



BARAUNI REFINERY

PRE-FEASIBILITY REPORT

OF

**BS-IV Project:
MS Quality Up-gradation &
HSD Quality Up-gradation
at
IOCL Barauni Refinery**

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Chapter-1

Project

1.1 Introduction of Project :

Barauni Refinery was commissioned in 1964 with an initial capacity of 2.0 MMTPA and through various revamps the crude processing capacity has been augmented to 6.0 MMTPA in 2002. CRU (Catalytic Reformer Unit) was commissioned in 1997 for production of unleaded motor spirit. To enhance productivity & profitability, Barauni refinery expansion project consisting of RFCCU, DHDT, SRU- I/II, ARU, SWSU and HGU along with associated utilities and offsite facilities were commissioned in 2002. Post Barauni Refinery Expansion, the refinery capacity was augmented to 6.0 MMTPA crude with 16% High Sulphur crude processing capability. Towards product quality up-gradation, a third reactor has been commissioned in DHDT in 2009 to reduce sulphur content in treated diesel to < 40 ppm at its design capacity. For the production of BS-III compliant Motor Spirit (MS), MSQ project with Naphtha Hydrotreating Units (NHDT), isomerization Unit (iSOM), Selective Hydrogenation Unit (SHU), Hydro desulfurization unit (HDS) and one new Hydrogen unit was commissioned in 2010 to produce 0.7 MMTPA BS-III MS.

Currently the refinery is implementing 1 x 100 TMTPA Biturox Unit for the production of VG-30 & VG10 bitumen, which is expected to be commissioned in 1st quarter of 2016. Coker-A revamp for modernization & reliability improvement is also under implementation and expected to be commissioned by last quarter of 2016. Post Biturox and Coker -A revamp scenario, the high sulphur processing capability of the refinery becomes approx. 20 wt% . Crude processing capacity of BR is normally about 6.3 MMTPA and produces about 100 TMTPM of BS-III MS and 280 TMTPD of BS-III HSD specifications. As per Auto Fuel Vision & Policy-2025, report of expert committee May'14, refineries have to produce 100 % BS-IV compliant auto fuels (MS & HSD) in w.e.f. 1st Aug'17.

1.2 Premises of the Project study

Total BS-IV MS/HSD supply being statutory requirement, low cost option has been explored with maximum utilization of the existing assets i.e, revamp of existing units has been considered instead of new units to the extent feasible. An in-house study has been performed by exploring various means of achieving 100 % BS-IV MS & HSD. The study is based on Licensor's input, literature survey/ in-house expertise and information. The study indicated, capacity revamp of DHDT, NSU, CRU and additional Prime G+ train is required to meet 100 o/o HSD and MS of BS-IV specifications, Except additional prime -G+ train with combined II stage selective HDS section, which is Grass root project , remaining BS-IV projects are revamp of existing unit/facility. It is worth to indicate, additional train of Prime G+ is recommended instead of revamping existing PG+ (as per M/s Axens) owing to high incremental capacity to be processed in PG+ unit.

1.3 Project Description & location

Major facilities envisaged/required unit capacity for 100 % BS-IV compliance are provided below:

Unit	Capacity Augmentation Envisaged, MMTPA		Remarks
	Present	Proposed	
NSU	0.464	0.76	Revamp / New Unit : Conversion of existing GSU of RFCC as NSU column or New NSU as existing NSU revamp not feasible (as per PDEC report)
NHTU	0.30	0.472	Revamp: Modifications to meet hydraulic load along with replacement of Hydro-treatment catalyst to suit application
CRU	0.3	0.47	Revamp: Provision of additional fourth reactor along with Catalyst Regenerator to operate in CCR mode with other three Reactors operating in SR mode (Dual forming)
PRIME-G+	0.32 (SHU) 0.22 (HDS)	0.76 (SHU) 0.56 (HDS)	New Unit : Parallel new PG+ Unit with combined II stage selective HDS
DHDT	2.2	3.3	Revamp: BS IV Revamp to 3.3 MMTPA

Capacity augmentation of DHDT, NHTU and CRU are envisaged through revamp route, the necessary hardware modifications shall be performed in the

existing available plot plan of the corresponding units' In case of additional new Prime G+, the area currently rendered vacant which is adjacent to existing ISOM unit (MSQ block) location is proposed.

As the required NSU capacity shall not be achieved even after revamping (as per PDEC report), it is proposed to revamp the GSU exist in the RFCC unit as second Naphtha splitter unit. At this preliminary juncture, it is proposed to perform the revamp of the GSU (for conversion as NSU) in the existing location. However if required, shifting of critical equipments of GSU to new location shall be decided after detailed study / execution of revamp philosophy.

With the above facilities, at current crude processing capacity of 6.3 MMTPA the refinery can produce about 1.3 MMTPA and 3.2 MMTPA of BS-IV MS & HSD respectively along with flexibility to produce 25 % of Euro V specifications. Comparison of units T'put in BS-IV scenario vis- a- vis current actual operating /design capacity is compiled below.

Attribute	Design	BS-III Scenario (Actual for 2013-14)	BS-IV Scenario**
	MMTPA	MMTPA	MMTPA
Crude Tput	6.0	6.47	6.3
HS%	13-15	8.98	19.7
Coker A T 'put	0.6	0.38	0.26
Coker B T 'put	0.5	0.12	0.00
RFCCU T 'put	1.43	1.69	1.7
Biturox T 'put	0.15	-	0.15
DHDT T 'put	2.2		
CRU T 'put	0.2	0.36	0.47
NHDT T 'put	0.18	0.25	0.24
ISOM T 'put	0.13	0.23	0.25
SHU	0.32	-	0.46
Prime G+ (HDS)	0.22	0.27	0.48
SRU	80 MTPD	28.4 MTPD	59.54 MTPD
F&L	-	8.99	9.0
Margin \$/bbl	-	5.45	6.84
BS-III MS	700	1190	Nil
BS-IV MS	Nil	Nil	1312
BS-III HSD	-	3249	Nil
BS-IV HSD	2156	Nil	3210
SKO	750	820	250

**** After execution of CRU Revamp / NSU revamp, parallel Prime G+ train, DHDT revamp to 3.3 MMTPA**

1.4 Additional offsite /utility system

Details of additional offsite & utility system which are planned to be installed as per project given below.

No major new utility facility or additional offsite facility is envisaged in BS-IV projects as the existing utility system is found to be adequate. Hydrogen balance, CW balance, Air, Nitrogen, and Steam & Power balance of the refinery post implementation of BS- IV projects is given below.

Utility	Unit of Measurement	Design	Present	Post
Power	MW/hr	90	43.5	50.85
Steam	MT/hr	1042	585	667
Air	NM3/hr	36900	11200	16865
Nitrogen	Nm3/hr	2040	1160	1373
Hydrogen	MT/hr	9.08	5.75	7.0
Cooling Water	m3/hr	52500	29900	30200

Offsite Facilities:

No new intermediate storage facilities have been considered as unit revamps are involved instead of new units (excluding new PG+ where hot feeding is envisaged). However new piping interconnections will be required to provide feed to new units and routing product streams to existing rundown tanks. However actual requirement of tankages and offsite facilities required if any shall be firmed up in DFR stage. At present no cost has been considered for offsite facilities.

1.5 MS Path-forward

MS pool consist of mainly three streams namely Heavy Reformate, Isomerate and treated RFCC gasoline. Thus meeting final MS pool specifications depends the combined performance of the units i.e CRU (including NHTU) , ISOM (including NHDT) and Prime G+ unit.

1.6 HSD Path-forward

With back-blending of about 25 wt % (BS IIL scenario), DHDT is currently operated at about 2.6-2.7 MMTPA against the design capacity of 2.2 MMTPA, but with bottlenecks. Execution of Low cost DHDT revamp along with replacement of advance catalyst variant is scheduled in Nov'14 S/D, will ensure operation of DHDT up to 2.9 MMTPA. Major modifications envisaged in the low cost revamp include augmentation of feed filter, replacement of feed & Product pumps, control valves etc'. Moreover the new catalyst charge is designed to handle 3.3 MMTPA scenario as well. Although new catalyst charge system (planned to be loaded in Nov'14) would be able to handle up-to 3.3 MMTPA operation, following modifications are essential for operation of DHDT at 3.3 MMTPA. Envisaged major modifications include

- Additional MUG compressor
- Wash water pump, Hydraulics limitation if any.

High capacity test run at 3.3 MMTPA is planned to be conducted after execution of low cost DHDT revamp to assess hydraulic limitation required if any and dovetail above mentioned modifications. As the targeted revamp capacity of DHDT is about 10 % more than 2.9 MMTPA (capacity achievable post execution of low cost DHDT revamp, it is proposed to straight away go for investment approval based on BDEP & cost estimate from PDEC.

1.7 Project Cost :-

Total capital expenditure has been estimated to be Rs 1327 Cr (with + 30 % accuracy), inclusive of 81 Cr. financial cost component, calculated on a period of four years of Phasing.

Chapter -2

MS Quality Up-gradation

2.1 Units considered for MS Quality Up-gradation

- | | |
|--------------------------------|--------|
| 1) PRIME-G+ Unit | (MSQ) |
| 2) Naphtha Splitting Unit | (NSU) |
| 3) Naphtha Hydro Treating Unit | (NHTU) |
| 4) Catalytic Reforming Unit | (CRU) |

2.2 Comparison of Present & Proposed

Unit	Capacity Augmentation Envisaged, MMTPA		Remarks
	Present	Proposed	
NSU	0.464	0.76	Revamp / New Unit : Conversion of existing GSU of RFCC as NSU column or New NSU as existing NSU revamp not feasible (as per PDEC report)
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PRIME-G+	0.32 (SHU) 0.22 (HDS)	0.76 (SHU) 0.56 (HDS)	New Unit : Parallel new PG+ Unit with combined II stage selective HDS

Chapter 3

Prime G+ (MSQ)

3.1 Introduction of Prime G+

IOCL Barauni refinery is considering revamping its existing MSQ (Prime-G+) unit in order to cope with tightening sulphur specifications for motor gasoline (MOGAS) on Indian market. Current design is composed of a Selective Hydrogenation Section (SHU) with two reactors in lead/lag, a three-cut splitter and a 1-stage selective HDS section with a single reactor.

It allows desulphurising 40.250 t/h (322,000 MTPA) of Light Cracked Naphtha (LCN) and 10.125 t/h (81,000 MTPA) of Heavy Cracked Naphtha (HCN) such as to achieve 150 wppm in the MOGAS pool.

Axens is now asked to offer a revamping solution for desulphurising 95.4 t/h (763,000 MTPA) of RFCC gasoline (LCN+HCN) in a staged investment:

First Investment Stage:

Down to 50 wppm of Sulphur in order to comply with BS IV specifications

Second Investment Stage:

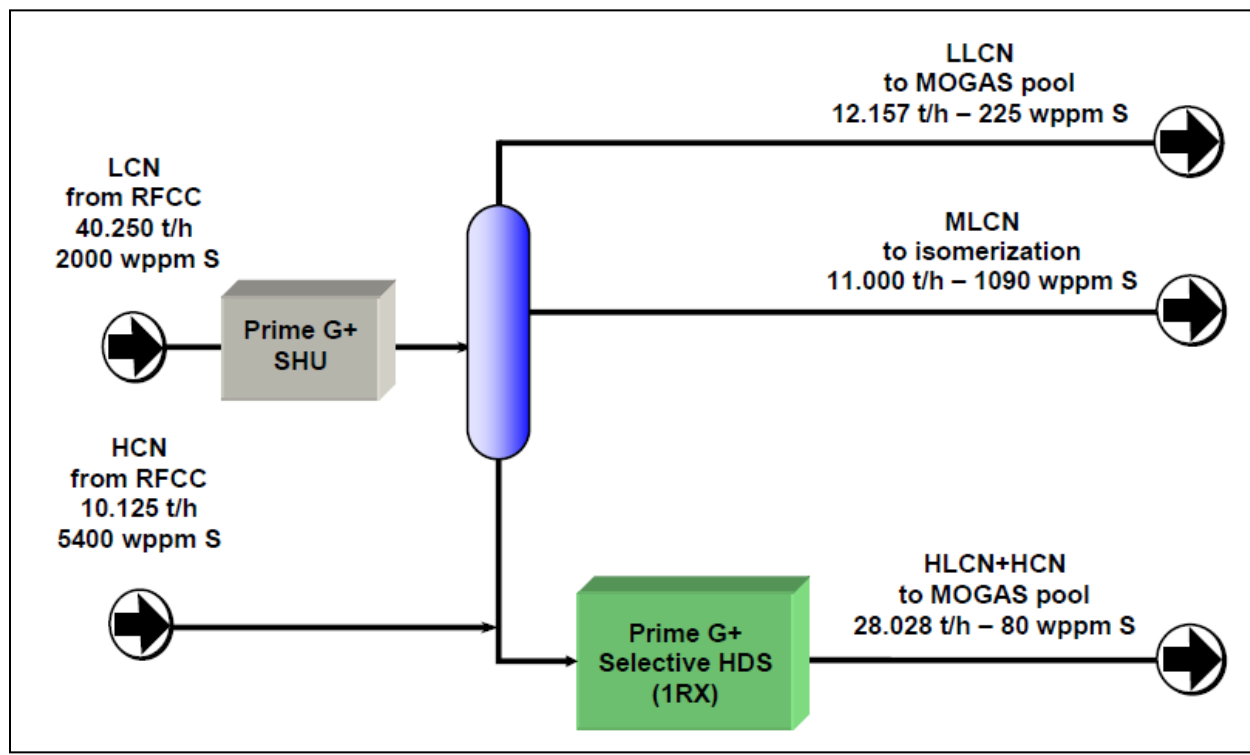
Down to 10 wppm of Sulphur in order to comply with BS V specifications

Owing to the important increase in capacity (+ 90%), Axens considered that part of the revamping solution would probably lie in duplicating part of the existing facilities (SHU and Splitter sections) and adding a new HDS section in parallel with the existing one to meet BS IV specifications.

For 10 wppm specification, configuration of one stage HDS section would lead to poor octane retention. Axens thus offers to upgrade the existing selective HDS scheme into a highly selective 2-stage Prime-G+ configuration to meet BS V specifications.

3.2 Current scheme

A simplified block flow diagram of existing Prime-G+ at Barauni refinery is provided hereafter including overall material balance from PDP at Start Of Run conditions with H₂ from HGU:



At PDP stage, 10% overdesign was considered on unit feed capacity. Current design capacity of SHU section is thus 44.275 tons/hr and current design capacity of HDS section is 30.938 t/h.

3.3 Proposed scheme for revamp

The objective of the revamping for IOCL Barauni is to increase Prime-G+ unit capacity and increase the desulphurisation rate with minimum impact on octane number in order to produce a gasoline pool complying with the following specifications:

a) First Investment Stage: Down to 50 wppm of Sulphur in order to comply with BS IV specifications

b) Second Investment Stage: Down to 10 wppm of Sulphur in order to comply with BS V specifications

Axens envisions the following for upgrading IOCL Barauni Prime-G+ meeting BS IV and BS V specifications while coping with higher FCC gasoline feed rate:

SHU section: Owing to the high incremental capacity (+90%) to be processed in Prime-G+ unit, Axens foresees to duplicate (after equipment optimization) the existing Prime-G+ SHU and splitter sections.

HDS sections: As different HDS severities are required to meet BS-IV and later, BS-V specifications, two revamp schemes allowing staged investment are proposed:

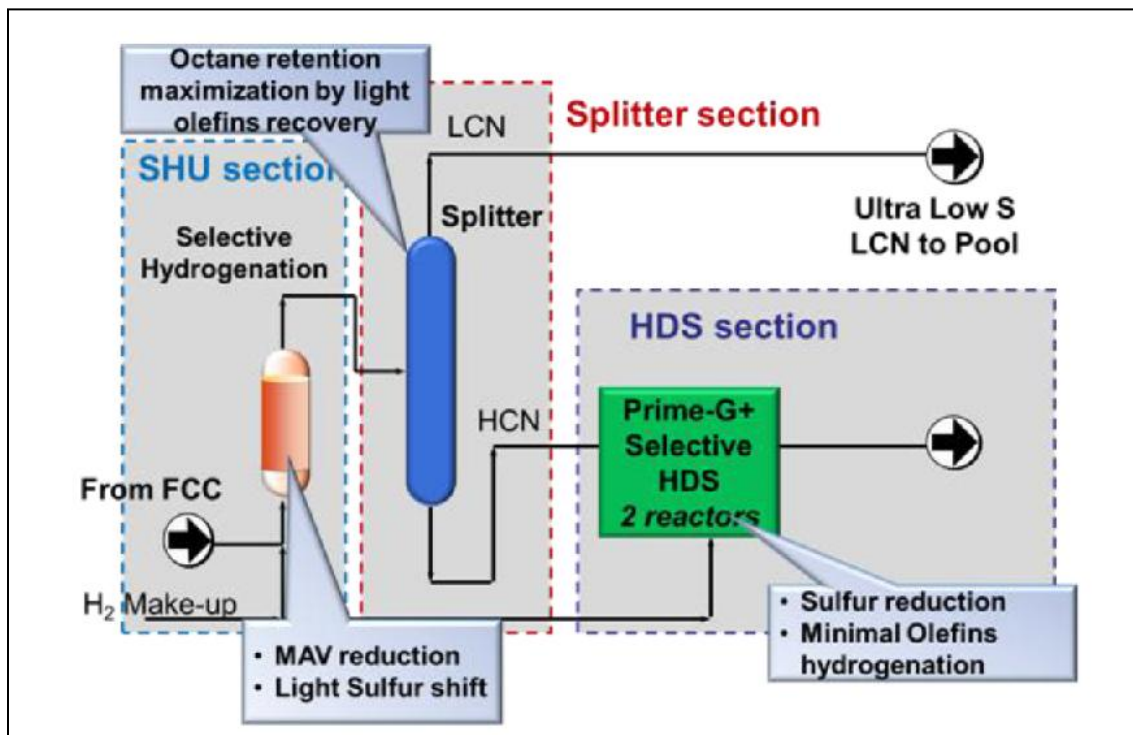
First Investment Stage (BS IV) - One Stage HDS Scheme: Due to the high incremental capacity (+90%) a new first-stage HDS section operating in parallel with the exiting section is foreseen.

Second Investment Stage (BS V) - Two Stages HDS Scheme: In order to ensure the highest octane retention to meet BS V specifications despite the deep desulphurisation, a new common HDS section is foreseen in series, with inter-stage H₂S removal.

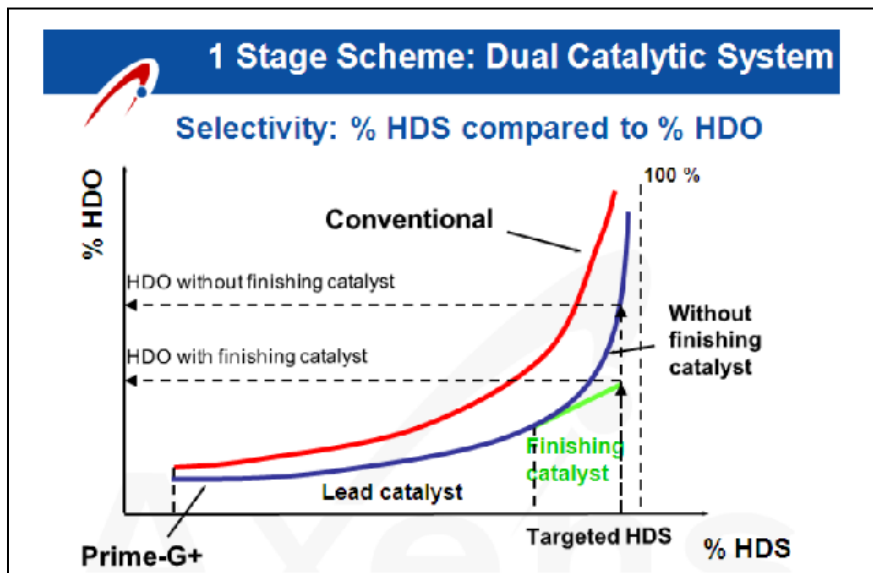
In the new configuration, existing HDS and new first stage HDS will be referred to as “first-stage HDS sections”. New, common, HDS section will be referred to as “second-stage HDS section”.

3.3.1 First Investment Stage (BS IV): One Stage HDS Scheme

To meet BS IV specifications (50 wppm of sulphur), one stage HDS scheme with dual catalytic system is proposed:

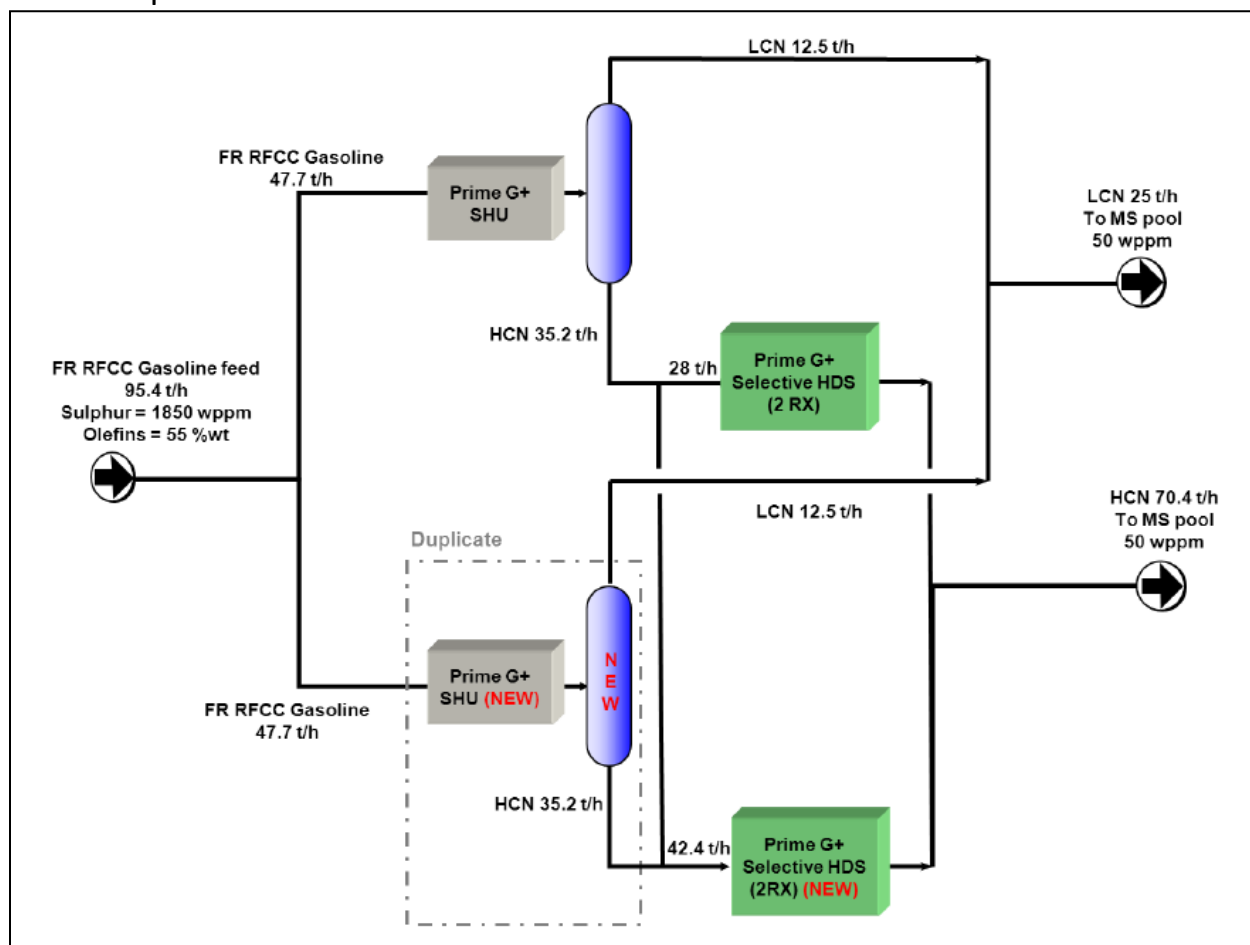


The heavy cut (HCN) is deeply desulphurized with minimum olefins saturation in in Prime-G+TM selective HDS section. Selective HDS section with two reactors relies on Axens dual catalytic system, whose principle is summarized in the figure below.



As compared to a conventional hydrodesulphurization catalyst, the lead HDS catalyst (main reactor) ensures a reduced HDO rate for the same HDS rate. The finishing catalyst enables completing HDS with almost no HDO. In addition, finishing reactor features excellent mercaptans control by getting rid of

mercaptans formed by recombination which is of major importance when targeting high levels of hydrodesulphurisation. It should be noted that hydrotreated gasoline at the outlet of Prime-G+TM process is sweet and can be routed directly to gasoline pool. This means that no further gasoline sweetening unit is required.



Owing to the important increase in capacity (+ 90%), a new SHU and Splitter section and a new first-stage HDS section operating in parallel with the exiting sections are specified.

Arrangement of new SHU and Splitter section is very similar to the existing sections and after optimization some equipment could be duplicated reducing time and cost of basic and detail engineering phases.

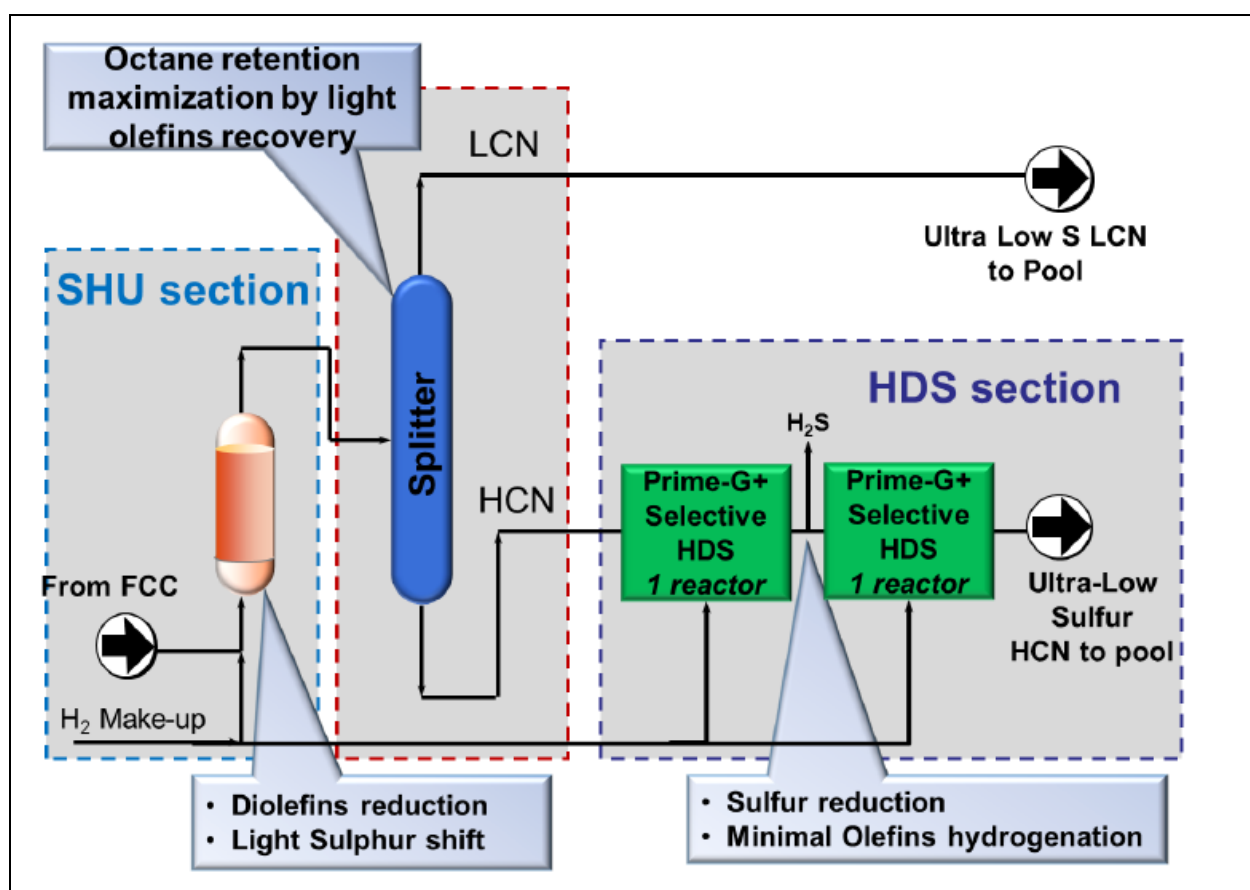
Arrangement of new first stage HDS section is similar to that of existing HDS section. It includes one fired heater, feed/effluent exchangers, a separator drum, quench pumps and a recycle gas loop (compressors, amine absorber, KO drums). In order to minimize the CAPEX as well as the plot space area, Axens can propose

mutualizing recycle gas loops of new first-stage HDS section and second-stage HDS section.

Each first stage HDS section will be equipped with two reactors. A new finishing HDS reactor shall be installed on existing first stage HDS section. In order to minimize CAPEX, finishing HDS reactors of both first stages HDS sections can be reused in future second stage HDS section (Second Investment Stage).

3.3.2 Second Investment Stage (BS V): Two Stages HDS Scheme

Axens' technical proposal to meet BS V specifications relies on the most advanced Prime-G+ configuration, the 2-stage HDS scheme, whose typical block flow is provided here below:

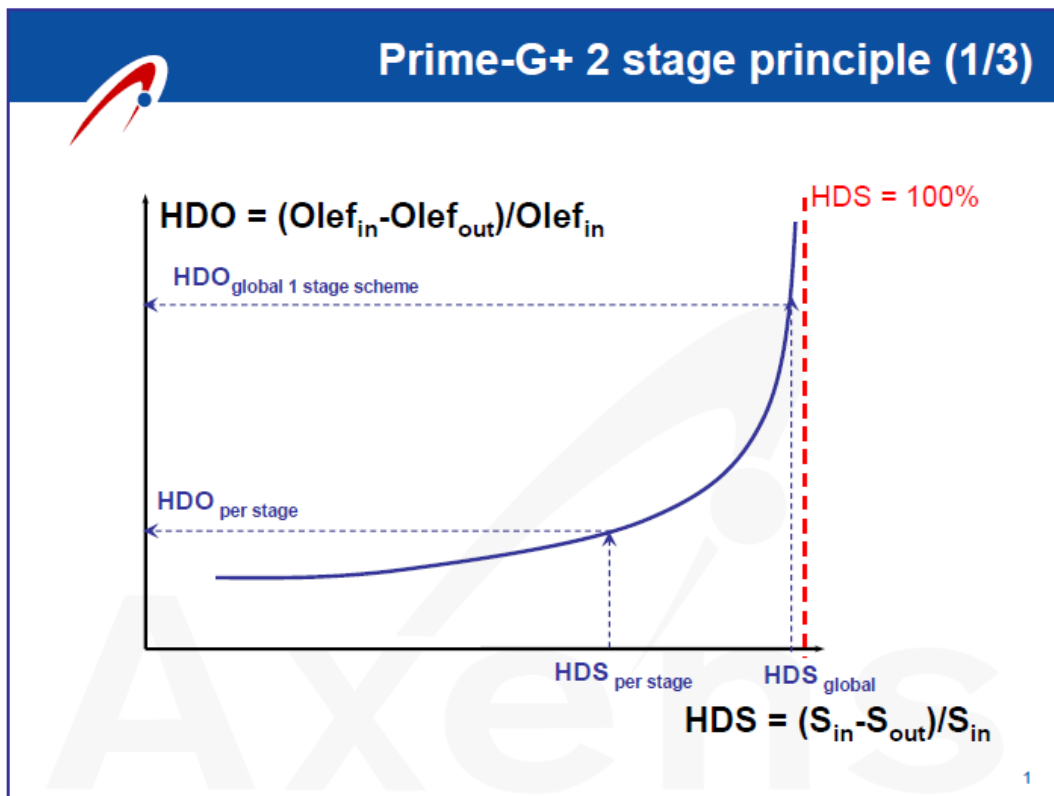


In selective hydrotreating, olefins saturation and consequently octane loss and H₂ consumption depend on the severity of the HDS required. Overall, one considers that octane loss increases exponentially with the extent of HDS. The reason is twofold: increase of temperature at very high HDS rate involving higher olefin saturation, sulphur in the product starting to be predominantly recombinant mercaptans (addition of olefins to H₂S to form RSH). In order to

further reduce sulphur content, olefins shall be hydrogenated leading to higher octane losses.

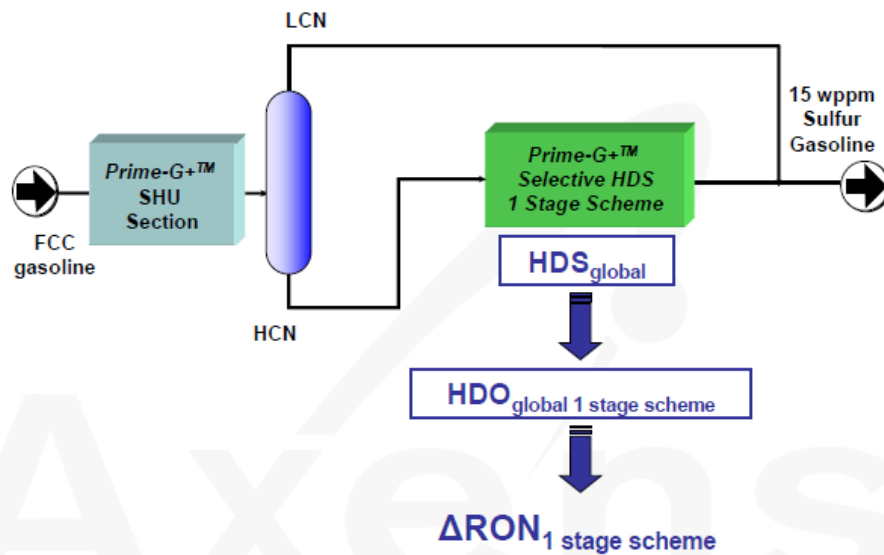
Thanks to lower HDS rate per stage and inter-stage H₂S removal, a 2-stage HDS scheme leads to higher octane retention and lower hydrogen consumption as compared to a 1- stage HDS scheme in cases of high severity HDS (generally to meet 10 wppm of sulphur in the product).

The principle of the Prime-G+ 2-stage selective HDS is summarized in the following figures; It is worth being emphasized that desulphurised product from Prime-G+ unit is sweet and does not require any further treatment before being sent to the pool.



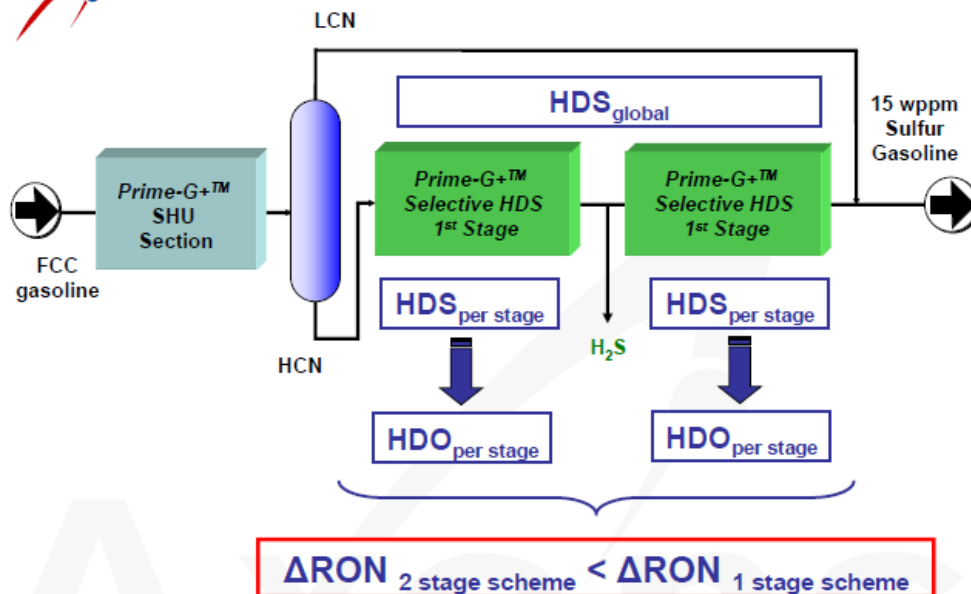
Prime-G+ 2 stage principle (2/3)

1 stage scheme



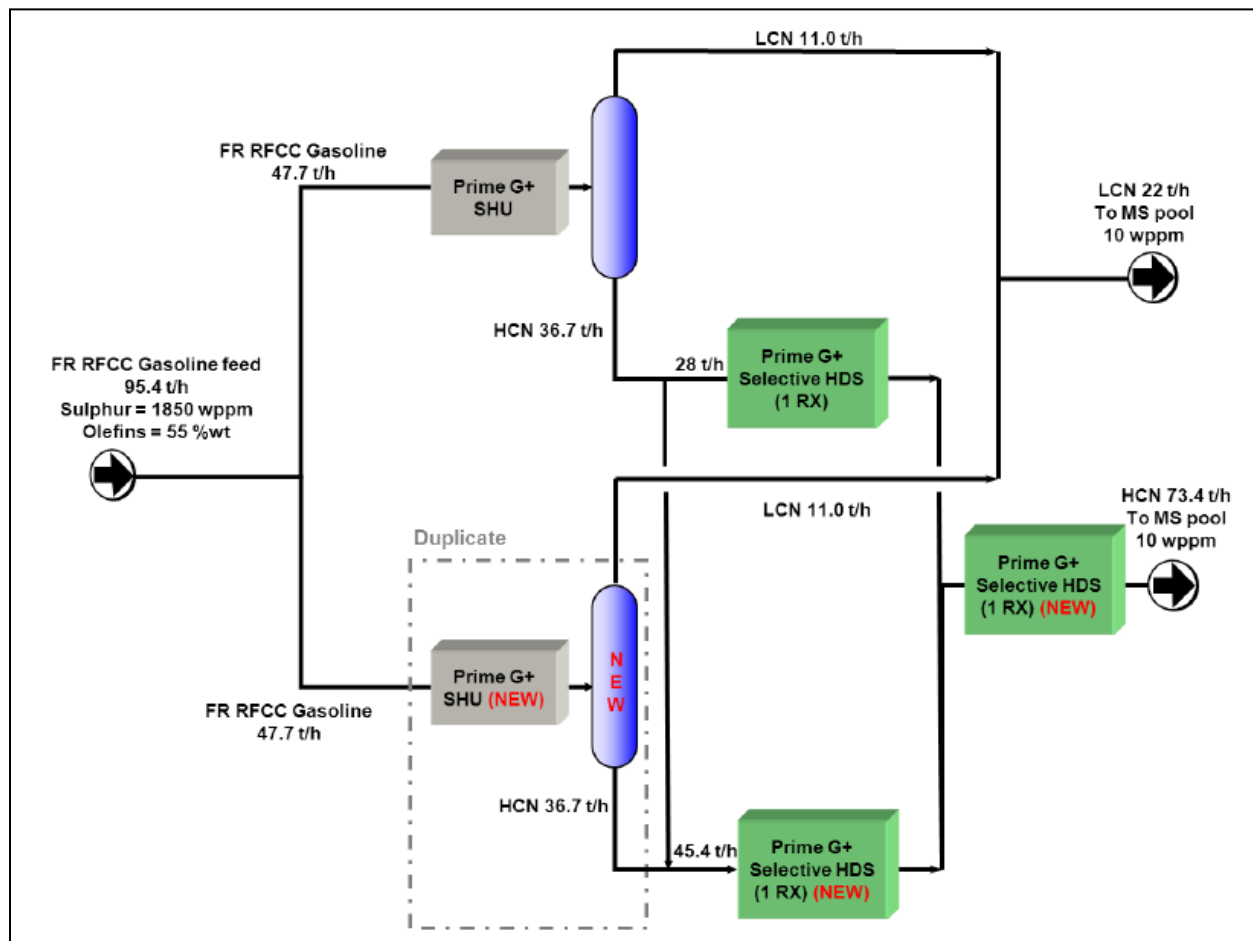
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Prime-G+ 2 stage principle (3/3)



3

Simplified block flow of Axens' solution is shown here below:



Second-stage HDS section arrangement is similar to that of existing HDS section. It includes one reactor, one fired heater, feed/effluent exchangers, a separator drum, quench pumps and a recycle gas loop (compressors, amine absorber, KO drums). In order to minimize the CAPEX as well as the plot space area, Axens can propose mutualizing recycle gas loops of first-stage HDS section and second-stage HDS section. H₂S removal from hydrocarbons in between first-stage HDS sections and second stage HDS section will be achieved in a new H₂S stripper column using part of the recycle gas from new recycle gas compressor.

Chapter -4

Naphtha Splitting Unit (NSU) **Naphtha Hydro-Treater Unit (NHTU)** **&** **Catalytic Reforming Unit (CRU)**

4.1 Introduction of NSU, NHTU & CRU

IOCL Barauni Refinery intends to revamp its existing Catalytic Reforming Unit (CRU) from existing capacity of 0.3 MMTPA to 0.47 MMTPA with Reformate RONC 100 for two feed cases considering a stream factor of 8000 hrs/year.

The current CRU includes the following units:

- Naphtha Splitting Unit (NSU);
- Naphtha Hydro-Treater Unit (NHTU);
- Catalytic Reforming Unit (CRU);
- Reformate Splitter Unit (RSU);

The existing CRU is licensed by M/s Axens France and was commissioned in April 1997 with a design capacity of 0.3 MMTPA with Reformate RONC 96. The CRU was debottlenecked in 2010 to meet the Euro III specifications thus to produce Reformate RONC 97. The capacity was maintained at 0.3 MMTPA. The process study of the CRU has been carried out in accordance with the Agreement number 05327 signed between IOCL BARAUNI and AXENS.

The bases of design were established during the kick off meeting held from 07th to 08th July 2014 in IOCL Barauni offices, Barauni – India. A hydraulic check-run and site survey have been performed during this meeting in order to evaluate the behavior of the Unit and constraints due to plot availability. The required capacity (58,75 t/hr) is 156% of initial process book design flow rate (37.54 t/hr) and 122% of the check-run capacity (48.05 t/hr). To achieve both throughput and reformate quality increase, Axens has proposed to revamp the existing reformer using DUALFORMING technology, which consists in a reaction section combining a fixed bed and moving bed technology. This Scoping Study Report contains the basic process information for Catalytic Reforming Unit (Unit 03) revamping.

4.2 PROCESS FLOW DESCRIPTION :

4.2.1 NAPHTHA SPLITTER UNIT (NSU):

IBP-140 °C cut naphtha from storage tanks is fed to splitter column (01-CC-001) under flow control by off site pump at tray No. 14. The feed is heated up to 95 °C in splitter feed/bottom exchanger against splitter bottom stream before it enters the column.

The overhead vapors are totally condensed in air condensers. The liquid collected is pumped by splitter reflux pump and one part sent as top reflux back to the column under flow control to maintain the top temperature. The balance, which constitutes the IBP-70°C cut naphtha is sent to HGU as their feed and rest light naphtha is sent to storage under reflux drum after cooling in a water cooler. Reflux drum boot water is drained in OWS manually.

The pressure of splitter is controlled at reflux drum by passing a part of hot column overhead vapors around the condenser or releasing the reflux vapors to flare through a split range controller.

The splitter bottom product which constitutes 70-140 °C cut naphtha is pumped to splitter feed/bottom exchanger by hydro treater feed pumps. The bottom product after exchanging heat with feed is split into two streams. One is fed to the hydro treater unit at a temp. of 65 °C and the other is sent to storage under column level control 01-LC-1102 after being cooled in splitter bottom column.

The heat necessary for splitter reboiling is supplied by splitter reboiler furnace and desired temperature maintained by controlling the fuel firing. Splitter reboiler pumps. provide the circulation through reboiler is double pass vertical cylindrical furnace having four burners fired from the bottom. It has soot-blowing facility for convection section.

4.2.2 NAPHTHA HYDROTREATER UNIT (NHTU):

a) REACTION AND SEPARATION SECTION:

The naphtha from NSU is fed to NHTU by a pump. The feed flow is controlled by flow control valve. The feed then mixed with Rich Hydrogen Gas from HP separator of reformer. Both the liquid naphtha and rich hydrogen gas are pre-heated in a series of exchangers, which are feed/reactor effluent heat exchangers. Then mixture is heated up to reaction temperature in a furnace and fed to the reactor. The furnace is four pass having three burners fired from bottom. The furnace is having facility of soot blowing. The reactor inlet temperature is maintained by temperature controller cascaded with either fuel oil or fuel gas Pressure Controller's. The furnace is provided with all safety shut down inter locks. It has also provision of decoking.

The desulfurisation and hydro treating reaction takes place in reactor at almost constant temperature since heat of reaction is quite negligible. The reactor is provided with facility of steam and air for regeneration of catalyst. The catalyst for reactor is HR-306. The reactor catalyst bed has been provided with five number of thermo couple points at various location to get the bed temperature during regeneration of the catalyst.

The reactor effluent after having heat exchanged in series with feed goes to air cooler. The air cooler fans pitch is variable i.e. cooling load can be adjusted as per situation requirement. After air cooler the effluent is cooled in a trim cooler. The product is collected in a separator vessel. Sour water is drained from the separator drum boot manually. The separator drum pressure is maintained by routing the gas to HGU compressor fully through and any excess gas can be routed to FG system through pressure controller. In event of emergency the separator excess pressure can be released to flare through an on-off control valve.

A line has been provided to feed the naphtha to stripper, during start up, bypassing the reaction/separation section.

b) STRIPPER SECTION :

The separator liquid is pumped under flow control cascaded with level controller to stripper feed/bottom exchanger when it gets heat exchanged by hot stripper bottom stream.

The stripper column consists of 28 Nos. of valve trays one to eight number of trays are single pass and the rest double pass. Feed enters at 9th tray from two sides. The overhead vapors are cooled down in air condenser and collected in stripper reflux drum. The fan load can be adjusted. The condensed hydrocarbons are returned to column top by pump under flow control cascaded with level as reflux to maintain the top temp. The water accumulated in the boot is sent for disposal as sour water. Pressure controller releasing excess gas in the FG system maintains the reflux drum pressure. The facility is there to inject corrosion inhibitor by pump.

Stripper bottom product exchanged heat with stripper feed in and then sent to reformer as hot feed. The excess or required hydro-treated naphtha is sent to storage after being cooled in under level control.

4.2.3 CATALYTIC REFORMER UNIT (CRU):

Hydro treated naphtha from Naphtha Hydro Treater Unit (NHTU) is pumped to required pressure by pump under flow control and mixed with recycle gas from the recycle gas compressor. The mixed feed is pre heated in the feed-effluent exchanger followed by feed/effluent exchanger. Then the mixture is brought up to the reaction temperature (480°C) by heating in the pre-heater and then fed to 1st reactor.

As the reaction is endothermic, the temperature drops, so the first reactor effluent is heated in the first inter heater prior to be sent to the second reactor. In the same way effluents are heated in the second inter heater prior to be fed to the third reactor.

The effluent from the last reactor is split into two streams and send for heat recovery parallely to feed/effluent exchanger and stabilizer reboiler. The outlet from the two exchangers is combined by a three way valve and then cooled down successively in the Zeemann Secathen exchanger, reformer effluent cooler

and effluent trim cooler. The cooled reactor effluent is flashed in the reformer separator.

Vapor and liquid phase are separated in separator. Part of the gas phase constitutes the hydrogen recycle gas to the reactor circulated by recycle gas compressor. The hydrogen rich gas compressor compresses remaining amount, corresponding to the amount of gas produced. The pressure control in separator is achieved by a kick back gas flow from HP Absorber to separator. Should the gas be produced in excess to capacity, degassing in split range to fuel gas is performed, through pressure control valves.

The separator liquid is sent by reformer separator bottom pumps under level control for recontacting with the gas compressed by compressor.

The hot flue gases from all the three reformer furnaces are combined and sent to steam generation system for waste heat recovery to produce MP steam. Provision is there to dry the recycle gas into a dryer. The dryer can later be regenerated.

The unit has also been provided with facilities for continuous chloriding, water injection, DMDS/ CCl_4 injection and caustic soda circulation.

The separator vapor after passing through KO drum is compressed in the H_2 Rich Gas Compressor and recontacted with separator liquid. The recontacted vapor and liquid is cooled in a cooler (03-EE-005) and then fed to HP absorber. The aim of this device is to allow for high recovery of the C_5 contained in the gas phase of separator and improves the quality (H_2 concentration) of the produced gas.

A part of hydrogen rich vapor goes to NHTU as a make up hydrogen through flow controller and balance goes to the suction KO drum of HGU compressor that is run to provide H_2 to DHDT after purification in a PSA unit through pressure controller. This pressure controller remains in line with fuel gas system at a slight higher set point. So that any excess gas can be routed to FG system through pressure controller.

The combined stream is cooled in LPG absorber feed cooler and flashed in LPG absorber. Off-gas is sent under pressure control to fuel gas system.

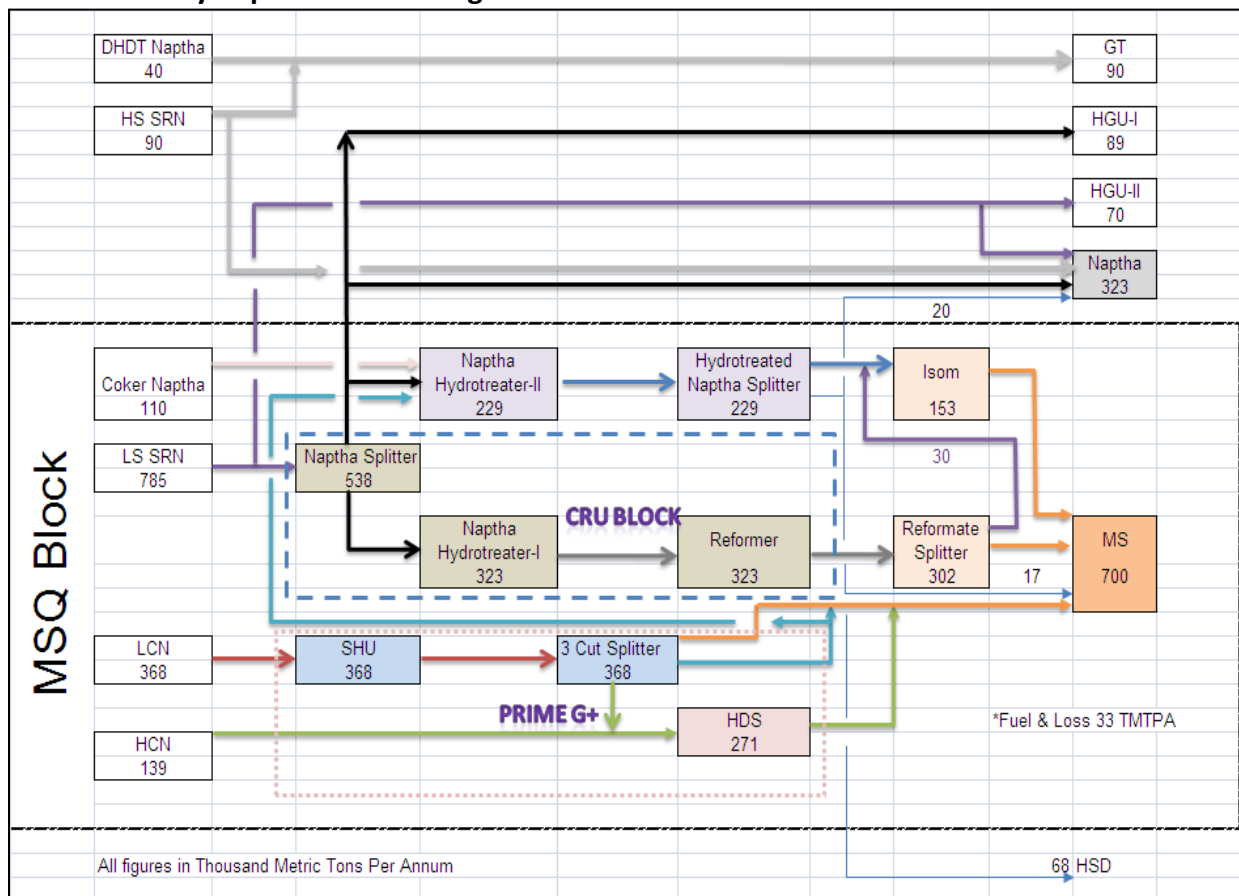
Stabilizer feed pumps. After pre heating in stabilizer feed/bottom exchanger the mixture is fed to the stabilizer at tray No. 13.

Stabilizer over head vapours is partially condensed in stabilizer condenser and flashed in stabilizer reflux drum. The vapor phase is sent to LPG absorber for C₃ and C₄ recovery. A part of condensed liquid is pumped as reflux to the column by stabilizer reflux pump under the flow control and the balance is sent to LPG Recovery Unit under level control of reflux drum.

The heat of reboiling to the stabilizer is provided by the hot reactor effluent in the stabilizer reboiler and the desired temperature maintained by controlling the flow of reactor effluent by the three way valve.

The bottom product, stabilized reformate, is cooled in the feed/bottom exchanger followed by reformate cooler and reformate trim cooler before being routed to storage Tanks.

❖ Refinery naphtha flow configuration: An overview



4.3 Revamp Strategy

The driving force of the revamp is to have a minimum operating cycle of 24 months in the semi-regenerative part of reaction section while producing a reformat RONC 100. The amount of feed has also been increased from 0.3 to 0.47 MMTPA. The second target is to maintain existing critical equipment to limit revamp cost.

From the point of view of the catalyst, the increase of capacity increases the flow rate across the catalytic bed. As a result, space velocity WHSV is increased, and the time where the naphtha is in contact with the catalyst is reduced. To maintain the current performance reactors WAIT would then have to be increased but it would result in higher coke production and thus a reduction of the cycle length.

Severity of the existing reactors has then been decreased to achieve RON 94. The remaining points can be retrieved by the Reactor operating in CCR (Continuous Catalytic Regeneration).

Targeting RON 94 at the outlet of the 3 existing reactors was not possible with the current catalytic system under revamp conditions. Catalyst system has to be updated. The solution retained consisted in loading in the 3 existing reactors the latest generation of Axens catalyst with TexicapTM solution. This leads to decrease the WHSV (dense loading) and catalyst dead space (cost savings) and finally to increase the cycle length and performances. It requires increasing the H₂/HC to sustain the coking rate to an acceptable level and so achieve the cycle length.

However, this new H₂/HC ratio tends to increase the hydraulics in the reaction loop and then raised constraints with installed equipment that are not designed for such flow rate. Knowing that, Axens has minimized wherever it was possible the pressure drops. In the same time, as reforming reactions require a substantial amount of heat, Axens has targeted to minimize the revamp of the heaters by implementing a new high efficiency exchanger.

4.4 CONCLUSION

This study has shown the revamping technical feasibility for 156% of original nominal capacity. However, this has highlighted that almost all existing equipment cannot be reused. Equipment reused such as reactors will be deeply revamped and internals will be changed. This is consistent with both capacity and severity increasing that lead to constraining operating conditions. But targeted performances and production can be reached with this technology update.

During Detail Engineering phase of the revamp project, specific characteristics of equipment and their relevant operating limitations (related to rotating machines, heaters and exchangers for instance) which are not known at this stage shall be investigated. Some significant costs may be added.

Limitations from Refinery networks (availability and properties of fuel gas, electricity, cooling water, flare network, etc) must be investigated as well. Such a Detail Engineering study is beyond AXENS' responsibilities as a Process licensor. It is expected that these tasks of a general nature will be performed by the Engineering Company and/or Unit Owner in accordance with the prevailing codes and rules and with good Engineering practice.

Chapter - 5

HSD Quality Up-gradation **Diesel Hydro-Treating Unit (DHDT)**

5.1 Introduction of DHDT

Petroleum fraction contains various amount of naturally occurring contaminants including organic sulfur, nitrogen, and metals compounds. These contaminants may contribute to increased levels of air pollution, equipment corrosion, and cause difficulties in the further processing the material. The Union fining process is a proprietary fixed-bed, catalytic process developed by UOP for hydro treating a wide range of feed stocks. The process uses a catalytic hydrogenation method to upgrade the quality of petroleum distillate fractions by decomposing the contaminant with negligible effect on the boiling range of the feed. Union fining is designed primarily to remove sulfur and nitrogen. In addition, the process does an excellent job of saturating olefinic and aromatic compounds while reducing Conradson carbon and removing other contaminants such as oxygenates and organ metallic compounds.

The desired degree of hydro treating is obtained by processing the feedstock over a fixed bed of catalyst in the presence of large amounts of hydrogen at temperatures and pressures dependent on the nature of the feed and the amount of contaminant removal required.

The hydrotreater unit is designed to improve the Diesel cetane number to 48.5(min) while meeting the diesel stability specification of 1.6 mg/100ml(max) and reducing sulfur content to 0.2wt%. Keeping in view the statutory requirement of supplying Bharat Stage – III compliant diesel to the market since the 1st of April 2010, one additional reactor.702-R-03 has been installed in series with R-01 & R-02 and the existing UOP catalyst has been replaced by the new generation Axens catalyst, HR-448 during the April-May'09 M&I shutdown. Now the unit can produce diesel with cetane no. 51 and further reduction of sulfur content to 0.05wt%. The two features of this process are Hydro treating and refining reactions. The refining reactions are

desulphurization and denitrification. The mercaptides, sulphides and disulphides react in an atmosphere of hydrogen to form corresponding saturated and aromatics compounds, H₂S and NH₃. The Hydrotreater feed consists of Straight Run Kerosene II (SRK II) and Coker Kerosene I (CK-I) and Diesel mode consisting of Straight Run Gas Oil from low sulphur imported crude (SRGO-LS), Straight Run Gas Oil from high sulphur imported crude (SRGO-HS), Total Cycle Oil from FCCU (TCO) and Light Coker Gas Oil (LCGO).

Union fining units are designed for dependable, stable operation. UOP's selective, high-activity catalysts operate for long periods of time between regenerations. Specific process objectives determine which UOP catalyst is best suited for a particular installation. The activity and selectivity of the catalyst is influenced only to a slight extent by the type of feed processed. The same catalyst in varying quantities can be used to hydro treat straight run naphtha, vacuum gas oil, catalytically and thermally cracked distillates. The widespread use of catalytic reforming units has made available large quantities of excess hydrogen, making it feasible to hydro treat many, or all, of the distillate produced by the refinery

5.2 Technology Features of DHDT

Union fining is carried out at elevated temperatures and pressures in a hydrogen atmosphere. Pressures and temperatures may range up to 454 deg C and 95 kg/cm²g. The Unionfining catalysts are formulated by compositing varying amounts of nickel or cobalt with molybdenum oxides on an alumina base. The specific catalyst system and unit design parameters will be evaluated on an economy basis for each unit. Each design will be based on feed quality, desired product properties, ease of operation, desired cycle length, operating flexibility, construction schedule and operating costs.

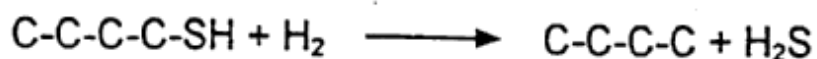
HYDROTREATING CHEMISTRY

The following chemical steps and/or reactions occur during the hydro treating process:

Sulfur Removal

Typical feed stocks to the Union fining unit will contain simple mercaptans, sulfides and disulfides. These compounds are easily converted to H₂S. However, feed stocks containing heteroatomic aromatic molecules are more difficult to process. Desulfurization of these compounds proceeds by initial ring opening and sulfur removal followed by saturation of the resulting olefin. Thiophene is considered 15 times more difficult to process compared to diethyl sulfide.

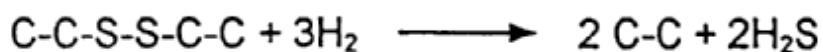
a. Mercaptan



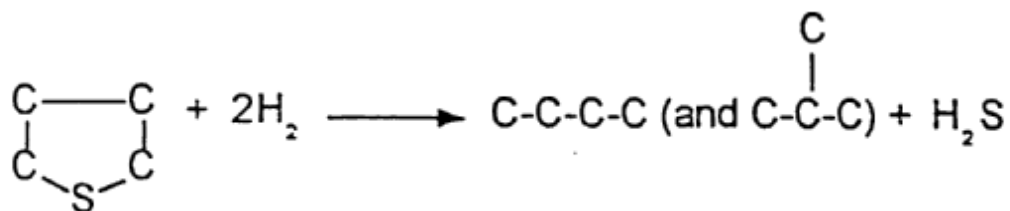
b. Sulfide



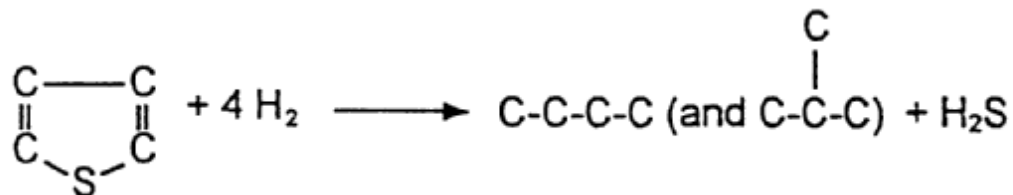
c. Disulfide



d. Cyclic Sulfide



e. Thiophenic

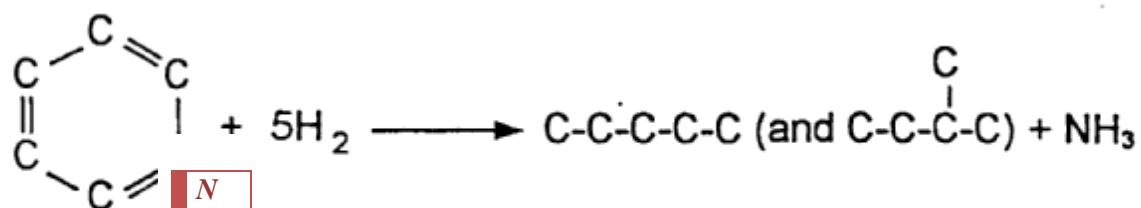


Nitrogen Removal

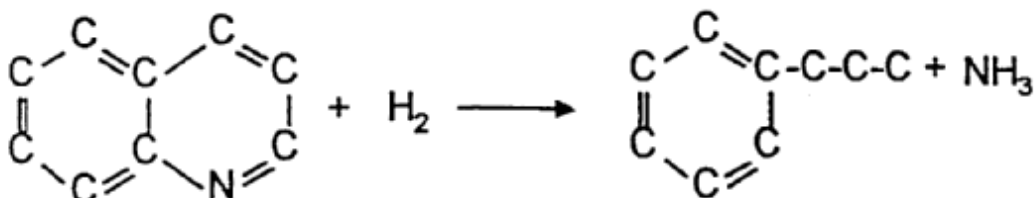
Denitrogenation is generally more difficult than desulfurization. Side reactions may yield nitrogen compounds more difficult to hydrogenate than the original reactant. Saturation of heterocyclic nitrogen-containing rings are also hindered by large attached groups (see Figure III-1)

The reaction mechanism steps are different compared to desulfurization. The denitrogenation of pyridine proceeds by aromatic ring saturation, ring hydrogenolysis, and finally denitrogenation.

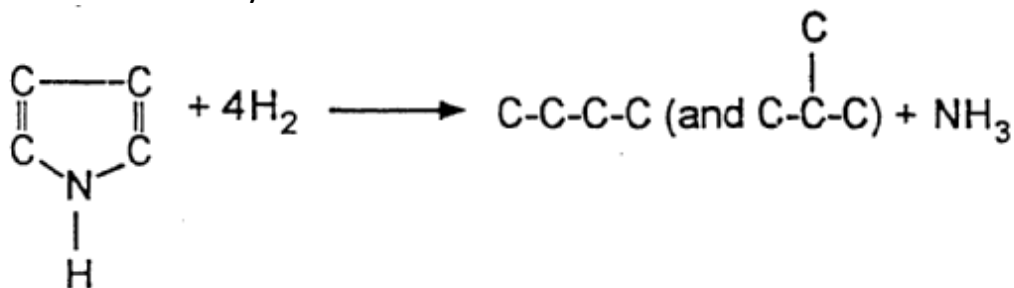
a. Pyridine



b. Quinoline



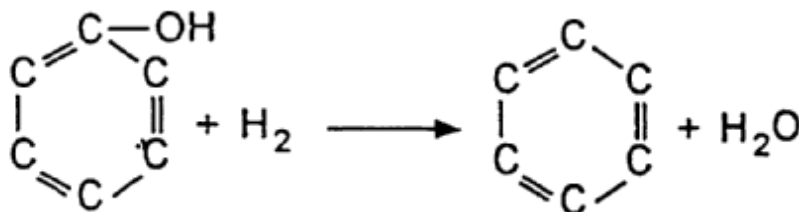
c. Pyrrole



Oxygen Removal

Organically combined oxygen is removed by hydrogenation of the carbon-hydroxyl bond forming water and the corresponding hydrocarbon.

a. Phenols



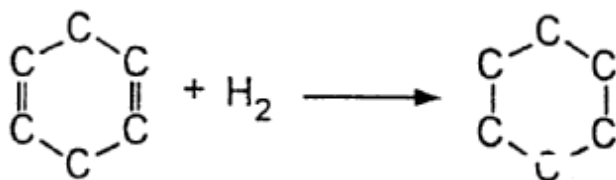
Olefin Saturation

Olefin saturation reactions proceed very rapidly and have a high heat of reaction.

a. Linear Olefin

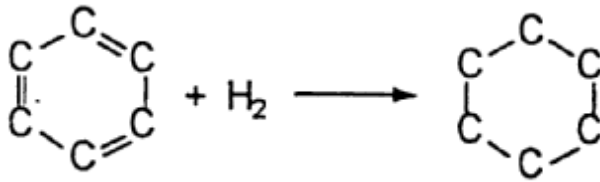


b. Cyclic Olefin



Aromatic Saturation

Aromatic saturation reactions are the most difficult. The reactions are influenced by process conditions and are often equilibrium limited. Unit design parameters would consider the desired degree of saturation for each specific unit. The saturation reaction is very exothermic.



Metals Removal

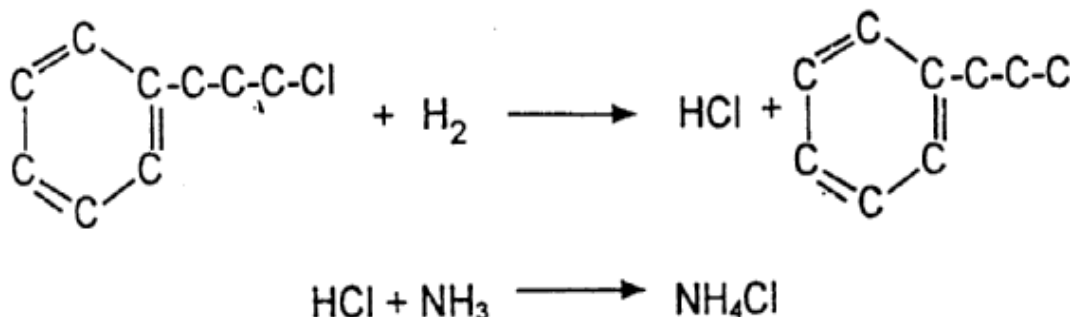
The mechanism of the decomposition of organo-metallic compounds is not well understood. However, it is known that metals are retained on the catalyst by a combination of adsorption and chemical reaction. The catalyst has a certain maximum tolerance for retaining metals. Removal of metals normally occurs 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. Lead may also deposit on the hydro treating catalyst from reprocessing leaded gasoline through the crude unit.

The useful life of the catalyst may be determined by the amount of metals that are accumulated on it during the course of operation. Most Union fining units are able to go through several operating cycles without exceeding the ability of the catalyst for removing metals. Metal removal is essentially complete above temperatures of 600°F to a metals loading of 2-3 wt% of the total catalyst. Above this level, the catalyst begins to approach equilibrium saturation and metals breakthrough is likely.

The total metals retention of the catalyst system can be increased by using a guard reactor or a guard bed of catalyst specifically designed for demetallization. Some demetallization catalysts may retain as much as 100 wt% metals based on fresh catalyst weight. Such catalysts typically have a lower activity for desulfurization and denitrogenation.

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 or leave with the stripper off-gas. Decomposition of organic halides is considered difficult with a maximum removal of ~90%.



Reaction Rates

The approximate relative heats of reaction per unit of hydrogen consumption for these reactions are:

Desulfurization	1
Olefin Saturation	2
Denitrification	1
Aromatics Saturation	1

All of the reactions discussed above are exothermic and result in a temperature rise across the reactor. Olefin saturation and some desulfurization reactions have similarly rapid reaction rates, but it is the saturation of olefin which generates the greatest amount of heat. The temperature rise expected for a given charge stock along with the desired product quality will play a very important role in determining the number, size, and arrangement of the reactors, heat exchange, and hydrogen circulation rate.

Hydro cracking Reactions

The products of the Union fining reactions are of a lower density than the feedstock. Therefore, the total liquid yield will be greater than the feed and may be as high as 102 lv%. Some hydro cracking may take place in the Union fining process. This is especially evident toward the end of an operating cycle when reactor temperatures are raised to compensate for lower catalyst activity. Total liquid yield and hydrogen consumption will increase as hydro cracking retractions proceed. Most of the increase in liquid volume yield will come from more net stripper overheads at the expense of lower stripper bottoms product. The Union fining unit is designed for maximum bottom production. Economic considerations will determine the amount of bottoms product that can be lost before the unit is shutdown for regeneration or catalyst change. Under normal operation, the net stripper overhead liquid produced in a Union fining unit should not exceed 2 lv% of the unit feed.

Hydrogen Consumption

The hydrogen consumed by the Union fining reactions is supplied externally and must be boosted to reaction system pressure. If the hydrogen is obtained from a naphtha reforming (Platformer) unit, its purity may vary between 70% to 90% H₂. Unionfining units are not designed to operate at hydrogen purities below 70% due to the adverse effect on the catalyst performance and excessive compressor horsepower. The amount of hydrogen required to complete the above reactions will vary depending upon feed and desired product quality.

5.3 Proposed Changes in DHDT :

Diesel Hydro-Treating Unit (DHDT) of Barauni Refinery is having capacity of 2.2 MMTPA and licensed by M/s UOP was commissioned in October 2002.

To produce the Ultra Low Sulphur Diesel (Euro-IV complied), IOCL Barauni Refinery has approached M/s Axens for necessary study @ 3.3 MMTPA unit throughput. M/s Axens has submitted a report for production of Ultra Low sulphur Diesel from DHDT Unit on July 2014. M/s Axens has studied adequacy of the reactor system only for 3.3 MMTPA scenario operation meeting Euro-IV product quality. Barauni has approached PDEC for studying DHDT unit (expect reactor) to operate the unit @ 3.3 MMTPA t'put.

Accordingly, PDEC has checked the adequacy of DHDT Unit @ 3.3 MMTPA operation meeting Euro-IV product quality. For cost estimation purpose, PDEC checked all the equipment (except control valves, PSV's, instruments and P&IDs) and the basic design engineering package with heat & material balance, PFDs, equipment adequacy summary and equipment datasheets is prepared.

5.4 Major Modifications in DHDT

The summary of major modifications @3.3 MMTPA is given below :

1.1 Column : DHDT Stripper Column (702-C-03) internals are required to be changed with high capacity trays.

2.2 Vessels : Modifications in few nozzles are required for CHPS, Flash Drum, Feed surge drum and stripper reflux drum.

1.3 Pumps :

- One identical new pump (702-P-02 C) is required for Feed charge pump services. 2 pumps will be in operation with one pump standby.
- Feed transfer pump (702-P-01 A/B/C) and Diesel product pump (702-P-06 A/B) impellers are required to be changed to maximum possible size.
- One identical new pump (702-P-08 C) is required for total hot feed pump services with philosophy of 2 operating and one standby @ 3.3 MMTPA unit t 'put. However, for partial hot feeding, the requirement of the new pump can be avoided.

1.4 Compressors:

One new Makeup Gas compressor (702-K-XX1) for Make up hydrogen gas.

1.5 Heat Exchanger:

- One additional identical bundle is required for Feed preheat exchanger 702 E-01 (fresh diesel feed vs .diesel product).
- One new exchanger is required for stripper feed vs diesel product exchanger (702-E-51 A/B).

1.6 Furnace:

Preliminary study shows, the furnace is just adequate to meet the required duty (20.0 Gcal/hr for EOR case) within acceptable pressure drop. Furnace vendor/DEC is to be consulted for final rating of the furnace.

1.7 Feed Filter:

BR has already proposed for one additional feed filter bank. However, vendor is to be consulted for the adequacy of the feed filter @ 3.3 MMTPA t'put.

1.8 Coalescer:

Diesel feed and product coalescer are required to be checked by vendors.

1.9 Power Recovery Turbine:

PRT is adequate and will operate at the existing rated capacity. BR may consider the higher capacity PRT in future.