venting Losses

If a purge of recycle gas off the Cold Separator is required for purity, the hydrogen which is removed is considered as a venting loss.

Of the above, the chemical hydrogen consumption by reaction is normally the major portion of the total makeup hydrogen requirement. The chemical consumption is going to be determined by the type of feed (hydrogen content of the feed), the type of products being made, and the design conversion level.

3.8   WASH WATER

The wash water injection in the reaction loop, upstream of the Hot Separator Vapor Condenser (E-07) has two functions:

1. It helps removing the ammonia from the reactor circuit (this ammonia inhibits the activity of the catalyst).

2. It dissolves the cristallized ammonium salts produced in the reactors. With no wash water, the salts would precipitate in the Hot Separator Vapor Condenser and lead to tube fouling and corrosion.

The unit should not be operated for long periods of time without water injection, since catalyst deactivation and tubes fouling will result. The reactors temperatures should not be increased while water is stopped to compensate for loss in catalyst activity.

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4. PROCESS CONTROL PHILOSOPHY

4.1 DISTRIBUTED CONTROL SYSTEM DESIGN SPECIFICATION

Enclosed the "Distributed Control System Design Specification" (document 5466S-1920-GG-RP-1543-01).

4.2   DESCRIPTION OF INTEGRATED CONTROL LOOPS OF THE UNIT

4.2.1   FEED FILTER AND AUTOMATIC BACKWASH SYSTEM

The feed filter FL-01 is made of six banks of 9 pair filter elements each. The backwash of the filter uses filtered feed oil circulating through the filter element against the normal flow direction. The backwash sequence consists in isolating and backflushing one pair of filter element after the other (1). It means that the filtration operation is not interrupted by the backwash of the filter, since all the filter remain in operation except the pair of filter element that is being flushed at the considered time.

The backwash sequence is automatically activated by PDSH-005 (when pressure drop through the filter exceeds a preset threshold). A second backwash is started immediately after the first backwash, if the pressure drop is still high. Furthermore, an independent timer will start a backwash sequence every fixed time (adjustable), in order to ensure that, in case of DPT-005 failure, dirt built up is avoided.

The backwash material, contaminated with the filter residue, is collected in the Backwash Surge Drum V-11, which is fuel gas blanketed and breathing to the LP Flare header. From V-11, the backwash material is pumped with P-05 A/B on V-11 level ON/OFF control (LSL-005 and LSH-005), and sent to the battery limit to the Heavy Slops Tank after cooling in the Backwash Cooler E-13. The temperature of this product to the battery limit is controlled by TIC-369 acting on the backwash rundown flow via TV-369.

Note (1) : The following sequence is carried out successively for all pairs of filter elements:

- close the inlet line (XV-9xy A) and opening the backflushing line (XV-9xy B), this will create an immediate reverse flow in the corresponding filter elements,

- switch back to the original filtration position (XV-9xy A open - XV-9xy B closed) after the specified backflushing time is elapsed.

4.2.2   BACKWASH SURGE DRUM CONTROL SYSTEM

Drum V-11 low level switch LSL-005 shall :

◆ Shutdown the backwash pumps 08-P-05 A & B,

◆ Start timer KY-005 set at 30 minutes for line flushing,

◆ Deenergize solenoid valve UYV-019, opening valve UV-019 to flush line from pump discharge to Heavy Slop Tank at a rate of 8 m3/h,

◆ Disconnect controller TIC-369 from valve TV-369, opening position of valve TV-369 being manually adjusted by operator via HIC-036.

Timer on cycle should be adjustable based on actual piping layout and control valve purchase.

Drum V-11 high level switch shall start the backwash pump 08-P-05 A (or B) and restore the action of controller TIC-369 on the valve TV-369.

Thus the backwash material rundown temperature is automatically controlled when pump P-05 is running.

Pump P-05 (A or B) shall always be running when drum V-11 level LT-005 is above low level switch LSL-005, in order to ensure complete draining of the drum between two backwash sequences of the filter (and prevent from drum overflow).

This is achieved by comparing pumps P-05 A/B status (XL-005 A/B) to the LT-005 value.

High high level switch LSHH-005 via time delay (10 sec.) shall send a signal to stop the backwash sequence of the feed filter FL-01.

4.2.3   CONTROL OF HYDROGEN FLOW TO EACH PASS OF HEATER H-01

The purpose of this control system is to maintain an equal split of hydrogen flow to each heater pass, through control valves FV-009 A thru J.

This split is accomplished by controlling the opening of one control valve in an almost wide open position, thus causing the other control valves to work in a throttling position. The control valve which is maintained in an almost wide open position is determined automatically by the individual loop resistance to flow, and can change from time to time as the individual loop resistances change.

The output signal of each loop FIC-009A thru J controller is monitored by the low signal selector ZY-001. The lower output signal passes through the selector and is used as the measurement signal of the ZIC-001 controller. The output signal of the ZIC-001 is used as the setpoint for each flow controller in the system. The setpoint of the ZIC-001 controller will normally be set at approximately 10% to maintain one control valve in the system in an almost wide open position.

4.2.4   MAKE-UP GAS COMPRESSOR CONTROL SYSTEM

The purpose of the Makeup Gas Compressor control system is to control the Cold Separator pressure (PIC-013) while protecting the compressors from low suction pressure and high discharge pressure conditions by controlling each interstage drum pressure. See Sketch "J" (on the following page).

During normal operation, the cold separator pressure controller PIC-013 controls the third stage spillback valves PV-020 C/D through a low signal selector PY-020K. Each interstage drum pressure controller will control the appropriate spillback valve through a low signal selector to maintain drum pressure. The last stage drum pressure controller will control the Makeup Gas rate to maintain drum pressure.

For a compressor suction low pressure condition, the appropriate interstage suction drum pressure controller will open the appropriate spillback valve through a low signal selector to maintain the compressor suction pressure.

For a high pressure condition in any interstage drum, the appropriate drum pressure controller will first close the spillback valve that is bringing gas into the drum (or open the Vent to Fuel Gas control valve), and then, through a low signal selector, open the spillback valve which will release gas from the drum in order to maintain drum pressure.

4.2.5   HEATER H-01 COMBUSTION CONTROL SYSTEM

The control system is designed to control the temperature of the heater process outlet, (TIC-024) by adjusting the quantity of fuel and combustion air. See Sketch "K" (on the following page).

The objective of this control system is to ensure that the air excess is always equal to the required value, by sending sufficient combustion air flow to the burners of heater H-01, when increasing or decreasing the load of the heater. In case the fuel gas demand to the burners increases, the system will first increase the combustion air flow (FIC-018) and then increase the fuel gas flow (FIC-024). Conversely, when the fuel gas demand decreases, the system will first decrease the fuel gas flow and then decrease the combustion air flow.

The two main parameters the operator can act upon are :

◆ The threshold of the fuel gas minimum pressure required at burners level, using HIC-013 (this parameter is to be adjusted on site, according to manufacturer's recommendations).

◆ The Air/Fuel ratio, using HIC-015. The operator adjusts this parameter according to the information of the flue gas O_2 content given by the analyser AI-003 (which gives information on the combustion air excess).

System Operation

The setpoint of the combustion air flow controller FIC-018 is reset by either the heater outlet controller TIC-024 output signal or the fuel flow signal FI-024B (Gas mass flow / hr), whichever is the greater, via the high signal selector TY-024B. The high signal selector output is multiplied by the Air/Fuel Ratio setting (mass of Air / mass of Gas) to provide the proper setpoint for the combustion air flow controller. Thus, the combustion air flow rate is adjusted, as required, to maintain the desired quantity of excess air.

The compensated combustion air flow signal from FI-018 (mass flow air / hr) is divided by the Air /Fuel Ratio setting (mass of Air/mass of Gas) to obtain the allowable Gas mass flow / hr signal.

The setpoint of the fuel flow controller FIC-024 is reset by either the heater outlet controller output signal (TIC-024), or the calculated allowable Gas mass flow / hr signal, whichever is the lower, via the low signal selector TY-024A. Thus, an increase in fuel flow is limited by the available combustion air.

Computations are based on the following relationship :

$$\text{Air / Fuel Ratio} = \frac{\text{mass Air}}{\textit{mass } \text{Gas}} \quad \text{(set by the operator, using HIC-015)}$$

$$\text{Combustion Air}_{mass/hr} = \text{Gas}_{mass/hr} \times \text{Air/Fuel Ratio}$$

4.2.6   **STEAM DISENGAGING DRUM (V-58) LEVEL CONTROL**

The summing function FY-039 will accept the linearized steam flow signal from FI-039 and add or subtract from the steam signal a signal which is the difference between the steam drum level controller (LIC-031) output and a bias signal. The bias signal shall be set at 50% signal to enable the level controller to make changes for both increasing and decreasing levels. The resultant output signal from the summing function is the setpoint signal to the boiler feed water flow controller FIC-038. The boiler feed water flow controller can be operated as an independent flow control when not on cascade.

Operation

The system is designed so that for every weight unit of steam make a weight unit of feed water is introduced and the level controller signal is only used to compensate for blowdown and error in steam to water ratio. A wide proportional band and a slow reset on the level controller is required so that sudden changes in level will not affect the feedwater flow controller. Only average level changes are to affect the feedwater flow controller.

Summer Output = Steam Flow + (Level Controller Output - Bias)

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4.2.7   HEATER H-51 COMBUSTION CONTROL SYSTEM

The control system is designed to control the temperature (or percent vaporization) on the outlet of the heater (TIC-267), by adjusting the quantity of fuel and combustion air. See Sketch "B" (on the following page).

The objective of this control system is to ensure that the air excess is always equal to the required value, by sending sufficient combustion air flow to the burners of heater H-51, when increasing or decreasing the load of the heater. In case the fuel gas demand to the burners increases, the system will first increase the combustion air flow (FIC-064) and then increase the fuel gas flow (FIC-063). Conversely, when the fuel gas demand decreases, the system will first decrease the fuel gas flow and then decrease the combustion air flow.

The two main parameters the operator can act upon are :

♦ The threshold of the fuel gas minimum pressure required at burners level, using HIC-024 (this parameter is to be adjusted on site, according to manufacturer's recommendations).

♦ The Air/Fuel ratio, using HIC-023. The operator adjusts this parameter according to the information of the flue gas O_2 content given by the analyser AI-006 (which gives information on the combustion air excess).

System Operation

The setpoint of the combustion air flow controller FIC-064 is reset by either the heater outlet controller TIC-267 output signal or the fuel flow signal FI-063B (Gas mass flow / hr), whichever is the greater, via the high signal selector TY-267B. The high signal selector output is multiplied by the Air/Fuel Ratio setting (mass of Air / mass of Gas) to provide the proper setpoint for the combustion air flow controller. Thus, the combustion air flow rate is adjusted, as required, to maintain the desired quantity of excess air.

The compensated combustion air flow signal from FI-064 (mass flow air / hr) is divided by the Air/Fuel Ratio setting (mass of Air / mass of Gas) to obtain the allowable Gas mass flow / hr signal.

The setpoint of the fuel flow controller FIC-063 is reset by either the heater outlet controller output signal (TIC-267), or the calculated allowable Gas mass flow / hr signal whichever is the lower, via the low signal selector TY-267A. Thus, an increase in fuel flow is limited by the available combustion air.

Computations are based on the following relationship :

$$\text{Air / Fuel Ratio} = \frac{\text{mass Air}}{\text{mass Gas}} \quad \text{(set by the operator, using HIC-023)}$$

$$\text{Combustion Air }_{mass/hr} = \text{Gas }_{mass/hr} * \text{Air / Fuel Ratio}$$

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