Feasibility Report for BS-VI Fuel Quality upgradation of IOCL, Panipat Refinery 15.0 MMTPA

CLIENT

INDIAN OIL CORPORATION LIMITED

PREPARED BY
ENGINEERS INDIA LIMITED
NEW DELHI

REPORT No. A814-FR-02-41-0001
March 2016

EXECUTIVE SUMMARY

1. EXECUTIVE SUMMARY

1.1. Introduction

Panipat refinery, a unit of Indian Oil Corporation Limited (IOCL) operates a 15.0 Million Metric Tons Per Annum (MMTPA) oil refinery at Panipat in Haryana. The refinery was commissioned in 1997-98 and started off with a crude oil processing capacity of 6.0 MMTPA (PR- Panipat Refinery). The refinery capacity was raised to 12.0 MMTPA with the addition of another crude unit and a full conversion hydrocracker as the secondary processing unit and Delayed Coker unit for bottom processing (PREP- Panipat Refinery Expansion Project). Through progressive revamps and addition of process units the refining capacity has been brought to the present operating capacity of 15.0 MMTPA (PRAEP- Panipat Refinery Additional Expansion Project). IOCL Panipat is also integrated with Naphtha Cracker and Aromatic Complex.

In current refinery operations data corresponding to year 2015, refinery produces Gasoline & Diesel conforming to BS-IV specifications. However with the objective of meeting the guidelines established in Auto Fuel Policy 2025 wherein it would be required to manufacture 100% BS-VI fuels, a study has been carried out (for existing refinery – 15.0 MMTPA) to analyze the potential for conforming to the mandate as described above by 2020 as envisaged by Govt. of India.

In base case of Panipat Refinery, 100% BS-IV grade fuel production has been considered and present operating scenario of Panipat Refinery has been considered.

1.2. Objectives

The major Scope/objectives of this Study are:

1.2.1. The major scope of the study is to identify discreet investments and operating costs associated with the reduction in sulfur content of gasoline and diesel from BS IV compliance to BS VI specification, whilst maintaining the yields of the major products (MS and HSD) and carry out CAPEX cost estimation of revamp / new units at ± 30% accuracy level.

1.2.2. Integration of Petrochemicals streams for production of MS and HSD products.

1.3. Basis of Configuration Study

1.3.1. Refinery throughput

As part of the study, the base case is established initially for 15.0 MMTPA, which corresponds to fuel production conforming to BS-IV specification, i.e. sulfur specification of 50 ppmw for both MS and Diesel.

1.3.2. Crude Mix

The study has been carried out for the design crude mix as per following table-

Table no. 1.3.2.1
Crude Mix

Crudes	Base Case (000 KTPA)	Assay (Reference no.)
Kuwait	3.25	KUWAT304
Basra Hy Basra Lt (Note-1)	4.0	BASH83A BASRA315
IRAN Mix(LS:HS) (75:25)	1.0	IRANL324 IRAH296S
Arab Mix (50:50)	1.25	ARBLRSH332 ARAHV277
Maya	0.75	MAYA204
Forcados	0.5	FRCDS315
Escravos	0.5	ESCRA324
Bonny Light	1.0	BONLT335
Quiboe	0.75	QUAIB3659
Zaffiro	0.5	ZAFRO309
Mangla	1.0	IOCL
Bombay High	0.5	BOMB82A

Note-1 Basra Light with 2.87% S content is considered for the study as per original design basis. However, for Sulphur balance 3.16% S will be considered. Additional margin in the feed sulphur content will be considered during finalization of new Hydrotreating unit like DHDT in line with the Basra Light assay with High Sulphur..

For RFCC Gasoline Selective HDS inlet, max Sulphur value of 2000 ppm in FCC gasoline is considered which is much above the normal Sulphur in RFCC Gasoline Selective HDS feed and no further margin to be considered.

1.3.3. Crude Assay

The crude assay utilized for the study is as per Design Basis. Refer Table 1.3.2.1 above.

1.3.4. Refinery on-stream Hours

8000 hours per annum

1.3.5. Feed and Product Prices

The feed and product prices considered for base case and 100% BS-VI case are provided below in table 1.3.5.1 & 1.3.5.2. The prices are based on average prices during FY 2011-14.

Table 1.3.5.1

Feed Prices

	Price (Rs/MT) (3 years Avg)
Crude	
Kuwait	42517
Basrah Light	41569
Iran Mix (50:50)	IranLight-42486 IranHeavy -42486
Arab Mix (50:50)	Arab Heavy-40775 Arab Light -44135
Forcados	46161
Escravos	46400
Bonny Light	46106
Quiboe	47129
Zaffiro	44634

Mangla	38391
Maya	40456
Bombay High	44908
Other Feed Streams	
C4H from PNCP	52420
Mathura Ref Naphtha to PX-PTA	44296
C-7 to C-8 streams from PNCP	59169
C-9 streams from PNCP	59169
RLNG ($/MMBTU-GCV)	13.48

Table 1.3.5.2

Product Prices

Products (in RS)	
LPG	52420
Propylene	55336
MS-BS-IV Regular	59959
MS-BS-IV Premium	63900
MS-BS-VI Regular	59959
MS BS-VI Premium	63900
SKO	56593
ATF	56098
HSD BS IV	54845
HSD BS VI	54845
HSFO	36601
Bitumen VG30	34625
Sulphur	7788
Coke	10340
Naphtha Export	54180
PNCP Naphtha	46867

PTA	-

1.3.6. Product specifications

MS Specifications

Presently IOCL is producing complete BS-IV Gasoline. The manufacturing specification is indicated below in Table 1.3.6.1.

Table 1.3.6.1
BS-IV Product Specifications

Parameter	Minimum	Product	Maximum
SPG	0.720	0.754	0.773
Sulfur, ppmw		20.0	45
RON	91.4	91.7	
MON	81	>81	
Benzene, max % vol		0.5	0.9
Aromatics, % vol		34.9	35
E-70% vol	11	25.1	38
E-100% vol	41	47.8	70
E-150% vol	75	86.6	
FBP DEG C	85	96.6	
Olefin % vol		8.2	21
Reid vap press, kpa @ 38 c		52.0	60
VLI (10 RVP + 7E-70)		695.7	

As part of up gradation of product qualities to BS-VI, the total gasoline pool is desired to be maintained at same level. The desired product quality is mentioned below in Table 1.3.6.2.

Table 1.3.6.2:
MS product critical specifications,

Sl. No.	Parameters	Unit	BS -VI Spec (Regular) (manufacturing)

1	Density @ 15 ^{0} C	Kg/m3	720-773.7
2	Distillation		
	E-70	% Vol	11-45
	E-100	% Vol	40-70
	E-150	% Vol	75
	FBP	^{0}C max	200
	Residue	% Vol. Max	2
3	Sulphur, Total	mg/kg max	8
4	RON	Min.	91.4 (Note-1)
5	MON	Min.	81.4
6	RVP @ 38 deg C	Kpa	60
7	VLI (10RVP+7E70)		
	Summer (May to Jul)	Max	750
	Others	Max	950
8	Benzene	% Vol-max	1
9	Aromatics	% Vol-max	35
10	Olefin	% Vol-max	21
11	Existent Gum	g/m3-max	
12	Gum(Solvent washed)	mg/100 ml max	5
13	Oxidation Stability	Minutes-Min	360
14	Lead as Pb	g/l-max	0.005
15	Oxygen	%wt-max	2.7

Note-1: In addition to this specification which is applicable to regular grade Gasoline, IOCL has also desired to have estimate of premium grade Gasoline with RON 95/ MON 85 with olefin content of 18 vol % with available blend streams.

Diesel Specifications

Diesel product manufacturing specification is indicated below:-

Table 1.3.6.3
BS-IV Diesel product critical specifications,

Product Qualities		Minimum	Product	Maximum
1.	Specific gravity	0.821	0.82	0.845
2.	Sulfur, ppm		46	46
3.	Flash index		7.1	14.6
4.	Cetane index	46	<49	
5.	Cetane number	51.4	<51.4	
6.	Pour pt index		0.9	3
7.	% Off at 358 c	95	95	
8.	Flash Point, deg- C	35	49.46	
9.	Viscosity, cst@38c	2	2.5	4.5
10.	Viscosity, cst@50c		1.8	
11.	Poly aromatic hydrocarbon		11*	

* Due to presence of Hydrocrackers in the complex, meeting PAH requirement is not envisaged to be a constraint.

In line with the gasoline pool, Diesel pool is also to be maintained at same level with upgraded desired product quality as mentioned in the following table:

Table 1.3.6.4
BS-VI Diesel product specifications,

Sl. No.	Parameters	Unit	BS -VI Spec (Manufacturing)
1	Density @ 15 ^{0} C	Kg/m3	821.5-845
2	Distillation- T -95	^{0}C max	360
3	Sulphur, Total	mg/kg max	8
4	Cetane Number	Min.	51.4
5	Cetane Index	Min.	46
6	Flash Point	^{0}C min	42
7	Kinematic Viscosity at 40 ^{0} C	Cst	2.15-4.5
8	PAH	% wt-max	11
9	Total Contaminants	mg/kg-max	24
10	Oxidation Stability	g/m3-max	21/18
11	RCR on 10% Residue	% wt-max	0.3
12	CCR on 10% Residue	% wt-max	-
13	Water Content	mg/kg-max	200
14	Lubricity corrected WSD	microns-max	420
15	Ash	%wt-max	0.01

1.3.7. Product Limits

As per present refinery operating data, IOCL Panipat Refinery is producing 1601 KTPA of BS-IV Gasoline & 6798 KTPA of BS-IV Diesel. However, as per base case analysis with agreed design crude mix by EIL, the BS-IV Gasoline works out to be 1489 KTPA and BS-IV Diesel 6907 KTPA. The difference may directionally be due to following:-

- Difference in operating crude mix & design basis crude mix
- Difference in Unit capacity utilization as in the study vis-à-vis actual. For e.g. operating capacity of HCU & OHCU is higher than that considered for design basis.

As a part of study for BS-VI production target of producing maximum Gasoline (min 1489 KTPA) & max diesel 6907 KTPA has been considered which is in line with retrofitted base case.

In addition to meeting the present Diesel pool quantity, as a part of BS-VI study, the maximum Kerosene upgradation to Diesel allowed by specification is being considered. The ATF product is to be maintained as per present demand.

1.3.8. Other considerations

Following are the general considerations taken for this study-

1) Sustainable operating capacity and product specifications to be considered for the base case for the existing units. Refer Section 1.3.6 for the same.
2) The likely impact of the planned projects or projects under implementation has been considered as per IOCL input.
3) No blend stock / stream from any other refinery or market is considered for meeting BS-VI specifications except naphtha from Matura refinery for Aromatics complex (150 KTPA).
4) Streams from Aromatics and Olefins complex respectively for blend pool are taken as per IOCL input.

1.3.9. Plant Fuel System

RLNG shall be utilized as:

- Feed and fuel to the existing as well as new hydrogen unit
- Fuel for GT's
- Refinery fuel pool option

Refinery Fuel gas will be used as plant fuel and the balance demand shall be met by low sulphur i.e. 0.5 wt% S fuel oil generated in the refinery.

1.4. Configuration Study Approach

a) Development of Base LP model of refinery to produce 100% BS-IV products for the existing capacity of 15 MMTPA based on the IOCL input data on present operating unit capacity, feed and product properties, blend streams and utility infrastructure. The fuel requirement as in present refinery has been aligned as part of Base case.

b) While going for production of BS-VI gasoline and diesel production, the main objective is the reduction of sulphur specification in the blend components and considering maximum Kerosene upgradation to Diesel.

Measures to be taken for the same, is mentioned in following Section.

1.5. Configuration result for 100 % BS-VI case

Based upon Details / analysis of individual streams in various units as agreed in design basis and further information (and as per discussions with IOCL), following are the conclusion regarding BS-VI Gasoline and Diesel blends and other associated units having impact due to up gradation.

1.5.1. Gasoline pool up-gradation to BS-VI specifications

The present Gasoline blend streams are identified in the table below.

Table 1.5.1.1

Blend pool for BS-IV grade gasoline pool (existing)

Stream	Qty, KTPA	Wt%	SUL, ppmw	RON
C7-C8 Stream from PNCP	180	12.08	0.1	100

Heavy Reformate	552.7	37.1	1	98
Isomerate	391.7	26.29	0.3	85.5
RFCC Gasoline Selective HDS Product	321.7	21.59	80	89.5
Raffinate	43.7	2.94	0.22	74
Total	**1489.7**	**100**	**18.0**	**91.8**

It is observed that only RFCC Gasoline Selective HDS unit product is the high Sulphur carrying streams in the gasoline pool. For BS-VI, this will warrant a revamp in existing RFCC Gasoline Selective HDS unit. Possible revamp options for RFCC Gasoline Selective HDS unit have been discussed with original licensor M/s Axens and following are the outcome:-

1. Splitter column revamping: Splitter cut-point to be adjusted to reach new LCN sulphur specifications (from 100 wppm to 30 wpmm).
2. Implementation of a Second Stage HDS Section: In order to perform a deeper hydrodesulphurization on HCN cut while maximizing octane retention. The Second Stage HDS section arrangement is similar to the existing HDS Reaction Section:
    - This includes a reactor, fired heater, feed/effluent exchangers, separator drum and new recycle gas loop (compressor, amine absorber, KO drums).
    - H2S removal from hydrocarbons between the First Stage HDS Reaction Section and the Second Stage HDS Section will be achieved in a new H2S Stripper column using part of the recycle gas from a new Recycle Gas Compressor.
3. In order to minimize CAPEX as well as the plot space area, spare compressor for First and Second Stage HDS Reaction Sections can be mutualized.
4. Re-use of the existing Finishing Reactor as one bed of Second Stage HDS Reactor. The existing finishing reactor is reused to reach the catalyst volume required for the future operation.
5. New reactor in second HDS stage: for second bed of second stage HDS reactor (Can be duplicated from existing finishing reactor)

This RFCC Gasoline Selective HDS revamp including a second stage HDS reactor will result in the complete BS-VI grade production which will reduce RFCC Gasoline Selective HDS product sulfur specification from 100 wppm to 30 wppm.

During the project execution activity, a more stringent specification of outlet sulfur in the RFCC Gasoline Selective HDS product can be aimed for to achieve further margin on the BS VI sulfur specification, however, taking the RON loss.

With only revamp of RFCC Gasoline selective HDS and available blend components the following regular gasoline pool in line with present refinery product profile is possible to be met.

Table 1.5.1.1

Blend pool for BS VI Regular grade gasoline

Sl. No.	Gasoline Blend Streams	KTPA	Sulfur, ppmw	RON	Olefins, wt%	Aromatics, wt%	RVP, kPa
1.	C7-C8 from PNCP	180	0.1	100	0	82	58.82
2.	Raffinate from Aromatic Complex	40	0.22	74.50	0	0.03	88.23
3.	Hy. Reformate	553	1	98.00 (Note-1)	0.10	68	9.80
4.	Isomerate	392	0.3	85.5	0.10	0	104.9
5.	RFCC Gasoline Desulphurized Product	322	30	87.4 (Note-2)	47.00	12.0	22.59
	Total	**1485**	**7.0**	**91.4**	**10.7**	**34.7**	**51.2**

Note-1 As per present operating data IOCL indicated RON of heavy reformate as 96. Considering design value of 100, final RON of 98 has been considered, which has been agreed with IOCL.

Note-2 As per original licensor input a RON loss of 2.6 across two stage hydrodesulphurization reactor RFCC Gasoline selective HDS

The Feasibility report has been worked out on the above basis, where the existing FCC Gasoline desulfurizer unit revamp has been considered. This option produces regular BS VI grade pool and no substantial quantity of premium grade can be produced with the available blend stocks. The cost has been considered as per licensor input pending details of modification, which can be substantiated at the Project execution stage.

1.5.2. Observation on FCC Gasoline HDS unit:

It is to be noted that a very extensive modification has been envisaged for the existing Prime G + unit which will include addition of new heater. New reactor, recycle gas compressor and columns in addition to modification in the existing equipment.

This extensive modification may call for extensive shut down, space and constructability issues which have to be reviewed in details during the execution stage with the detailed information from selected Licensor.

Considering that, an alternative option of a grass root FCC Gasoline selective hydrotreater to meet the requirement also can be reviewed. This unit can be located in a virgin area and can be constructed easily without any extended shutdown and production loss. The penalty in Capex is expected to be of the order of INR 86.8 Crores (Plant and Machinery) between the revamp and grass root unit.

1.5.3. Observation on production of premium gasoline:

As mentioned, no substantial quantity of premium gasoline is feasible to be made within the refinery with the available blend streams. The possible options of production of premium grade are:

- By considering external blend stream such as Ethanol: It is possible to meet part premium pool by using these streams. By limiting the Ethanol blend to 5% vol, 25% Gasoline pool can be converted to premium grade with ramification on the overall gasoline production.

However the modality of Ethanol blending to be reviewed in details with IOCL in line with their operating and marketing philosophy.

- By installing additional process units to produce additional octane boosting streams such as TAME or Alkylate in a captive way. For this case the following alternative options are possible

1.5.2.1 Alternate Option-1

RFCC Gasoline Selective HDS revamp with a TAME unit

For determining potential for producing 95 octane MS, an option of TAME unit has been assessed. TAME unit, if considered will provide benefit of 1 RON number in the Prime G Product with overall increase of 3% in Gasoline overall Yield. It can be noted that already the Regular Gasoline Pool is limited by RON and that impact of the TAME unit in Gasoline pool will not be significant thus possibility of producing substantial quantity of 95 octane premium MS doesn't exist by only TAME unit.

1.5.2.2 Alternate Option-2

RFCC Gasoline selective HDS revamp with an Alkylation unit

Panipat refinery with an existing RFCC and DCU has feed potential for an alkylation block. As per preliminary calculations, it is possible to produce around 60 KTPA of alkylate from RFCC & DCU LPG.

This case will directionally lead to reduction of LPG yield from the refinery and will result in production of around 13.% of Premium grade Gasoline. The estimated Gasoline pools are as follows

Blend pool for Gasoline for this case is as follows-

Table 1.5.1.2

Blend pool for Regular/Premium grade gasoline-Option C

Case-C							
Sl. No.	Gasoline Blend Streams	KTPA	Sulfur, ppmw	RON	Olefins, wt%	Aromatics, wt%	RVP, kPa

		Regular	Premium					
1.	C7-C8 from PNCP	140	40	0.1	100	0	82	58.8
2.	Raffinate from Aromatic Complex	20	-	0.22	74.50	0	0.03	88.2
3.	Hy. Reformate	488	64	1	98	0.10	68	9.80
4.	Isomerate	352	40	0.3	85.5	0.10	0	104.9
5.	RFCC Gasoline Desulphurized Product	321	24	30	87.40	47.00	12.0	22.6
6.	Alkylate	-	60	Neg.	96.0	1.00	2.5	37.9
Total Regular		**1322**		**8**	**91.4**	**12.1**	**33.7**	**50**
Total Premium			**203**	**1**	**95**	**1**	**35**	**50**

It can be noted that the premium pool is limited by the Aromatics and RON Number.

It is to be noted that the present Feasibility Study does not include the expenditure towards Alkylation unit

1.5.3 Diesel pool up-gradation to BS-VI specifications

Presently Panipat refinery produces BS-IV diesel of 6798 KTPA. After rationalization as per design crude mix and individual unit capacity / yield as per design basis the base case Diesel production works out to be 6907 KTPA along with ATF production of 1125 KTPA and kerosene production of 870 KTPA. The diesel pool components for BS-IV base case are described below

Table 1.5.2.1

Blend pool for Diesel BS-IV (existing) Scenario

	KTPA	SUL, ppmw	Density	Cetane Index	flash pt deg C
C9+ STREAM	100	141	916	30	38

SR Hy Nap from CDU1	111.3	150	788.7	30.2	43
SR KEROSENE	258.3	503	823.9	37	66
SR L.DIESEL	90.3	1060	863.3	43.79	105
OHCU HV. NAPHTHA	119.7	1	750	30	15
OHC KEROSENE	371	5	792	49	45
OHCU DIESEL	718.7	5	837	60	66
DHDS DIESEL	679.3	21	833	51.06	70
KERO (HCU)	257.7	5	802	44.5	38
DIESEL (HCU)	724.3	15	838	52.8	66
DIESEL (DHDT)	3456	8	825.7	59.1	50
PX HEAVY AROM.	20.3	0.1	934.4	30	38
Total	**6907**	**46**	**820**	**>49**	**49.46**

As part of BS VI study, Diesel pool component including upgradation of maximum kerosene stream results in finding as follows -

1. The total diesel production will be 7750 KTPA, which considers upgradation of 870 KTPA of kerosene.
2. All the straight run & cracked streams will necessarily have to be treated before blending into diesel pool of BS-VI.
3. Existing DHDS unit is designed for 30 ppm outlet sulphur. With catalyst change as planned by IOCL , is it supposed to achieve 10 ppmw sulphur. To achieve 8 ppm in this unit, catalyst change with de-rating of the unit is considered. Preliminary assessment indicates that the achievable unit capacity with change of catalyst and at the most a change of internal in case warranted by the Licensor at execution stage will be 550 KTPA.
4. IOCL apprised that Existing DHDT Diesel Sulphur hovers around 15ppmw due to equipment/maintenance issues presently encountered in the unit. However, as per mutual agreement EIL has considered that these mechanical issues will be sorted out and DHDT will be able to produce 8 ppmw Sulphur product at

design capacity with change of catalyst at the most a change of internal in case warranted by the Licensor at execution stage. Existing design capacity has been retained for this case.

5. To accommodate the additional streams which could earlier be blended into diesel pool and balance capacity of DHDS a new DHDT unit of 1920 KTPA is required meeting outlet specification of 8 ppmw Sulphur which includes the estimated kerosene upgradation.
6. The C-9 Stream from PNCP block contains large amount of styrenics, indenes and DCPD therefore a pre-treatment of this stream is considered before it is processed in New DHDT unit. Pre-treatment is considered to be part of new DHDT unit, hence, CAPEX implication for pre – treatment of C-9 stream is considered in DHDT unit.
7. **However, the design capacity of the unit has been finalized as 2200 MMTPA in consultation with IOCL** to take care of -
   a. Variations in crude design mix and opportunity crude processing having higher middle distillate potential.
   b. Reprocessing requirements, in case of exigencies in existing units.

Blend pool for BS-VI diesel is as follows-

Table 1.5.2.2

Blend pool for Diesel BS-VI Scenario

Sl. No.	Diesel Blend Streams	KTPA	Wt%	SPG	Sulphur, ppmw	Cetane Index	Flash Pt., Deg C
1.	HCU Heavy. Naphtha	79	1.0	0.768	4.0	18	10.0
2.	OHCU Diesel	719	9.2	0.837	5.0	60	66.0
3.	DHDS Diesel	534	6.9	0.8201	8.0	59.5	70.0
4.	HCU Kero	383	4.9	0.801	5.0	44.5	38.0
5.	HCU Diesel	724	9.3	0.838	10.0	52.8	66.0

6.	DHDT Diesel	3417	44.0	0.8133	8.0	59.5	50.0
7.	Hy. Aromatics	20	0.2	0.934	0.1	30	38.0
8.	New DHT Diesel	1874	24.2	0.8287	8.0	51.5	50.0
Total		**7750**	**100**	**0.8215**	**7.7**	**>49**	**48**

1.5.4 Finding on other Units-

1.5.4.1 Hydrogen Generation Unit (HGU)

The estimate of H2 consumption has been carried out considering existing infrastructure in the refinery on requirement of H2 in Hydrotreating unit and recovery from other H2 rich streams.

A new H2 generation unit of 32 KTPA will be required in case desired margin on overall requirement is considered.

As per operating feed back, the maximum operating capacity of HCU and OHCU are 1900 & 2100 KTPA respectively. Though credit towards the same has not been considered in the configuration, considering the additional requirement for these units in hydrogen demand, HGU capacity requirement will be ~44 KTPA.

It is to be noted that existing HGU-I PSA capacity is limited to 38 KTPA which includes H2 rich gas from CCR and OHCU & HCU. Therefore additional capacity of 20 KTPA will be considered in PSA with new HGU. The PSA Capacity shall be 64 KTPA. The present feasibility study includes the above option.

1.5.4.2 SRU

Presently the refinery Sulphur recovery units are having following design capacity-

SRU-I - 2 X 115 TPD (common incinerator allowing only one SRU train to operate)

SRU-II – (2+1) X 225 TPD

The estimated operating capacity of new SRU required is about 125 TPD, retaining the existing philosophy of a standby SRU train in PREAP unchanged. An additional SRU chain of 225 TPD is envisaged to fulfill the requirement for handling additional Sulphur rejection considering the following:

- Flexibility of inter-changeability
- Flexibility to operate with 100% HS Crude in block out mode.

TGTU

A New TGTU of 225 equivalent Capacity along with New SRU train is considered.

The present Feasibility Study includes the additional SRU provision

1.5.4.4 Alternative option of Sulphuric Acid production unit instead of SRU:

With the recently developed technologies, it is possible to evaluate another option of Sulfuric acid unit instead of SRU which will also generate a substantial quantity of high pressure steam to add to it's advantage. With the available additional sour gas a WSA unit of 375 TPD (Design 400 TPD) will be feasible.

This unit can be located in a separate area. The Capex for the 400 TPD WSA plant is estimated to be INR 234.5 Crores.

In case owner decides so, the associated requirement (e.g., utilities etc.) for the sulphuric acid plant can be established in the next stage of the Project.

1.5.4.5 SWS

There will be an additional generation of 10m3/hr of sour water from new facilities. A two stage SWS of capacity 16.4 m3/hr (in line with existing Hydroprocessing SWS-II unit) is considered as part of feasibility report.

1.5.4.6 ARU

There will be additional generation of 84 TPH of rich amine which needs to be recovered. An amine regeneration system is envisaged for the same as part of feasibility report.

1.5.4.7 Amine Treating Unit

As per IOCL the Existing ATU's are operating at maximum capacities and there are no margins available for treating additional Fuel Gases.

For additional fuel gas treatment under BS VI operation scenario existing a New LP Fuel Gas Amine Treating Unit of 6 TPH Fuel Gas Treating capacity is envisaged.

1.5.4.8 New Process Units

A list of new process units required for BS-VI upgradation is as follows-

Table 1.5.3.3

New Process Units

	Unit	Capacity (Design)	Operating Capacity
1.	DHDT	2200 KTPA	1920 KTPA
2.	HGU	44 KTPA	44 KTPA
3.	Sour water Stripper (2 stage)	16.4 m3/hr	10 m3/hr
4.	ARU (MDEA based)	84 TPH	84 TPH
5.	SRU + TGTU	225 TPD	125 TPD
6.	FG Amine Treating Unit	6 TPH	6TPH

1.6 Material Balance

Table 1.6.1

Material Balance for BS-VI Scenario

FEED	Units	Weight (KTPA)
KUWAIT	KTPA	3250.00
FORCADOS	KTPA	500.00
ESCR	KTPA	500.00
BONNYLIGHT	KTPA	1000.00
QUAIBOE	KTPA	750.00
ZAFIRO	KTPA	500.00
BOMBAY HIGH	KTPA	500.00
ARAB MIX	KTPA	1250.00
IRAN MIX	KTPA	1000.00
Basarah 90L10H	KTPA	4000.00
Mangala	KTPA	1000.00
MAYA	KTPA	750.00
LNG	KTPA	266
TOTAL CRUDES		15206
MATHURA REF. NAPHTHA	KTPA	150.00
C7-C8	KTPA	180.00
C9+ STREAM	KTPA	100.00
PFO (PNCP)	KTPA	112.00
C4H (PNCP)	KTPA	200.00
Other Feed		742
TOTAL FEED		**15948**
Product Sales	**Units**	**Weight**
MIXED LPG'S	KTPA	649.0

RFCC PROPYLENE	KTPA	112.0
BS-VI REG. GASOLINE	KTPA	1485.0
BENZENE	KTPA	20.3
PTA SALES	KTPA	553.0
PNCP FEED	KTPA	1372.0
KEROSENE	KTPA	0.00
JET	KTPA	1125.00
BS-VI DIESEL	KTPA	7750.0
HIGH SUL.F.OIL	KTPA	225.0
BITUMEN	KTPA	360.00
COKE	KTPA	874.0
PRODUCT SULPHUR	TPD	652

1.7 Operating Capacities at a glance – BS VI Scenario

Operating & Design capacities of Major Process units for BS-VI scenario is listed in following table-

Table 1.7.1

Unit Capacities

S.NO.	UNIT NAME	CAPACITY	
		Design MMTPA	Operating MMTPA
1	CDU, VDU	7.5	7.5
3	OHCU	1.9	1.9
4	HGU	0.038	0.038
5	RFCCU	0.85	0.85

6	PSU	0.255	0.232
7	CRU	0.64	0.64
8	DHDS	0.7	0.55
9	VBU	0.4	-
10	BBU	0.5	0.36
14	SRU-I	115 TPD	115 TPD
15	SRU-I	115 TPD	-
16	ARU	400M^3/HR	400M^3/HR
17	SWS-I (REFINERY)	71.8 M^3/HR	71.8 M^3/HR
18	SWS-II (HYDROPROCESSING)	16.4 M^3/HR	16.4 M^3/HR
PANIPAT REFINERY EXPANSION (PRE) UNITS			
S. NO.	**UNIT NAME**	**CAPACITY**	
		Design MMTPA	**Operating MMTPA**
1	CDU, VDU	6.00 MMTPA + 35%	7.5
3	HGU	2 X 70 KTPA	0.140
4	HCU	1.7 MMTPA	1.7
5	DCU	3.0 MMTPA	2.9
7	DHDT	3.5 MMTPA	3.5
8	SRU-II	2 X 225 TPD	450
9	TGU	EQ.450 TPD SULPHUR	EQ.450 TPD SULPHUR

10	ARU	410 M^3/HR	410 M^3/HR
11	SWS-I (REFINERY)	170 M^3/HR	170 M^3/HR
12	SWS-II (HYDROPROCESSING)	40 M^3/HR	40 M^3/HR
MS QUALITY UPGRADATION (MSQ) UNITS			
S. NO.	**UNIT NAME**	**Design MMTPA**	**Operating MMTPA**
1	NHT /PENEX	410 / 400 TMTPA	0.41/0.4
PANIPAT REFINERY ADDITIONAL EXPANSION PROJECT (PRAEP) UNITS			
S. NO.	**UNIT NAME**	**CAPACITY**	
1	SWS	40 M^3/HR	40 M^3/HR
2	SRU	225 TPD	225 TPD
3	TGU	EQ.450 TPD SULPHUR	EQ.450 TPD SULPHUR

1.8 Utility Systems

Following additional utilities are envisaged along with new facilities-

Table 1.8.1

Additional Utility Requirement & systems

Utility	Unit	Quantity	Remarks
Power	KW	10300	*Can be met through existing facilities. However, spinning reserve will be reduced which can be managed by either change in operating configuration from 4GT+2TG to*

			5GT+1TG or by importing power from grid. No CAPEX considered due to same.
Steam			
HP	TPH	-32 (generation)	No additional hardware envisaged.
MP	TPH	22	-do-
LP	TPH	8.0	-do-
Cooling Water	M3/hr	2500	*Additional CT of 4000 m3/hr (1+1) and pump of 2*4000 m3/hr will be considered near the DHDT/ HGU area.*
Instrument Air	Nm3/hr	950	*One additional compressor of 6500 m3/hr and drier of 5000 nm3/hr will be considered*
Plant air	Nm3/hr	140	No additional hardware
Condensate handling	TPH	18	-do-
Nitrogen	Nm3/hr	6681 (Max. intermittent requirement for DHDT start-up) Continuous requirement: Nil	*Will be met from existing facility considering no simultaneous shutdown/start-up of existing DHDT/HCU/OHCU and the new DHDT.*

Surface condensate and power condensate to be routed to existing system (CPU/ Dearator)

The utilities are estimated as per following basis-

- Requirement for following units as per their design capacities.

DHDT	2200 KTPA
HGU	44 KTPA
Revamp of Existing Prime – G	(Revamp)

Sour water Stripper (2 stage)	16.4 m3/hr
ARU (MDEA based)	84 TPH
SRU	225 TPD
New Cooling Towers	4000 m3/hr (For the DHDT/ HGU block)
New Compressed air system	6500 Nm3/hr
Additional offsites facilities	DHT feed storage and pumping system

1.9 Offsite System

The offsite system shall be augmented by adding following storages and pumping systems-

Table 1.9.1
New Storage Tank

Sl no.	Service	No of tanks	Type	Liquid stored capacity of each tank
1	Intermittent Feed storage tank for new DHDT	1	Floating Roof	20,000 m3

Table 1.9.2
New Pumps

Sl no.	Service	No of pumps	Flow (m3/hr)	Type
1	New DHDT unit feed pumps	1W + 1S	250	Centrifugal motor driven

1.10 Flare System

Additional loads will be generated from the new DHDT, HGU and incremental load from FCC Gasoline HDS. As IOCL has informed that existing Flare system is already operating at its saturation level post PRAEP. Therefore, additional flare system has been considered under BS VI project. For this additional Flare load a New Flare Stack of size 64” and associated systems / piping is included in CAPEX. However, during execution a detailed flare adequacy needs to be carried out with mitigation philosophies adopted in new units as well as some of existing units to estimate the requirement of Flare system.

1.11 ETP

As part of additional facility additional effluent mostly from Sour water stripper (approx 5 m3/hr, intermittent) will be generated. It is considered that existing ETP will be adequate to handle this additional amount. Additional Cooling Tower blow down (15 m3/hr), Boiler blow down (1 m3 / hr from SRU), OWS and CRWS will also be accommodated in existing ETP system. A detailed analysis can be carried out at the execution stage.

1.12 Control Room and Substation

New Control Room and Substation has been considered along with BS VI upgradation project.

1.13 CAPEX ESTIMATION

The CAPEX required for new units and revamp of existing units has been estimated to be Rs 2745.15 Crores with a cost accuracy level of +/- 30%. The cost estimation is subjected to the following stipulations and based on the existing Prime G + revamp and new SRU+ TGTU.

- Presently the CAPEX estimate has been carried out for the Process units utilities and offsites facilities.
- The new DHDT CAPEX has been worked out based on in house information considering the selected technology for recently concluded evaluation. A variation in unit CAPEX upto +25% has been observed amongst competing technologies.
- DHDT/HGU Cost is only for B/L of Process unit which excludes substation, control room, interconnection with existing plant facilities such as fire water, cooling water, Instr. Air, Plant air etc.
- The capex estimate for the RFCC gasoline selective HDS unit is based on the Licensor inputs obtained for the subject units. The cost estimate is worked out on the assumption that it will be feasible to carry out the

suggested hardware modifications within stipulated schedule / shutdown time. No constructability study, has been carried out at this stage to establish the feasibility. Further inputs from the Licensor on the details of the changes is awaited.

1.14 PLOT PLAN

Overall Plot Plan of Panipat refinery has been reviewed along with IOCL with respect to inclusion of new units envisaged as part of BS VI Fuel Upgradation and Location of New units have been identified and marked-up in the latest Plot Plan.

Indian Oil Corporation Limited
Panipat, Haryana, India

Px Expansion Project
Process Revamp Study Report
UOP Project Number 9000352-359

October, 2015

A Honeywell Company
25 East Algonquin Road
Des Plaines, Illinois 60017-5017
U.S.A.

Table of Contents

1. Executive Summary

1.1. Overview

Indian Oil Corporation Limited ("IOCL") is operating a UOP designed and Licensed Aromatics Complex at its refinery located in Panipat, Haryana, India. IOCL would like UOP to assess the feasibility of processing C8 Aromatics recovered from the Pygas Splitter along with the Naphtha Feed in the Aromatics Complex for increasing PX production to 460 KMTA based on 8000 hrs annual operating. To assist IOCL in evaluating this project, UOP has conducted a Process Revamp Study to identify the adequacy of major pieces of equipment and recommend necessary changes to meet the objective

The results of this study will provide IOCL with information to help them complete preliminary project economics and establish a firm design basis and direction for further engineering work. UOP is positioned to use the study results as a basis in developing a Revamp Schedule A for the complex at this capacity The basis and scope of the study are described below.

1.2. Objectives / Background

The main objective of the project is to increase Px production to 460 KMTA from available current Naphtha feed and additional feeds from the Pygas Splitter. The Study will identify the major modifications required to achieve the increased capacity by targeting minimum possible revamp capital cost.

The Aromatics Complex at IOCL Panipat Refinery was designed by UOP in 2000 and consists of the following process Units:

- UOP Naphtha Hydrotreating Process Unit ("NHT Unit"), UOP project no 903292.
- CCR Platforming Process Unit ("Platforming Unit"), UOP project no 903293.
- Sulfolane Process Unit ("Sulfolane Unit"), UOP project no 903295.
- Tatoray Process Unit ("Tatoray Unit"), UOP project no 903340
- Benzene -Toluene Fractionation Unit ("B-T Fractionation Unit"), UOP project no 903341.
- Parex Process Unit ("Parex Unit"), UOP project no 903297.
- Isomar Process Unit ("Isomar Unit"), UOP project no 903299.
- UOP Xylene Fractionation Unit including Reformate Splitter, Clay Treater, Xylene Column, and Heavy Aromatics Column ("Xylene Fractionation Unit"), UOP project no 903298.

The NHT and Platforming units constitute the "front end" of the Aromatics Complex. The NHT and the Platforming units are not included in the scope of this report. Separate study reports were made for these units. The Sulfolane, Tatoray, B-T Fractionation, Parex, Isomar and Xylene Fractionation units constitute the "back end" of the Aromatics complex.

The original design objective of the Aromatics Complex was to process 500 KMTA of a heart cut naphtha derived from Bombay High or Bonny Light Crude, while producing 360 KMTA of Paraxylene and Benzene commensurate to the heart cut naphtha feed composition. The Aromatics Complex was commissioned between May to August, 2006 and has been operating since then.

1.3. Basis / Approach

The process revamp study for the Aromatics complex is done for a single case considering the below catalysts.

- R-334 catalyst in the Platforming unit.
- TA-30 catalyst in the Tatoray unit.
- ADS-47 Parex adsorbent in the Parex unit.
- I-400 EB-isomerization catalyst in the Isomar unit

Paraxylene production of 460 KMTA and HA Column Bottoms production of 20 KMTA is targeted with 8000 operating hrs/year.
The three feed stocks to the Aromatics complex are as below

- Reformate from Debutanizer Bottoms @ 565 KMTA
- Pygas Splitter Bottoms Stream @ 57KMTA
- Pygas Splitter Overhead Stream @ 23 KMTA

A CCR Platforming yield estimate was first generated for the Feed Case considering the Feed compositions of the Panipat and Mathura streams provided to UOP. The feed to the PX Complex was taken from this yield estimate which is the CCR Platforming Unit – Debutanizer bottoms or reformate stream. A yield estimate for the PX Complex then was generated using the reformate and the Pygas Splitter Bottoms and overhead streams The yield estimate numbers are as follows:

- CCR Platforming YE – P065487
- Aromatics PX Complex YE – AC15274

The yield estimate for the Aromatics complex is based on the addition of a new BT Splitter Column upstream of the Reformate Splitter in the Xylene Fractionation unit.

The study consisted of the review of all major equipment including Fired Heaters, Reactors, Fractionators, Combined Feed exchangers, major Vessels, major Heat Exchangers, major Pumps and Compressors. The equipment evaluation was performed based on UOP's simulation of the operation of the process units at the revamp case conditions, and the ability of the existing equipment to meet those requirements based on as-built equipment data sheets and drawings provided by IOCL

IOCL provided the existing equipment data for the study. These included general vessel arrangement drawings and layout drawings of trays, vendor data sheets and construction drawings for heat exchangers, pump API sheets, pump curves and pump motor data sheets, as well as Piping & Instrument Diagrams for the Unit. Revision 0 of the Basic Engineering Design Questionnaire ("BEDQ") dated July 2, 2015, was used for utility system values.

The overall basis, scope and deliverables in this report are in accordance with the Engineering Agreement numbered 15A0075 and dated March 10, 2015 between IOCL and UOP LLC included as Appendix 9.6.

1.4. Summary of Results

The following equipment list shows the major equipment considered for the operating cases and indicates the status of each item. Please note that minor equipment such as instruments, pressure relief valves, injection pumps and utilities are not considered a part of this Study.

(OK = acceptable, NG = not good, NA = not applicable)

List of major equipment and equipment status

1.4.1. Sulfolane Process Unit

Column				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Trays**	**Revamp Status**
204-C1	Extractor	OK	OK	Adequate
204-C2	Raffinate Water Wash Column	OK	NG	New jet-deck trays required
204-C3	Stripper	OK	OK	Adequate
204-C4	Recovery Column	OK	OK	Adequate
204-C5	Water Stripper	OK	OK	Adequate

Vessel

Item Number	Equipment Name	Design Temp. / Pressure	Residence Time	Vapor/ Liquid Separation	Revamp Status
204-V2	Stripper Receiver	OK	OK	OK	Adequate
204-V3	Stripper Receiver	OK	OK	OK	Adequate
204-V4	Recovery Column Receiver	OK	OK	OK	Adequate

Exchanger				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Duty/ Pressure drop**	**Revamp Status**
204-AC1	Stripper Condenser	OK	OK	Adequate
204-E1	Lean-Rich Solvent Exchanger	OK	OK	Adequate
204-E2	Raffinate Cooler	OK	OK	Adequate
204-E3	Stripper Reboiler	OK	OK	Adequate
204-E4	Stripper Trim Condenser	OK	OK	Adequate
204-E5	Recovery Column Reboiler	OK	OK	Adequate
204-E7	Solvent Regenerator Reboiler	OK	OK	Adequate
204-E8	Water Stripper Reboiler	OK	OK	Adequate
204-E9	Recovery Column Condenser	OK	OK	Adequate
204-E12	Raffinate Rundown Cooler	-	-	New

Pump				
Item Number	**Equipment Name**	**Capacity**	**Head**	**Revamp Status**
204-P1A/B	Extractor Charge Pumps	OK	OK	Adequate
204-P2A/B	Raffinate Wash Water Recycle Pumps	OK	OK	Adequate
204-P3A/B	Raffinate Pumps	OK	OK	Adequate
204-P4A/B	Stripper Bottoms Pumps	OK	OK	Adequate
204-P5A/B	Stripper Overhead Water Pumps	OK	OK	Adequate

204-P6A/B	Extractor Recycle Pumps	OK	OK	Adequate
204-P7A/B	Lean Solvent Pumps	OK	OK	Adequate
204-P9A/B	Wash Water Pumps	OK	OK	Adequate
204-P10A/B	Recovery Column Overhead Pumps	OK	OK	Adequate

1.4.2. Tatoray Process Unit

Column				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Trays**	**Revamp Status**
208-C1	Stripper	OK	NG	Bottom section trays need replacement

Fired Heaters				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Duty**	**Revamp Status**
208-F1	Tatoray Charge Heater	OK	OK	Adequate

Vessel					
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Residence Time**	**Vapor/ Liquid Separation**	**Revamp Status**
208-V1	Feed Surge Drum	OK	OK	OK	Adequate
208-V2	Separator	OK	OK	OK	Adequate
208-V3	Stripper Receiver	OK	OK	OK	Adequate

Reactor				
Item Number	**Equipment Name**	**Design Temp / Pressure**	**Flow Distribution/ Pressure drop**	**Revamp Status**
208-R1	Reactor	OK	OK	Adequate

Exchanger				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Duty/ Pressure drop**	**Revamp Status**
208-E1	Combined Feed Exchanger	OK	OK	Adequate
208-E2	Product Trim Condenser	OK	OK	Adequate
208-E3	Stripper Feed-Bottoms Exchanger	OK	OK	Adequate
208-E4	Stripper Reboiler	OK	OK	Adequate
208-E5	Stripper Feed-Overhead Exchanger	OK	OK	Adequate
208-E6A/B	Stripper Trim Condenser	OK	NG	Replace with larger shell
208-E7	Feed Tank Cooler	OK	OK	Adequate

Air Cooled Exchanger				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Duty/ Pressure drop**	**Revamp Status**
208-AC1	Product Condenser	OK	OK	Adequate
208-AC2	Stripper Condenser	OK	OK	Adequate

Compressor				
Item Number	**Equipment Name**	**Capacity**	**Discharge Temperature**	**Revamp Status**
208-K1	Recycle Compressor	OK	OK	Adequate

Item Number	Equipment Name	Capacity	Head	Revamp Status
208-P1A/B	Charge Pumps	NG	NG	New pump and Motor required
208-P2A/B	Stripper Auxiliary Reboiler Pumps	OK	OK	Adequate
208-P3A/B	Stripper Overhead pumps	NG	NG	New pump and Motor required

1.4.3. UOP Benzene Toluene Fractionation Unit

Column

Item Number	Equipment Name	Design Temp. / Pressure	Trays	Revamp Status
209-C1	Benzene Column	OK	OK	Adequate
209-C2	Toluene Column	OK	NG	Replace with PFMD Trays

Vessel

Item Number	Equipment Name	Design Temp. / Pressure	Residence Time	Vapor/ Liquid Separation	Revamp Status
209-V1A/B	Clay Treaters	OK	N/A	N/A	Adequate
209-V2	Benzene Column Receiver	OK	OK	OK	Adequate
209-V3	Toluene Column Receiver	OK	OK	OK	Adequate

Exchanger- Shell & Tube

Item Number	Equipment Name	Design Temp. / Pressure	Duty/ Pressure drop	Revamp Status
209-E1	Clay Treater Charge Exchanger	OK	OK	Adequate
209-E2	Clay Treater Charge Heater	OK	OK	Adequate
209-E3	Benzene Column Reboiler	OK	OK	Adequate
209-E4	Benzene Product Cooler	OK	NG	Add identical shell in series
209-E5	Toluene Column Reboiler	OK	NG	New

Exchangers- Air Cooled

209-AC1	Benzene Column Condenser	OK	OK	Adequate
209-AC2	Toluene Column Condenser	OK	NG	Add additional bay in parallel

Pump

Item Number	Equipment Name	Capacity	Head	Revamp Status
209-P1A/B	Clay Treater Charge Pumps	OK	OK	Adequate
209-P2A/B	Benzene Column Bottoms	OK	OK	Adequate

	Pumps			
209-P3A/B	Benzene Product Pumps	OK	OK	Adequate
209-P4A/B	Benzene Column Reflux Pumps	OK	OK	Adequate
209-P5A/B	Toluene Column Bottoms Pumps	NG	NG	New pump and Motor would be required
209-P6A/B	Toluene Column Overhead Pumps	OK	OK	Existing pump to be used for Net Overhead service only
209-P7A/B	Toluene Column Reflux Pumps	NG	NG	New Pump and Motor to be installed for Reflux service Only

1.4.4. Parex Process Unit

Column				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Trays**	**Revamp Status**
205-C1	Raffinate Column	OK	OK	Re-Tray with ECMD Trays
205-C2	Extract Column	OK	OK	Adequate
205-C3	Finishing Column	OK	NG	Re-tray bottoms section
205-C4	Desorbent Rerun Column	OK	OK	Adequate

Vessel – Adsorption			
Item Number	**Equipment Name**	**Design Temp / Pressure**	**Revamp Status**
205-V1	Adsorbent Chamber No 1	OK	New internals required
205-V2	Adsorbent Chamber No 2	OK	New internals required

Vessel					
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Residence Time**	**Vapor/ Liquid Separation**	**Revamp Status**
205-V3	Raffinate Sidecut Surge Drum	OK	OK	NA	Adequate

205-V4	Raffinate Column Receiver	OK	OK	OK	Adequate
205-V5	Extract Column Receiver	OK	OK	OK	Adequate
205-V6	Finishing Column Receiver	OK	OK	OK	Adequate
205-V7	Paraxylene Chloride Treater	OK	NA	NA	Adequate
205-V8	Water Injection Drum	OK	OK	NA	Adequate
205-UV1	Sump Tank	OK	OK	NA	Adequate

Exchanger

Item Number	Equipment Name	Design Temp. / Pressure	Duty/ Pressure drop	Revamp Status
205-E1	Raffinate Column Feed-Bottoms Exchanger	OK	OK	Adequate
205-E2	Raffinate Column Reboiler	OK	OK	Adequate
205-E3	Raffinate Column Vent Condenser	OK	OK	Adequate
205-E4	Extract Column Feed-Bottoms Exchanger	OK	OK	Adequate
205-E5	Desorbent Pumpout Cooler	OK	OK	Adequate
205-E6	Extract Column Reboiler	OK	OK	Adequate
205-E7	Extract Column Auxiliary Reboiler	OK	OK	Adequate
205-E8	Paraxylene Trim Cooler	OK	OK	Adequate
205-E9	Finishing Column Feed-Bottoms Exchanger	OK	OK	Adequate
205-E10	Finishing Column Desorbent Reboiler	OK	OK	Adequate
205-E11	Finishing Column Steam Reboiler	OK	OK	Adequate*
205-E12	Desorbent Rerun Column Reboiler	OK	OK	Adequate
205-E13	Sump Tank Pumpout Cooler	OK	OK	Adequate
205-E14	Start-up Heater	OK	OK	Adequate

*Minimum MP (16ATA) steam pressure required – 15.8 kg/cm^2g.

Air Cooled Exchanger

Item Number	Equipment Name	Design Temp. / Pressure	Duty/ Pressure drop	Revamp Status
205-AC1	Raffinate Column Condenser	OK	OK	Adequate
205-AC2	Extract Column Condenser	OK	OK	Adequate
205-AC3	Finishing Column Condenser	OK	OK	Adequate
205-AC4	Paraxylene Cooler	OK	OK	Adequate

Item Number	Equipment Name	Capacity	Head	Revamp Status
205-P1A/B	Line Flush Pumps	OK	OK	Adequate
205-P2A/B/C	Chamber Circulation Pumps	OK	NG	Larger impeller required. Motor Adequate
205-P3A/B	Raffinate Column Bottoms Pumps	OK	OK	Adequate
205-P4A/B	Raffinate Column Reflux Pumps	OK	NG	Larger impeller required. Motor is adequate
205-P5A/B	Extract Column Bottoms Pumps	OK	OK	Adequate
205-P6A/B	Extract Column Overhead Pumps	OK	OK	Adequate
205-P7A/B	Finishing Column Bottoms Pumps	OK	OK	Adequate
205-P8A/B	Finishing Column Overhead Pumps	OK	OK	Adequate
205-P9A/B	Sump Tank Pumps	OK	OK	Adequate
205-P10A/B	Desorbent Transfer Pumps	OK	OK	Adequate
205-P11A/B	Paraxylene Transfer Pumps	OK	OK	Adequate
205-P12A/B	Water Injection Pumps	OK	OK	Adequate
205-P13A/B	Desorbent Rerun Column Bottoms Pump	OK	OK	Adequate
205-P14A/B	Desorbent Makeup Pumps	OK	OK	Adequate

1.4.5. Isomar Process Unit

Fired Heaters

Item Number	Equipment Name	Design Temp. / Pressure	Duty	Revamp Status
207-F1	Charge Heater	OK	OK	Adequate.

Column				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Trays**	**Revamp Status**
207-C1	Deheptanizer	OK	OK	Adequate
207-C2	Stripper	OK	OK	Adequate

Vessel					
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Residence Time**	**Vapor/ Liquid Separation**	**Revamp Status**
207-V1	Water Injection Drum	OK	OK	OK	Adequate
207-V2	Separator	OK	OK	OK	Adequate
207-V3	Clay Treater	OK	OK	OK	Adequate
207-V4	Deheptanizer Receiver	OK	OK	OK	Adequate

Reactor				
Item Number	**Equipment Name**	**Design Temp / Pressure**	**Flow Distribution/ Pressure drop**	**Revamp Status**
207-R1	Reactor	OK	OK	Adequate

Exchanger				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Duty/ Pressure drop**	**Revamp Status**
207-E1	Combined Feed Exchanger	OK	OK	Adequate
207-E2	Deheptanizer Feed-Isomer Feed Exchanger	NG	OK	Exchanger re-rating required
207-E3	Deheptanizer Feed-Parex Feed Exchanger	OK	OK	Adequate
207-E4	Deheptanizer Feed-Bottoms Exchanger	OK	OK	Adequate
207-E5	Deheptanizer Trim Condenser	OK	OK	Adequate
207-E6	Deheptanizer Reboiler	OK	OK	Adequate

207-E7	Stripper Reboiler	OK	OK	Adequate
207-E8	Stripper Bottoms Cooler	OK	OK	Adequate
207-AC1	Product Condenser	OK	OK	Adequate
207-AC2	Stripper Condenser	OK	OK	Adequate

Item Number	**Equipment Name**	**Capacity**	**Head**	**Revamp Status**
207-P1A/B	Charge Pumps	OK	NG	Larger size impeller, New motor required
207-P2A/B	Separator Pumps	OK	OK	Adequate
207-P3A/B	Deheptanizer Overhead Pump	OK	OK	Adequate
207-P4A/B	Deheptanizer Bottoms Pump	OK	OK	Adequate

Compressor				
Item Number	**Equipment Name**	**Capacity**	**Discharge Temperature**	**Revamp Status**
07-K1	Recycle Compressor	OK	OK	Adequate

1.4.6. UOP Xylene Fractionation Unit

Fired Heaters				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Duty**	**Revamp Status**
206-F1 A/B	Xylene Column Reboiler Heater	OK	OK	Adequate

Column				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Trays**	**Revamp Status**
206-C1	Reformate Splitter	OK	NG	Replace Bottom Section trays
206-C2	Xylene Column	OK	NG	New ECMD trays
206-C3	Heavy Aromatics Column	OK	OK	Adequate
206-C4	Benzene Toluene Splitter	New	New	New

Vessel					
Item Number	**Equipment Name**	**Design Temp/ Pressure**	**Residence Time**	**Vapor/ Liquid Separation**	**Revamp Status**
206-V1	Reformate Splitter Receiver	OK	OK	OK	Adequate
206-V2 A/B	Clay Treaters	OK	NA	NA	Adequate
206-V3	Xylene Column Receiver	OK	OK	OK	Adequate
206-V4	Parex Feed Surge Drum	OK	OK	OK	Adequate
206-V5	Heavy Aromatics Column Receiver	OK	OK	OK	Adequate
206-V7	Benzene Toluene Splitter Receiver	New	New	New	New

Exchanger – Air Cooled				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Duty/ Pressure drop**	**Revamp Status**
206-AC1	Reformate Splitter Condenser	OK	NG	Two bays to be added in parallel
206-AC2	Heavy Aromatics Column Condenser	OK	OK	Adequate
206-AC3	Heavy Aromatics Column Bottoms Cooler	OK	OK	Adequate

Exchanger – Tubular				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Duty/ Pressure drop**	**Revamp Status**
206-E1	Reformate Splitter Feed-Bottom Exchanger	OK	OK	Adequate
206-E2	Reformate Splitter Reboiler	OK	NG	Replace with a new exchanger
206-E3 A/B	Reformate Splitter Net Overhead Cooler	NA	NA	Bypass in revamp
206-E5	Clay Treater Feed Heater	OK	OK	Adequate
206-E6	Heavy Aromatics Column Reboiler	OK	OK	Adequate
206-E9	Benzene Toluene Splitter Condenser	New	New	New

206-E10	Benzene Toluene Splitter Reboiler	New	New	New

Pump				
Item Number	**Equipment Name**	**Capacity**	**Head**	**Revamp Status**
206-P1 A/B	Reformate Splitter Overhead Pumps	OK	NG	Larger size impeller required. Motor adequate
206-P2 A/B	Reformate Splitter Bottoms Pumps	OK	OK	Adequate
206-P3 A/B	Parex Feed Pumps	OK	NG	Larger size impeller, new motors required
206-P4 A/B/C	Xylene Column Bottoms Pumps	OK	OK	Adequate
206-P5 A/B	Xylene Column Sample Pumps	OK	OK	Adequate
206-P6 A/B	Xylene Column Overhead Pumps	OK	NG	Larger size impeller, new motors required
206-P7 A/B	Heavy Aromatics Column Overhead Pumps	OK	OK	Adequate
206-P8 A/B	Heavy Aromatics Column Bottoms Pumps	OK	OK	Adequate
206-P9 A/B	Benzene Toluene Splitter Overhead Pumps	New	New	New Pump and Motor required
206-P10 A/B	Benzene Toluene Splitter Bottoms Pumps	New	New	New Pump and Motor required

1.5. Recommendations

UOP can assist IOCL Panipat in developing this revamp by conducting a Schedule A .

The Study results indicate that the objectives of the study as outlined in Section 1.2 can be achieved by implementing the equipment modifications and adjusting the process parameters as elaborated below.

1.5.1. Sulfolane Process Unit

Based on the proposed revamp flow-scheme, modifications required in the Sulfolane unit are as mentioned below:

1. Raffinate Water Wash Column (204-C2) would need replacement with 7 new jet-deck trays.
2. New Raffinate Rundown Cooler (204-E12) is required on net Raffinate stream.

1.5.2. Tatoray Process Unit

Based on the proposed revamp flow-scheme, modifications required in the Tatoray unit are as mentioned below:

1. The requirement of catalyst for the revamp scenario is higher than the currently loaded catalyst volume. The Reactor to be loaded with the additional amount of catalyst during the next available opportunity.
2. The Stripper column bottom section trays are inadequate for the revamp conditions. Existing bottoms section trays need replacement with new ECMD trays.
3. Stripper Trim Condenser (208-E6) is under surfaced for the revamp duty and operating conditions. The exchanger will be replaced by a single shell suited for the revamp duties and flowrates.
4. Charge pumps (208-P1A/B) and Stripper Overhead Pumps (208-P3A/B) are inadequate for revamp operating conditions. New pumps are required.

1.5.3. UOP Benzene Toluene Fractionation Unit

Based on the proposed revamp flow-scheme, modifications required in the B-T Fractionation unit are as mentioned below:

1. The Toluene Column trays are inadequate to handle the revamp flows and need to be replaced with new PFMD trays
2. Existing Benzene Product Cooler is not adequate for the revamp condition. One additional shell in series is required to the existing shell to meet revamp requirement.
3. Existing Toluene Column Reboiler is not adequate for the revamp condition. A new Reboiler is required.
4. The Toluene Column Condenser requires additional bay in parallel to the existing two bays to meet revamp duty requirement.
5. New Toluene Column Bottoms Pumps and Toluene Column Reflux Pumps are required as the existing pumps are inadequate for the revamp requirement.

1.5.4. Parex Process Unit

Based on the proposed revamp flow-scheme, modifications required in the Parex Unit are as mentioned below:

1. UOP recommends replacement of Adsorbent Chamber Internals(ACI) with the New Adsorbent (ADS – 47).
2. The Finishing Column bottom section trays are inadequate for the revamp conditions. Replacing the trays in the Bottoms section with new Valve Trays is required.
3. Replacement of the impeller of the Chamber Circulation Pumps (205-P2A/B/C) is required.
4. Replacement of the impeller of the Raffinate Column Overhead Pumps (205-P4A/B) is required.
5. For the Raffinate Column (205-C1) to handle high non-aromatics in the Parex feed, replacement of the existing trays in the Raffinate Column with ECMD trays is recommended

The recommendations noted above need to be taken in conjunction with the ADS-47 reload plus ACI replacement and ACCS modification.

1.5.5. Isomar Process Unit

Based on the proposed revamp flow-scheme, modifications required in the Isomar Unit are as mentioned below:

1. Replacement of the impeller and Motor of the Charge Pumps (207-P1A/B) is required.
2. Re-rating of the Deheptanizer Feed-Isomer Feed Exchanger (207-E2) for higher design pressure is required.

1.5.6. UOP Xylenes Fractionation Unit

Based on the proposed revamp flow-scheme, modifications required in the Xylenes Fractionation Unit are as mentioned below:

1. The Reformate Splitter (206-C1) bottoms section trays are inadequate for revamp conditions. The existing bottom section trays need to be replaced with new valve trays.
2. The Xylene Column (206-C2) trays are inadequate for revamp conditions. The existing MD trays need to be replaced with new ECMD trays.
3. Two additional identical bays are required in the Reformate Splitter Condenser (206-AC1) to meet the revamp duty requirement.
4. Replacement of existing Reformate Splitter Reboiler (206-E2) with a new exchanger is required.

5. Replacement of the impeller of the Reformate Splitter Overhead Pumps (206-P1A/B) is required.
6. Replacement of the impeller and Motor of the Parex Feed Pumps (206-P3A/B) and Xylene Column Overhead Pumps (206-P6A/B) is required.
7. New Benzene Toluene Splitter (206-C4), along with associated condenser, reboiler, receiver and pumps is required to meet the revamp objectives.

2. Design Basis and Scope

2.1. Background

Indian Oil Corporation Limited ("IOCL") is operating a UOP designed and Licensed Aromatics Complex at its refinery located in Panipat, Haryana, India. The original design objective of the Aromatics Complex was to process 500 KMTA of a heart cut naphtha derived from Bombay High or Bonny Light Crude, while producing 360 KMTA of Paraxylene and Benzene commensurate to the heart cut naphtha feed composition. The Aromatics Complex was commissioned between May to August, 2006 and has been operating since then. IOCL would like UOP to assess the feasibility of processing C8 Aromatics recovered from the Pygas Splitter along with the Naphtha Feed in the Aromatics Complex for increasing the PX production to 460 KMTA based on 8000 hrs annual operating.
This report provides the details of the revamp modifications required in the Aromatics Complex to achieve 460 KMTA Paraxylene production in the complex.

2.2. Objectives and Constraints

The main objective of the project is to increase Px production to 460 KMTA from available current Naphtha feed and additional feeds from the Pygas Splitter. The Study will identify the major modifications required to achieve the increased capacity by targeting minimum possible revamp capital cost.

For the increased capacity under the revamp scenario, heater duty increase might warrant modifications. In such a scenario, "helper" heater addition is preferable since these can be easily hooked up to the existing system with minimum downtime requirement.

Column shell needs to be retained. However, if needed, existing trays may be replaced with high capacity trays.

The replacement of impellers for existing pumps is IOCL's preferred option. Providing an additional pump, or two new pumps/drivers, may also be considered. The availability of plot space will govern the addition of a pump.

Provision of new heat exchangers and air cooled exchangers shall be based on plot space availability as determined during the site visit. Actual implementation of such with respect to foundation capability and crane access will be assessed by IOCL.

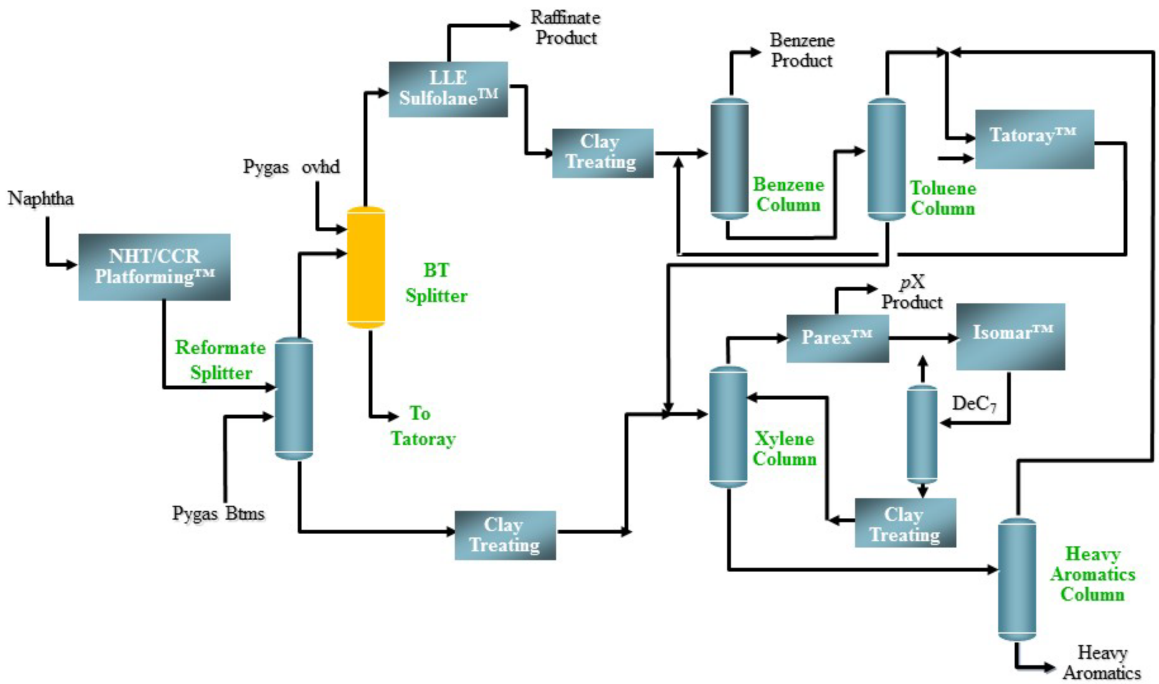
2.3. Basis and Scope

UOP Aromatics Complex YE – AC15274 served as the basis for the study. The yield estimate for the Aromatics unit is based on the addition of a new BT Splitter Column upstream of the Reformate Splitter in the Xylene Fractionation unit.

2.4. Deliverables

1. A report containing a discussion of the results of the study.
2. A summary of the operating conditions, yields and product specifications. The operating conditions are provided in the discussion sections of the individual unit.
3. UOP will supply budgetary equipment costs for new major equipment.
4. These costs will be estimated on a US Gulf Coast basis.
5. UOP will supply scaled utility requirements for the revamp operation.
6. This information is provided in the discussion sections of the individual unit.
7. UOP will supply the PEDS of the new equipment required to meet revamp objective.
8. UOP will provide the flagged PFD.

2.5. Complex flow Scheme



3. Sulfolane Process Unit

3.1. Process Description

Net overheads from the new Benzene Toluene Splitter column in Xylene Fractionation Unit, and Stripper Bottoms from the Isomar unit enter the Raindeck Extractor and flow upward, countercurrent to a stream of lean solvent. As the feed flows through the Raindeck Extractor, aromatics are selectively dissolved in the solvent. A raffinate stream, very low in aromatics content, is withdrawn from the top of the Raindeck Extractor.

The rich solvent, loaded with aromatics, exits the bottom of the Raindeck Extractor, it is heated by exchanging heat with lean solvent from the Water Stripper Reboiler, and then enters the Stripper. The nonaromatic components having volatilities higher than that of benzene are completely separated from the solvent by extractive distillation and removed overhead along with some aromatics and water. This overhead stream is totally condensed. Water vapor, from the Water Stripper, containing small quantities of hydrocarbon joins the Stripper overhead stream between the Stripper Condenser and Stripper Trim Condenser. The water is separated off in the Stripper Receiver boot and sent to the Water Stripper. The overhead distillate is recycled to the Raindeck Extractor where the light nonaromatics displace the heavy nonaromatics from the solvent phase leaving the bottom of the Raindeck Extractor.

The bottoms stream from the Stripper, substantially free of nonaromatic impurities, is sent to the Recovery Column, where the aromatic product is separated from the solvent. The Recovery Column operates under a vacuum, which is maintained by the Recovery Column Ejector system. Because of the large difference in boiling point between the sulfolane solvent and the heaviest aromatic component, this separation is accomplished easily, with minimal energy input. Lean solvent from the bottom of the Recovery Column is used to reboil the Water Stripper and heat the feed to the Stripper before being returned to the Raindeck Extractor. In addition, the lean solvent is routed through the Lean Solvent Filter to remove pipe scale and debris that could potentially plug the holes in the raindeck trays.

The aromatic product, or extract, is recovered overhead and sent on to the Clay Treater Charge Tank at the Benzene-Toluene Fractionation Unit for recovery of the individual benzene and toluene products. Wash water recovered in the Recovery Column Receiver boot is recycled to the Raffinate Water Wash Column.

The raffinate stream exits the top of the Raindeck Extractor and is directed to the Raffinate Water Wash Column. In the Raffinate Water Wash Column, the raffinate is contacted with water to remove dissolved solvent. The solvent-rich water from the Raffinate Water Wash Column is combined with the water from the Stripper Receiver boot and sent to the Water Stripper. Water is vaporized in the Water Stripper Reboiler by exchange with hot circulating solvent. A small amount of the water vapor is used to strip off hydrocarbons in the Water Stripper. The Water Stripper overhead vapor is routed to the Stripper overhead system. Most of the water vapor is used as stripping steam in the Recovery Column. Accumulated solvent from the bottom of the Water Stripper is educted to the Recovery Column.

The raffinate product exits the top of the Raffinate Wash Column. The amount of sulfolane solvent retained in the raffinate is negligible. The raffinate product is sent to storage. Under normal operating conditions, sulfolane solvent undergoes only minor oxidative degradation. A small Solvent Regenerator is included in the design of the unit as a safeguard against the possibility of air leaking into the unit. During normal operation, a small slip-stream of circulating solvent is directed to the solvent regenerator for removal of oxidized solvent.

3.2. Operating Conditions

3.2.1. Operating Parameters

Feed to Sulfolane Unit decreases to 76% of design in revamp. At the same time the raffinate production is expected to increase significantly. As a result of addition of Benzene-Toluene Splitter in Xylenes Fractionation Unit, the Net overhead from Benzene-Toluene Splitter consists of 54.8 mol% benzene, 4.5 mol% toluene and 40.7 mol% non-aromatics. Comparatively, this stream in original design consisted of 25.8 mol% benzene, 53.7 mol% toluene and 20.5 mol% non-aromatics.

Expected benzene recovery in extract is estimated to be 99.9%.

Operating pressures for Sulfolane unit are very similar to the original design. However, due to higher non-aromatic content in feed, the solvent circulation and water circulation rates have changed.

The loads on Raffinate Water Wash column and associated equipment have increased. On the other hand, Raindeck Extractor, Stripper and Recovery column sections have reasonable capacity due to reduced aromatic content and thereby extract rate.

Key operating conditions are listed below:

Operating Ratios	Molar Basis
Solvent Loading, % hydrocarbon	30
Extractor Recycle/Extract	0.64
Stripping Steam/Lean Solvent	0.135
Lean Solvent water content	4.6 mol%

3.2.2. Material Balance

Table below represents the material balance for the revamp case and the original design.

	Revamp
Feed, kg/h	14369
Extract, kg/h	7969
Raffinate, kg/h	6408
Feed, mol% aromatics	59
Feed, mol% non-aromatics	41

3.2.3. Product Specifications

Benzene product specification of 500 wppm non-aromatics is the target for Sulfolane unit. The non-aromatics separation for meeting this specification can be accomplished under revamp operating conditions.

3.3. Major Equipment Review

The design information of the existing equipment has been referenced from the drawings and the information provided by IOCL.

This section will include a brief summary of the results of the equipment evaluation. Other equipment such as the instruments, pressure relief valves and normally not operating equipment were not considered as a part of equipment evaluation

3.3.1. Vessels

Equipment data for vessels and fractionating columns in the form of vendor drawings was made available by IOCL Panipat. The evaluation is preliminary in nature and regardless to the result of the assessment the contractor will be responsible for the following, should the revamp proceed to the detailed design phase:

- Conducting a detailed field inspection based upon consultation with the vendors.
- Ensuring that vessels are refurbished, as required, to provide for safe operation at the revamp conditions.
- Verifying that the structural and mechanical integrity of the vessels is satisfactory for the revamp conditions.
- Ensuring compliance with existing codes and practices, including environmental regulations.
- Ensuring that any replacement parts or modifications comply with the latest appropriate API / ISO Standards. As a minimum, the replacement parts or modifications shall comply with the same standards as the original equipment.
- Verify that the vessel is suitable for the new operating conditions based on actual or modified vessel conditions. If necessary, re-rate the vessel to the new design/vacuum conditions.

4. Tatoray Process Unit

4.1. Process Description

The UOP Tatoray Process unit selectively converts Toluene and C9+ aromatics (A9+) into Benzene and Xylenes. The Toluene Feed to the Tatoray units comes from the Toluene Column Overhead in the Benzene Toluene Fractionation Unit and the new BT Splitter Column bottoms .The A9+ Feed comes from the Heavy Aromatics Column Overhead in the Xylene Fractionation Unit. The combined feed is introduced into the Charge Pumps (208-P1A/B) through Feed Surge Drum (208-V1) and mixed with recycle hydrogen-rich gas at the pump discharge. The combined feed stream is then preheated in the Combined Feed Exchanger (208-E1) by exchange against reactor effluent stream. The combined feed is then heated up to the reaction temperature by the Tatoray Charge Heater (208-F1) before passing to the Reactor (208-R1). This is a single fixed bed Reactor loaded with catalyst TA-30S. After the feed passes through the down-flow reactor, the effluent from the Reactor is routed to the Combined Feed Exchanger. Final cooling of the reactor effluent is achieved in the Product Condenser (208-AC1) followed by a water-cooled Product Trim Condenser (208-E2). The cooled reactor effluent from Product Trim Condenser is then separated into vapor and liquid hydrocarbon in the Separator (208-V2) operating at controlled pressure of 27.0 kg/cm^{2}g.

Vapor from the Separator is compressed and joins the hydrocarbon feed upstream of the Combined Feed Exchanger. The Recycle Compressor (208-K1) recycles the vapor from the Separator to the Combined Feed Exchanger. The Hydrogen-rich makeup gas from the Platforming unit joins downstream of the Recycle Gas Compressor and maintains the required hydrogen purity (80%) in the process. A portion of the recycle gas is purged to maintain recycle gas hydrogen purity, to remove accumulated light ends from the recycle gas loop and the purge stream is routed to Isomar unit along with Platforming unit Hydrogen-rich stream as the Hydrogen makeup in the recycle gas circuit.

Liquid hydrocarbon from the Separator enters the tube side of the Stripper Feed-Overhead Exchanger (208-E5) and shell side of Stripper Feed-Bottoms Exchanger (208-E3) and then enters the Stripper (208-C1). The Stripper Column has 63 trays (18 valve trays above feed and 45 MD trays below feed). Stripper overhead vapor is partially condensed in Stripper Condenser (208-AC2) and followed by water-cooled Stripper Trim Condenser (208-E6). The overhead material is collected in the Stripper Receiver (208-V3) operating at 6.8 kg/cm^{2}g. Stripper Receiver off gases are sent to Recovery plus at Platforming Unit which can handle and remaining sent to Flue Gas Header. Stripper reflux is pumped through Stripper Overhead Pumps (208-P3A/B). The net overhead liquid product is sent to Debutanizer at Platforming Unit. The Stripper column bottoms are reboiled in the Stripper Reboiler (208-E4), which is a HP Steam Heated Exchanger and Charge Heater Convection section. The net Stripper column bottoms is cooled by exchanging heat with the Stripper feed in the Stripper Feed-Bottoms Exchanger and is sent to the Toluene Column at Benzene-Toluene Fractionation Unit.

Toluene Product scenario

A high level assessment was done to check the adequacy of the major equipment's in the Tatoray unit for the production of 30KMTA Toluene. In this scenario the BT Splitter bottoms unit will be routed to a new Sulfolane unit to extract the Non Aromatics from the Toluene. This Toluene will be fed to the Tatoray Unit. The Reactors, Heaters and Compressor appear to be adequate for the Toluene production scenario. The Stripper and the associated Reboiler, condenser and pumps would be new. Modifications will be required in the BT Fractionation unit too and are captured in the report.

4.2. Operating Conditions

4.2.1. Operating Parameters

For the current Study, the catalyst used was TA-30S. The Separator operating pressure was maintained at 27.0 kg/cm2g. The WHSV on fresh feed was maintained at 3 hr-1. The reactor outlet temperature at Start of Run (SOR) and End of Run (EOR) are to be maintained at 480.0oC and 320.4 oC respectively. The Stripper column operating pressure is to be maintained at 6.8 kg/cm2g.

4.2.2. Material Balance

The overall material balance for the Tatoray unit is given in the following table:

	Units	Case
Fresh Feed	kg/hr	78,060
Platforming Unit Makeup Gas	kg/hr	1,230
Purge gas to Isomar Unit	kg/hr	1,517
Stripper Receiver Off Gas to Recovery Plus	kg/hr	1,590
Stripper Receiver Off Gas to Fuel Gas Header	kg/hr	2,575
Stripper Net Overhead to Platforming unit	kg/hr	10,035
Stripper Net Bottoms to Platforming unit	kg/hr	63,412
Off Gas from Separator to Fuel Gas Header	Kg/h	159

4.2.3. Product Specifications

The yield estimates were generated with the objective of providing the Tatoray specification as shown in Section 10.6.

5. UOP Benzene-Toluene Fractionation Unit

5.1. Process Description

The Benzene-Toluene Fractionation Unit consists of two Fractionation columns, the Benzene Column (209-C1) and the Toluene Column (209-C2), and two Clay Treaters (209-V1A/B). The purpose of this unit is to separate purified benzene product from the C7+ aromatics (in the Benzene Column) as well as to separate toluene from the C8+ aromatics (in the Toluene Column). The Toluene is sent to the Tatoray unit and the C8+ material is sent to the Xylene Fractionation Unit. Feeds to the Benzene Toluene Fractionation unit are overhead extract from Recovery Column at Sulfolane Unit, Parex Finishing Column overhead from Parex Unit and the Stripper bottoms from Tatoray unit.
Extract from the Recovery Column is feed to the Clay Treater Charge Tank (209-T1). The feed is then pumped to the Benzene Column by Clay Treater Charge Pumps (209-P1A/B) via Clay Treater Charge Exchanger (209-E1), Clay Treater Charge Heater (209-E2) and finally through the Clay Treaters (209-V1A/B). Clay Treater Charge Exchanger and Clay Treater Charge Heater preheat the cold feed to the desired process outlet temperature of 199 0C before it is fed to the Clay Treaters, where the olefins are removed. The effluent from the Clay Treaters goes through Clay Treater Charge Exchanger before mixing with the overhead stream coming from the Parex Finishing Column. The combined feed stream then enters the Benzene Column. The Benzene Column feed from the Clay Treater Charge Tank contains saturated water, which is removed in the boot of Benzene Column Receiver (209-V2). The Benzene Column produces a side-draw benzene product, which is cooled in Benzene Product Cooler (209-E4) before it is pumped by Benzene Product Pumps (209-P3A/B) to Storage. Benzene Column overhead vapor is condensed via Benzene Column Condenser (209-AC1) into Benzene Column Receiver (209-V2). From there it pumped back to column as reflux via Benzene Column Reflux Pumps (209-P4A/B). The toluene-rich Benzene Column bottoms are fed to the Toluene Column after mixing -with Stripper Bottoms from the Tatoray Unit. The Toluene Column overhead vapor stream is condensed by Toluene Column Condenser (209-AC2) and flows to Toluene Column Receiver (209-V3). The net overhead from Toluene Column Receiver is sent to the Tatoray Unit via Toluene Column Overhead Pumps (209-P6A/B). The Toluene Column bottoms stream is pumped to Xylene Column in the Xylene Fractionation Unit by Toluene Column Bottoms Pumps (209-E5).

<u>Clay Treaters:</u>
The Clay Treaters removes trace quantities of olefinic contaminants from the Benzene-Toluene Fractionation Unit feed. HP steam is used in the Clay Treater Charge Heater to heat up the feed to the desired temperature of 199˚C.

<u>Benzene Column:</u>
The Benzene Column separates 99.9 percent pure benzene product from water, Toluene, Xylenes and Heavier Aromatics. The maximum Toluene limit in Benzene

product is 100 wt ppm. The Benzene product is taken as a side-draw from the column in order to reduce its water content. A Benzene drag (normally no flow) is taken from the overhead receiver in order to remove any light materials that might get accumulated in the Benzene Column Receiver. The Benzene drag stream is returned to the stripper condenser located at the Sulfolane Unit. Water drained from the receiver is sent to CBD/OWS. The Benzene column is reboiled with 16 ATA steam.

Toluene Column:
The Toluene Column separates toluene from the xylenes and the heavier aromatics. The Toluene Column is fed by the Benzene Column bottom and Stripper bottoms from the Tatoray unit. The Toluene Column net overhead is sent to the Tatoray Unit. The net bottoms are sent to the Xylene Fractionation Unit. The Toluene column is reboiled with 16 ATA steam.

5.2. Operating Conditions

5.2.1. Operating Parameters

The Benzene column receiver operates at a pressure of 0.07 kg/cm2(g). For the revamp operation the Toluene Column receiver operating pressure has been increased to 0.7 kg/cm2(g) to reduce vapor load in the column. To reduce the vapor-liquid load in the Toluene column to utilize existing column shell, 1000 ppm Toluene is allowed in the bottom and 1 wt% Xylenes are allowed in the overhead.

	Original Design	**Revamp**
Benzene Column		
Reflux to Feed Ratio	1.11	1.74
Reboiler Duty, MM kcal/hr	1.65	1.64
Receiver Temperature, °C	71	66
Receiver Pressure, kg/cm^2g	0.07	0.07
Top Pressure, kg/cm^2g	0.39	0.39
Bottoms Temperature, °C	138	134
Toluene Column		
Reflux to Feed Ratio	0.96	1.28
Reboiler Duty, MM kcal/hr	4.93	9.1
Receiver Temperature, °C	113	130
Receiver Pressure, kg/cm2g	0.07	0.7
Top Pressure, kg/cm2g	~0.34	1.6
Bottoms Temperature, °C	174	193

5.2.2. Material Balance

The overall material balance for the Benzene Toluene Fractionation Unit is given in the following table:

	Units	
Feed from Sulfolane Unit	kg/hr	7936
Feed from Parex Unit	kg/hr	1480
Feed from Tatoray Stripper Bottoms	kg/hr	65194
Benzene to storage	kg/hr	7213
Toluene to Tatoray Unit	kg/hr	29785
Toluene Column Bottom to Xylene Fractionation Unit	kg/hr	37601
Boot Water to CBD/OWS	Kg/hr	11

5.2.3. Product Specifications

Benzene Product specification is tabulated below:

Parameters	Target	Value Achieved
Purity, wt%	99.9 min	>99.9
Toluene, wt%	0.01 max	0.01
Non Aromatics, wt%	0.05 max	< 0.05
Specific Gravity @ 15.6 ^{0}C	0.882 - 0.886	0.884

6. Parex

6.1. Process Description

Separation of para-xylene from the other C8 aromatics occurs in the Adsorbent Chambers (205-V1/V2). Each Adsorbent Chamber is divided into twelve adsorbent "beds". A specialized grid, which also serves to distribute the flow, supports each bed of adsorbent from below. Each flow distributor is connected to the Rotary Valve ("Coplanar Manifolding Indexer" - 205-ME1) by a "bed line". The flow distributors between each adsorbent bed are used to inject or withdraw liquid from the chamber or to simply redistribute the liquid over the cross-sectional area of the Adsorbent Chamber.
The Adsorbent Chambers have 24 adsorbent beds with 24 grids and 24 bed lines connecting the grids with the Rotary Valve. There are two Adsorption Chambers in series with 12 beds each and pumparound/pusharound pumps.

There are four major streams that are distributed to the Adsorbent Chambers by the Rotary Valve. These "net" streams are:

a. **Feed In -** Mixed xylenes feed from Xylene Column overhead

b. **Dilute Extract Out -** Para-xylene product diluted with desorbent

c. **Dilute Raffinate Out -** EB, meta-, and ortho-xylene diluted with desorbent

d. **Desorbent In –** Recycle desorbent from the fractionation section

At any given time, seven of the bed lines are active, four carrying the net streams into and out of the Adsorbent Chamber and three carrying flushes. The Rotary Valve is used to periodically switch the positions of the liquid feed and withdrawal points as the composition profile moves down the chamber. The Chamber Circulation Pumps (205-P2A/B/C) provides the liquid circulation from the bottom of one Adsorbent Chamber to the top of the other.

The dilute extract from the Rotary Valve is preheated in the Extract Column Feed-Bottoms Exchanger (205-E4) and sent to the Extract Column (205-C2) for separation of the extract from the desorbent. The overhead from the Extract Column is heated in the Finishing Column Feed-Bottoms Exchanger (205-E9) and sent to a Finishing Column (205-C3), where the highly pure para-xylene product is separated from any toluene that entered the unit with the feed.

The dilute raffinate from the Rotary Valve is preheated in the Raffinate Column Feed-Bottoms Exchanger (205-E1) and sent to the Raffinate Column (205-C1) for separation of the raffinate from the desorbent. The sidecut from the raffinate column contains the "unextracted" C8 aromatics: ethylbenzene, meta-xylene, and ortho-xylene. The raffinate product is then sent to the Isomar unit, where additional para-xylene is formed and then recycled back to the Parex unit.

The desorbent from the bottom of both the Extract and Raffinate Columns is recycled back to the Adsorbent Chambers through the Rotary Valve. Any heavy contaminants in the feed will accumulate in the desorbent. In order to prevent this accumulation, provision is made to take a slip-stream of the recycle desorbent to a small Desorbent Rerun Column (205-C4) where heavy contaminants are rejected. During normal operation, mixed xylenes are stripped, clay treated, and rerun prior to being sent to the Parex unit. Thus, there are normally no heavy contaminants to be removed from the bottom of the Desorbent Rerun Column.

6.2. Operating Conditions

6.2.1. Operating Parameters

A single feed case was considered for the Study with the following operating conditions:

Adsorption Section	
Adsorbent Type	ADS-47
Rotary Valve Size	5
A/Fa	0.477
L2'/A	0.635
L3'/A	1.6
L4'/A	-0.450
Cycle Time, minutes	33.6
Temperature, °C	177
Feed Flow Rate, kmta	2503.16
PX Product Flow Rate, kmta	460.29
PX Purity, wt%	99.7
PX Recovery (per pass), wt%	97.0
Raffinate Column	
Internal Reflux to Feed Ratio, molar	0.72
Reboiler Duty, MM kcal/hr	42.12
Receiver Temperature, °C	113

Receiver Pressure, kg/cm^2g	0.08
Extract Column	
Reflux to Feed Ratio, molar	0.64
Total Reboiler Duty, MM kcal/hr	11.8
Receiver Temperature, °C	127
Receiver Pressure, kg/cm^2g	0.07
Finishing Column	
Reflux to Feed Ratio, molar	1.39
Total Reboiler Duty, MM kcal/hr	8.58
Receiver Temperature, °C	66
Receiver Pressure, kg/cm^2g	0.07

6.2.2. Material Balance

The overall material balance for the Parex unit is given in the following table

	Units	Case
Fresh Feed	kg/hr	312895
Raffinate to Isomar	kg/hr	253898
Toluene to B-T Fractionation	kg/hr	1480
Paraxylene Product	kg/hr	57522

6.2.3. Product Specification

The yield estimates were generated with the objective of providing the Para-Xylene specification as shown in Section 10.6.

7. Isomar

7.1. Process Description

The UOP Isomar unit is a fixed bed catalytic isomerization process that converts a mixture of C8 aromatics from the Parex unit Raffinate to a near-equilibrium mixture. Isomar unit is associated with the recovery of para-xylene isomers. Isomar is combined with the UOP Parex process for recovery of para-xylene. The raffinate from the Parex unit is introduced into the suction of Charge Pumps. The feed from charge pumps is combined with hydrogen-rich recycle gas. Make-up gas is added to replace the small amount of hydrogen consumed in the Isomar reactor. Make-up hydrogen is supplied from the Tatoray unit.

The combined feed is then vaporized by exchange with reactor effluent in the Combined Feed Exchanger and then heated to reactor operating temperature in the charge heater. The hot feed vapor is then sent to the Reactor where it is passed radially through a fixed bed of catalyst. This is a single bed Reactor loaded with catalyst I-400.

The reactor effluent is cooled by exchange with the combined feed and condensed in the Product Condenser. The cooled reactor effluent from Product condenser is then separated into vapor and liquid hydrocarbon in the Separator operating at controlled pressure of 8.1 kg/cm2g (SOR) & 15.9 kg/cm2g (EOR) . Hydrogen-rich gas is taken off the top of the Product Separator and recycled back to the Reactor through Recycle Compressor. A small portion of the recycle gas is sometimes purged to remove accumulated light ends from the recycle gas loop.

Liquid from the bottom of the Product Separator is charged to the Deheptanizer. The C8+ fraction from the bottom of the Deheptanizer is clay treated to remove the small amount of olefins generated across the reactor and then it is recycled back to the Xylene Column. The C7- overhead from the Deheptanizer is cooled and separated into gas and liquid products. The Deheptanizer overhead gas is exported to the fuel gas system and the Deheptanizer net overhead liquid is sent to a Stripper column to stabilize the liquid product.

The Stripper bottoms are sent to Sulfolane unit so that any Benzene in this stream can be recovered in the Sulfolane unit. The Stripper overhead vapor is returned to the Deheptanizer overhead vapor line.

7.2. Operating Conditions

7.2.1. Operating Prameters

For the current Study, the catalyst used is I-400. The Separator operating pressure is maintained at 8.1 kg/cm^{2}g (SOR) & 15.9 kg/cm^{2}g (EOR). Separator operating pressure for EOR case has been limited to keep a margin below design pressure. H2/HC ratio for SOR case is to be kept at 3.5 whereas for EOR the ratio has to be kept at 5.5. The variation in H2/HC ratio is required for compressor operation with new feed conditions. The LHSV on fresh feed was maintained at 3.5 hr-1. The reactor inlet temperature at Start of Run (SOR) is to be maintained at 373˚C and 429˚C for End of Run (EOR). The Deheptanizer receiver operating pressure is to be maintained at 3.2 kg/cm^{2}g.

7.2.2. Material Balance

The overall material balance for the Isomar unit is given in the following table:

	Units	**SOR**	**EOR**
Fresh Feed	kg/hr	253893	253893
Taotoray Unit Makeup Gas	kg/hr	1337	1443
Deheptanizer Receiver Off Gas to Fuel gas system	kg/hr	3101	3198
Deheptanizer Net Bottoms to Xylene Fractionation unit	kg/hr	251645	251652
Stripper Bottoms to Sulfolane unit	kg/hr	517	497

8. UOP Xylene Fractionation Unit

8.1. Process Description

The Xylenes Fractionation Unit consists of a Reformate Splitter, Benzene Toluene Splitter, Clay Treaters, Xylene Column and Heavy Aromatics Column along with associated equipment. This unit processes Debutanizer bottoms from the CCR Platforming Process unit, Xylenes rich Pygas from Pygas Splitter bottoms, Toluene rich Pygas from Pygas Splitter overhead, Toluene Column bottoms from Benzene-Toluene Fractionation Unit and Deheptanizer Bottoms from Isomar Unit.

The purpose of the Xylenes Fractionation unit is to fractionate these streams to get Mixed Xylenes feed for Parex Unit for separating para-Xylene product, to prepare feed for the Sulfolane process unit for further extraction of Benzene, and to prepare feed stock consisting of Toluene & C9/C10 aromatics for Tatoray Unit.

The Reformate from the Platforming unit and the Xylenes rich Pygas Splitter bottoms is mixed and heated in the Reformate Splitter Feed-Bottoms Exchanger (206-E1) and is then sent to the Reformate Splitter (206-C1). The aromatics rich Reformate Splitter overhead stream and the toluene rich Pygas Splitter overhead stream are sent to the Benzene Toluene Splitter (206-C4) for separating Benzene and Toluene. The Benzene rich stream from the Benzene Toluene Splitter overheads is sent to Sulfolane Unit for extraction of Benzene from non-aromatics and the Toluene rich stream from the Benzene Toluene Splitter bottoms is sent to Tatoray Unit where toluene and C9+ aromatics (A9+) are converted into benzene and xylenes.

Reformate Splitter bottoms is combined with the Toluene Column bottoms from Benzene Toluene Fractionation Unit and then is preheated to 200°C in Clay Treater Feed Heater (206-E5) before being sent to the Clay Treaters (206-V2A/B). Inlet temperature to the Clay Treaters is controlled by the flow rate of the Xylene Column Bottoms through the Clay Treater Feed Heater. Clay Treaters having high activity clay, help to remove olefinic material which is detrimental to the Parex adsorbent. The clay treated stream is then fed to the Xylene Column (206-C2), along with the Deheptanizer bottoms from Isomar Unit.

The Xylene Column Receiver (206-V3) operates at an elevated pressure of ~ 6 kg/cm^2(g) to allow the reboiling of the Reformate Splitter, Benzene Toluene Splitter, Parex Extract Column and Parex Rafinnate Column using the Xylene Column overhead vapors as the heating medium. The pressure control for the Xylene Column is via the flow rate of the Xylene Column overhead vapors from the Parex Extract Column Reboiler.

Condensed xylenes from Reformate Splitter Reboiler (206-E2), Benzene Toluene Splitter Reboiler (206-E10), Parex Extract Column Reboiler (205-E7) and Parex Raffinate Column Reboiler (205-E2) return to the Xylene Column Receiver to be pumped as reflux to the Xylene Column and as net overhead to the Parex Feed Surge Drum (206-V4).

Mixed xylenes from the Parex Feed surge Drum, having ~ 460 KMTA p-xylene are sent to Parex Unit for recovering p-xylenes from mixed xylenes via the Deheptanizer Feed-Parex Feed Exchanger located in Isomar Unit.

The reboiling duty for this column is provided by the Xylene Column Reboiler Heaters (206-F1A/B). Fuel firing to the heater is controlled by the PDIC at the heater outlet. The Circulating Xylene Column bottoms are used as heating medium in the Clay Treater Feed Heater, Heavy Aromatics Column Reboiler (206-E6), Deheptanizer Reboiler in Isomar Unit, Parex Extract Column Auxiliary Reboiler and Parex Desorbent Rerun Column Reboiler.

Xylenes Column bottoms are sent to Heavy Aromatics Column (206-C3), where C9/C10’s rich overhead stream is sent to Tatoray Unit. The heavy aromatics from the column bottoms at the rate of ~ 20KMTA are sent to storage.

Major Revamp Modifications

- A new Benzene Toluene splitter was added along with the associated equipment in this unit to reduce the revamp loads of the BTS loop (consisting of Sulfolane Unit and Benzene Toluene Fractionation Unit). The Reformate splitter overheads consisting of C7-, instead of direct routing to Sulfolane Unit, are now being routed to the new Benzene Toluene Splitter where Benzene and Toluene are separated. The Benzene Toluene Splitter Receiver will operates at a low pressure of 0.07 $kg/cm^2(g)$.The Benzene fraction from the Benzene Toluene Splitter Overhead is routed to the Sulfolane Unit and the bottoms Toluene fraction is routed to Tatoray Unit. This results in the reduction of the load to the Sulfolane Unit ~ 55%.

- To allow effective utilization of heat in this unit, the Xylene column overhead vapors were used as heating medium in the Benzene Toluene Splitter Reboiler to provide the required reboiling duty in the column.

8.2. Operating Conditions

8.2.1. Operating Conditions

One feed case was considered for this revamp Study. The operating pressure of existing Reformate Splitter Receiver and the Heavy Aromatics Column Receiver were maintained at the original set pressure of 0.07 kg/cm^2(g) and the Xylene Column overhead operating pressure was maintained at 7.4 kg/cm^2(g). The operating pressure of the new Benzene Toluene Splitter Receiver was set at 0.07 kg/cm^2(g). The flow through the Xylene Column Reboiler Heater was maintained same as in the original design.

The operating parameters for all the columns in this unit are tabulated below:

Reformate Splitter	
Reflux to Feed Ratio, mass	0.649
Reboiler Duty, MM kcal/hr	9.57
Receiver Temperature, °C	62
Receiver Pressure, kg/cm^2(g)	0.07
Top Pressure, kg/cm^2(g)	0.35
Bottoms Temperature, °C	172
Benzene Toluene Splitter	
Reflux to Feed Ratio, mass	0.411
Reboiler Duty, MM kcal/hr	3.33
Receiver Temperature, °C	40
Receiver Pressure, kg/cm^2(g)	0.07
Top Pressure, kg/cm^2(g)	0.35
Bottoms Temperature, °C	133
Xylene Column	
Reflux to Feed Ratio, mass	1.79
Reboiler Duty, MM kcal/hr	88.07
Receiver Temperature, °C	226
Receiver Pressure, kg/cm^2(g)	6.0
Top Pressure, kg/cm^2(g)	7.4
Bottoms Temperature, °C	282
Heavy Aromatics Column	
Reflux to Feed Ratio, mass	0.394
Reboiler Duty, MM kcal/hr	1.56
Receiver Temperature, °C	152
Receiver Pressure, kg/cm^2(g)	0.07
Top Pressure, kg/cm^2(g)	0.5
Bottoms Temperature, °C	231

8.2.2. Material Balance

The overall material balance for the Xylenes Fractionation unit is given in the following table:

Stream	Units	Flow Rate
Reformate from Platforming Unit	kg/hr	79,535
Pygas from Pygas Splitter Bottoms	kg/hr	7,120
Toluene Column bottoms from BT Fractionation Unit	kg/hr	37,825
Pygas from Pygas Splitter Overheads	kg/hr	2,875
Deheptanizer Bottoms from Isomar Unit	kg/hr	250,651
Benzene Toluene Splitter overhead to Sulfolane Unit	kg/hr	13,400
Benzene Toluene Splitter bottoms to Tatoray Unit	kg/hr	19,637
Parex Feed to Isomar Unit	kg/hr	312,058
Heavy Aromatics Column bottoms to Storage	kg/hr	2,539
Heavy Aromatics Column overhead to Tatoray Unit	kg/hr	30,369

8.2.3. Product Specification

The feed to the Xylenes Fractionation Unit was set based on 460 KMTA production of p-Xylenes. The specifications targeted in Xylene column are primarily to limit the A9+ components & MEB going along with the parex feed.

The HA column bottoms rate was targeted at 20 KMTA based on IOCL's requirement.

9.3. Over all Utility Summary

NORMAL UTILITIES CONSUMPTION

Unit		Electric Power, KW	HP Steam, kg/hr	MP Steam, kg/hr	LP Steam, kg/hr	BFW, kg/hr	Condensate, kg/hr	Water Loss, kg/hr	Cooling Water, m^3/hr	Fuel, MMKcal/h /hr
NHT	Produced						7,984			3.0
	Consumed	480	6,895			1,089	3,152		4	4.5
CCR Platforming unit	Produced			66,500	700		300			
	Consumed	3,215	59,800			8,700		700	532	17.5
Regenerator Section	Produced						45			
	Consumed	61		45					56	
Sulfolane	Produced									
	Consumed	136		5,500					561	
Tatoray	Produced						16,000			31.0
	Consumed	835	13,800			2,200			247	3.7
B-T Frac	Produced						23,753			
	Consumed	217	535	23,118		100			10	
Isomar	Produced						550			36.1
	Consumed	3,262		550					168	10.0
Parex	Produced						13,090			
	Consumed	1,946		13,090			63		85	
Xylenes Frac	Produced									
	Consumed	2,302							418	101.5
Total	***Produced***	***0***	***0***	***66,500***	***700***	***0***	***61,722***	***0***	***0***	***70.2***
	Consumed	***12,453***	***81,030***	***42,303***	***0***	***12,089***	***3,215***	***700***	***2,081***	***137.2***

Note 1 : Condensate Consumed in NHT and Parex is cold condensate

Indian Oil Corporation Limited
Panipat Aromatics Complex Expansion Project

Process Revamp Study Report
Naphtha Hydrotreating Process Unit

UOP Project Number 9000352

October, 2015

A Honeywell Company

25 East Algonquin Road
Des Plaines, Illinois 60017-5017
U.S.A.

Table of Contents

1 Executive Summary

1.1 Overview

Indian Oil Corporation Limited ("IOCL") is operating a UOP designed and Licensed Aromatics Complex at its refinery located in Panipat, Haryana, India. IOCL would like UOP to assess the feasibility of processing C8 Aromatics recovered from the Pygas Splitter along with the Naphtha Feed in the Aromatics Complex for incremental Paraxylene (PX) production of 460 KMTA @ 8000 hrs annual operating. The Naphtha Hydrotreating (NHT) Unit revamp design considers processing sufficient quantity of naphtha so that the PX production of 460 KMTA can be achieved. To assist IOCL in evaluating this project, UOP has conducted a Process Revamp Study.

The results of this study will provide IOCL with information to help them complete preliminary project economics and establish a firm design basis and direction for further engineering work. The basis and scope of the study are described below.

1.2 Objectives / Background

The main objective of the Aromatics Complex project is to target 460 KMTA PX production from the Naphtha feed and additional feeds from the Pygas Splitter. This Study will identify the major modifications required in the NHT unit to achieve the increased capacity by targeting minimum possible revamp modifications.

The original capacity of the Aromatics Complex was to process 500 KMTA of a naphtha derived from Bombay High or Bonny Light Crude, while producing 360 KMTA of Paraxylene and Benzene commensurate to the naphtha feed composition. The Aromatics Complex was commissioned between May to August, 2006 and has been operating since then.

This report provides the details of the revamp modifications required in the NHT unit to process 1885 MTD or 628 KMTA Naphtha to achieve 460 KMTA Paraxylene production in the complex.

1.3 Basis

The process revamp study for the Aromatics complex is done for a single case considering the HYT-1119™ catalyst in the NHT unit.

A NHT yield estimate (P106907) was generated for the Feed Case considering the Feed compositions of the Panipat and Mathura streams provided to UOP.

The study consisted of the review of all major equipment including Fired Heaters, Reactors, Fractionators, major Vessels, major Heat Exchangers, major Pumps and Compressors. The equipment evaluation was performed based on UOP's simulation of the operation of the process units at the revamp case conditions, and the ability of the existing equipment to meet those requirements based on as-built equipment data sheets and drawings provided by IOCL.

IOCL provided the existing equipment data for study. These included general vessel arrangement drawings and layout drawings of trays, vendor data sheets and construction drawings for heat exchangers, pump API sheets, pump curves and pump motor data sheets, as well as Piping & Instrument Diagrams for the Unit. Revision 0 of the Basic Engineering Design Questionnaire ("BEDQ") information for the dated July 2, 2015, was used for utility system values.

The overall basis, scope and deliverables in this report are in accordance with the Engineering Agreement numbered 15A0075 and dated March 10, 2015 between IOCL and UOP LLC.

1.4 Scorecard and Results

The following equipment list shows the major equipment considered for the operating cases and indicates the status of each item. Please note that minor equipment such as instruments, pressure relief valves, injection pumps and utilities are not considered a part of this Study.

(OK = acceptable, NG = not good, NA = not applicable)

List of major equipment and equipment status

Fired Heaters				
Item Number	**Equipment Name**	**Design Temperature and Pressure**	**Duty**	**Revamp Status**
201-F1	Charge Heater	OK	OK	Adequate, 1. UOP recommends SS 347H as tube metallurgy for new designs. 2. Charge Heater- Convection Section Stripper Reboiler service is adequate for the revamp duty.

Column				
Item Number	**Equipment Name**	**Design Temperature and Pressure**	**Trays**	**Revamp Status**
201-C1	Stripper	OK	NG	Replace trays in Stripper Bottoms section with new valve trays.

Vessels					
Item Number	**Equipment Name**	**Design Temperature and Pressure**	**Residence Time**	**Vapor/ Liquid Separation**	**Revamp Status**
201-V1	Feed Surge Drum	OK	OK	NA	Adequate

201-V2	Water Break Tank	-	-	-	New Pressurized tank is recommended instead of existing open atmospheric tank to avoid hydrocarbon release to atmosphere in case of backflow from the reactor section.
201-V3	Separator	OK	OK	OK	Adequate
201-V4	Recycle Compressor Suction Drum	OK	OK	OK	Adequate
201-V5	Stripper Receiver	OK	OK	OK	Adequate

Reactor				
Item Number	**Equipment Name**	**Design Temperature and Pressure**	**Flow Distribution/ Pressure drop**	**Revamp Status**
201-R1	Reactors	OK	OK	Adequate

Exchangers				
Item Number	**Equipment Name**	**Design Temperature and Pressure**	**Duty/ Pressure drop**	**Revamp Status**
201-E1	Combined Feed Exchanger	OK	NG	Inadequate, Additional surface area required.
201-E2	Stripper Feed-Bottoms Exchanger	NG	OK	Adequate, The exchanger to be re-rated for higher design pressure for the increased shutoff pressure of Stripper Bottoms Pumps.
201-E3	Stripper Reboiler	OK	OK	Adequate
201-E4	Stripper Bottoms Cooler	NG	NG	Inadequate, 1. Can cool upto 35% of slip stream instead of normal 50%. 2. The exchanger to be re-rated for higher design pressure due to the increased shutoff pressure of Stripper Bottoms Pumps.
201-AC1	Product Condenser	OK	OK	Adequate
201-AC2	Stripper Condenser	OK	OK	Adequate

Compressor				
Item Number	**Equipment Name**	**Capacity**	**Discharge Temperature**	**Revamp Status**
201-K1A/B	Recycle Compressors	OK	OK	Adequate

Centrifugal Pumps					
Item Number	**Equipment Name**	**Capacity**	**Head**	**Seal Status**	**Revamp Status**
201-P1 A/B	Charge Pumps	OK	OK	OK	Adequate
201-P2 A/B	Wash Water Injection Pumps	OK	OK	OK	Adequate
201-P4 A/B	Stripper Reflux Pumps	OK	OK	OK	Adequate
201-P8 A/B	Stripper Bottoms Pumps	OK	NG	OK	Inadequate, Install higher size impeller. Motor is adequate.

1.5 Recommendations

This Study will allow IOCL to evaluate the benefits of the proposed revamp and arrive at a design basis for the next stage of development of the project. UOP can assist IOCL in conducting a Revamp Schedule A package to further develop more detailed engineering.

The study results indicate that the objectives of the Study as outlined in Section 1.2 can be achieved by equipment additions and modifications and adjusting the process parameters as elaborated below.

a. The Naphtha feed feeding to the NHT Feed Surge Drum comes from Gas Blanketed Storage Tanks. Out of 3 Storage Tanks, Tank C is floating roof tank as per IOCL FOS Technical Audit report of DEC 2009. The feed from the Floating roof tank is to be routed to Oxygen Stripper in order to avoid oxygenates in the NHT unit which can cause fouling.

b. The metallurgy of the existing tubes in the Charge Heater is A 335 P9. As the amount of H2S in feed is high, UOP recommends SS 347H (A 376 TP 347H) as coil metallurgy to give better corrosion resistance and to avoid frequent catalyst bed skimming in reactor.

c. The Hydrotreating catalyst, HYT-1119™ and two layers of top graded beds (Cat Trap 10, Cat Trap 30) are recommended for Reactor (201-R1).

d. As per current design practice, UOP's recommends a new pressurized Water Break Tank (201-V2) to replace existing atmospheric water break tank with a nitrogen-blanketed pressurized vessel to avoid hydrocarbon release to atmosphere in case of any backflow from the reactor section.

e. For the Stripper Column (201-C1), the trays in the bottom section are inadequate to handle the revamp loads and need to be replaced with new trays with slopped downcomer.

f. The Stripper receiver pressure is reduced from 8.09 Kg/cm^2 (g) to 7.9 Kg/cm^2 (g) in order to have reasonable approach between the Steam and Process side for the Steam Heated Stripper Reboiler service.

g. For Combined Feed Exchanger (201-E1), 10 entirely new shells in series are required in order to keep minimum dew point margin for the Charge Heater feed. This was discussed during the review meeting on 25th August 2015.

h. The Stripper Feed-Bottoms exchanger (201-E2) Shell Side and Tube Side design pressure to be re-rated for the increased shutoff pressure. This is due to recommendation of a higher impeller size for the Stripper Bottoms Pumps.

i. The Stripper Bottoms Cooler to be re-rated for higher design temperature and design pressure. Design temperature is increased considering bypass of Stripper Feed-Bottoms exchanger whereas design pressure is increased due to recommendation of a higher impeller size for the Stripper Bottoms Pumps.

j. For Stripper Bottoms Pumps (201-P8 A/B), the impeller size to be increased from existing 306.5 mm to 336.5 mm for the revamp conditions.

2 Design Basis and Scope

2.1 Background

The main objective of the project is to target 460 KMTA PX production from available current Naphtha feed and additional feeds from the Pygas Splitter. The Study will identify the major modifications required to achieve the increased capacity by targeting minimum possible revamp modifications.

The original design objective of the Aromatics Complex was to process 500 KMTA of a heart cut naphtha derived from Bombay High or Bonny Light Crude, while producing 360 KMTA of Paraxylene and Benzene commensurate to the heart cut naphtha feed composition. The Aromatics Complex was commissioned between May to August, 2006 and has been operating since then.

This report provides the details of the revamp modifications required in the NHT unit to process 1885 MTD or 628 KMTA Naphtha to achieve 460 KMTA Paraxylene production in the complex.

2.2 Objectives and Constraint

The objective of the Study is to identify the major equipment modifications or additions required to allow the NHT Unit to process 1885 MTD of Naphtha feed to provide acceptable feed for the downstream Platforming Unit. The study will identify the major modifications required to achieve the increased capacity by targeting minimum possible revamp modifications.

The following equipment constraints will be assumed:

1. The existing NHT reactor shell should be re-used. Modifications to internals, if required, are acceptable.
2. The existing NHT recycle compressor should be re-used.

2.3 Basis/ Scope of Work

- The Feed to the NHT Unit is a blend of Panipat and Mathura Naphtha in 70/30 weight ratio. The PONA for the Naphtha streams was provided by IOCL. UOP used the average compositions of the Panipat Naphtha streams and the Mathura Stream composition to arrive at the blended Feed used for the study as defined in section 7.1 of the report. The contaminants in the Naphtha were 500 wt ppm Sulfur, 1.5 wt ppm Nitrogen, 0 wt ppm Oxygen and 40 wt ppb Metals as defined by IOCL. UOP yield estimate P106907 served as the basis for the NHT unit study, refer section 6.1 of this report. The NHT unit Separator pressure considered for the revamp is 28.1 kg/cm^2 (g). The NHT Unit design case yield estimate is based on UOP HYT-1119™ catalyst.
- For equipment evaluation, the revamp equipment process requirements were compared with existing equipment vendor datasheets. In the absence of the existing equipment vendor datasheets, UOP Schedule A project specifications were used. The current condition of the equipment is assumed to be identical to that shown on the existing vendor datasheets or as per the UOP project specifications.
- Hydraulic evaluation of pump discharge circuits were performed based on standard piping lengths used by UOP and pipe diameters from the as-built P&ID's. Details hydraulics for the pump discharge circuits will be developed during the next phase of the revamp.
- The scope for the Study includes the major equipment shown on the following, current as-built process flow diagrams:

    1. NHT Unit: Reactor and Stripper section as-built PFD No. 903292-110-01-A1 Rev 3 and 903292-110-02-A1 Rev 2

- Offsites, feed pretreatment, and downstream processing are outside the scope of the Study. Piping, instruments, pressure safety valves, injection pumps and other minor equipment are also outside the scope of the Study.
- The process simulations performed for the Study were used to provide stream flagging information on the Process Flow Diagrams of the NHT Unit. Flagging information provided for the internal streams include mass flow rates, temperatures, controlled pressures and fired heater and heat exchanger duties.

2.4 Deliverables

- A report containing a discussion of the results of the study
- A summary of the operating conditions, yields and product specifications. The operating conditions are provided in the discussion sections of the individual unit.
- UOP will supply budgetary equipment costs for new major equipment.
- These costs will be estimated on a US Gulf Coast basis.
- UOP will supply scaled utility requirements for the revamp operation.
- This information is provided in the discussion sections of the individual unit.
- UOP will supply the PEDS of the new equipment required to meet revamp objective.
- UOP will provide the flagged PFD.

3 Process Description

The UOP Naphtha Hydrotreating unit is a catalytic process that removes organic sulfur, oxygen and nitrogen from the naphtha stream. The Naphtha from gas blanketed storage is introduced into the unit and sent to the fuel gas blanketed Feed Surge Drum which is maintained at an operating pressure of 1.76 kg/cm^2g. The Feed Surge Drum is a horizontal vessel with a boot, to remove free water. Feed from Feed Surge Drum is transferred through the Charge Pumps to the Combined Feed Exchangers and mixed with recycle hydrogen-rich gas and then preheated by exchange against reactor effluent in the Combined Feed Exchanger. The combined feed is then heated up to the hydrotreating reaction temperature by the Charge Heater before passing to the Reactor. There is one single bed Reactor loaded with hydrotreating catalyst HYT-1119™. Top graded bed materials Crystaphase CatTrap 10 and Crystaphase CatTrap 30 are recommended to reduce any potential for pressure drop buildup along with main catalyst HYT-1119™ bed. After the feed passes through the downflow reactor, the effluent from the Reactor is routed to the Combined Feed Exchanger. Final cooling of the reactor effluent is achieved in the air-cooled Product Condenser. Water is injected in the effluent stream before the cooler to prevent deposition of ammonium salts that can corrode and foul the Product Condenser. The cooled reactor effluent from Product Condenser is separated into vapor, liquid hydrocarbon and sour water in the Separator operating at controlled pressure of 28.1 kg/cm^2g.

Wash water from Water Break Tank operating at 0.7 kg/cm^2g is pumped through Wash Water Pumps to mix with cooled reactor effluent before entering to the product condenser.

Vapor from the separator is compressed and joins the hydrocarbon feed upstream of the Combined Feed Exchanger. The Recycle Compressor recycles the vapor from the Separator to the combined feed exchanger. The Hydrogen-rich makeup gas from the Platforming unit joins upstream of the Product Compressor and maintains the required pressure level as hydrogen is consumed in the process. The Separator pressure controller controls the makeup gas rate. Sour water from the Separator is sent to the existing Sour Water Stripping Unit system outside the unit battery limit.

Liquid hydrocarbon from the Separator enters the shell side of the Stripper Feed-Bottoms Exchanger where it is preheated prior to entering the Stripper. The Stripper Column has 25 valve trays to ensure Hydrogen Sulfide, water and lighter hydrocarbons are stripped out in Stripper Column. Stripper overhead vapor is condensed in Stripper Condenser. The overhead material is collected in the Stripper Receiver operating at 7.9 kg/cm^2g. This Stripper Receiver pressure is reduced from original pressure of 8.1 Kg/cm^2 (g) in order to have reasonable approach between the Steam and Process side for the Steam Heated Stripper Reboiler service. Stripper Receiver off gases is sent to Fuel Gas Header. The

Stripper operates on total reflux. Stripper reflux and net overhead liquid (normally no flow) are pumped through Stripper Reflux Pumps. The net overhead liquid product is sent to Refinery outside the unit battery limit. Stripper Receiver boot water is sent to the existing Sour Water Stripping Unit system outside the unit battery limit. The Stripper column bottoms are reboiled with Steam heated thermosyphon reboiler and Convection section of Charge Heater which is a fired heater service. The net Stripper column bottoms is cooled by exchanging heat with the Stripper feed in the Stripper Feed-Bottoms Exchanger before being sent as feed to the CCR Platforming Process unit at optimum feed temperature (114°C). When the CCR Platforming unit is not in operation, the net bottoms stream at the outlet of the Stripper Feed-Bottoms Exchanger is sent to the water cooled Stripper Bottoms Cooler exchanger, before being sent to storage. The existing Stripper Bottoms Cooler exchanger is designed to cool for 50% flow of the existing unit throughput.

4 Operating Conditions

4.1 Cases

The Study considers one feed case with a feed rate of 1885 MTD. The feed to the NHT unit considered for the Study is a blend of 70% Panipat and 30% Mathura feed to the unit.

For the current Study, the catalyst used was HYT-1119^{TM}. The Separator operating pressure was maintained at 28.1 kg/cm^{2}g. The LHSV on fresh feed was maintained at 11.3 hr^{-1}. The reactor outlet temperature at Start of Run (SOR) operation is to be maintained at 330 ^{o}C and 343.3 ^{o}C at End of Run (EOR) operation. The H2/feed ratio is to be maintained at 71.4 Nm^3/m^3. Further, for the revamp, the Stripper Receiver will operate at controlled pressure of 7.9 kg/cm^2 (g).

4.2 Overall Material Balance

The overall material balance for the NHT unit is given in the following table:

	Units	**Revamp Design**
Naphtha Feed from Storage	kg/hr	78710
Makeup Gas from Platforming Unit	kg/hr	144
Cold Clean Condensate	kg/hr	3152
Sour Water from Separator	kg/hr	3113
Sour Water from Stripper Receiver	Kg/hr	34
Off Gas from Stripper Receiver	kg/hr	225
Stripper Net Bottoms to Platforming unit	kg/hr	78634

4.3 Product Specifications

The NHT Unit yield estimates were generated with the objective of providing hydrotreated feed to the CCR Platforming Process unit with Sulfur and Nitrogen contents less than 0.5 wt ppm.

6 Utility Summary

Item Number	Service	Revamp Condition
Fuel Gas Consumed, MMKcal/hr		
201-F1	Charge Heater	4.48
	Total Fuel Gas Consumed, MMKcal/hr	**4.48**
Fuel Gas Produced, MMKcal/hr		
201-V5	Stripper Receiver Offgas	3.01
	Total Fuel Gas Produced, MMKcal/hr	**3.01**
Electricity, kW		
201-AC1	Products Condenser	44.8
201-AC2	Stripper Condenser	
201-P1 A	Charge Pump	194.4
201-P2 A	Wash Water Injection Pump	18.1
201-P4 A	Stripper Reflux Pump	5.4
201-P8 A	Stripper Bottoms Pump	67.5
201-K1 A	Recycle Compressor	149.7
	Total Electricity, kW	**479.9**
Cooling Water, m^3/hr		
201-E4	Stripper Bottoms cooler	[115.4]*
201-P8 A	Stripper Bottoms Pump	1.6
201-K1 A	Recycle Compressor- Aux Lube Oil Pump	2.3
	Total Cooling Water, m^3/hr	**3.9**
HP Steam Consumed, Kg/hr		
201-E3	Stripper Reboiler	6895
	Total HP Steam Consumed, Kg/hr	**6895**
Boiler Feed Water Consumed, Kg/hr		
201-E3	Stripper Reboiler	1089
	Total Boiler Feed Water Consumed, Kg/hr	**1089**
HP Steam Condensate Generation, Kg/hr		
201-E3	Stripper Reboiler	7984

Total HP Steam Condensate Generation, Kg/hr		**7984**
Cold Clean Condensate Consumed, Kg/hr		
201-V2	Water Break Tank	3152
Total Cold Clean Condensate Consumed, Kg/hr		**3152**
Remarks: 1. []* Indicates Intermittent service**.** 2. Intermittent service pumps like Feed Inhibitor Injection Pump, Sulfide Injection Pump, Regeneration Water Return Pump and Stripper Inhibitor Injection Pump were not considered in the Utility estimates.		

Indian Oil Corporation Limited
Panipat Aromatics Complex Expansion Project

Process Revamp Study Report
CCR Platforming Process Unit
&
CCR Regenerator Section
UOP Project Number 9000353

October, 2015

25 East Algonquin Road
Des Plaines, Illinois 60017-5017
U.S.A.

Table of Contents

1. Executive Summary

1.1 Overview

Indian Oil Corporation Limited ("IOCL") is operating a UOP designed and Licensed Aromatics Complex at its refinery located in Panipat, Haryana, India. IOCL would like UOP to assess the feasibility of processing C8 Aromatics recovered from the Pygas Splitter along with the Naphtha Feed in the Aromatics Complex for incremental PX production of 460 KMTA @ 8000 hrs annual operating. The CCR Platforming unit revamp design considers processing sufficient quantity of naphtha so that the PX production of 460 KMTA can be achieved. To assist IOCL in evaluating this project, UOP has conducted a Process Revamp Study to identify the adequacy of major pieces of equipment and recommend necessary changes to meet the objective

The results of this study will provide IOCL with information to help them complete preliminary project economics and establish a firm design basis and direction for further engineering work. UOP is positioned to use the study results as a basis for our work in developing a Revamp Schedule A for the complex at this capacity The basis and scope of the study are described below.

1.2 Objectives / Background

The main objective of the project is to target 460 KMTA PX production from available current naphtha feed and additional feeds from the Pygas Splitter. The Study will identify the major modifications required to achieve the increased capacity by targeting minimum possible revamp capital cost. The CCR Platforming and CCR Regenerator units are part of the "front end" of the Aromatics Complex.

The original design objective of the Aromatics Complex was to process 500 KMTA of a heart cut naphtha derived from Bombay High or Bonny Light Crude, while producing 360 KMTA of Paraxylene. The Aromatics Complex was commissioned between May to August, 2006 and has been operating since then.

This report provides the details of the revamp modifications required in the CCR Platforming unit and CCR Regenerator section to produce sufficient reformate in order to achieve 460 KMTA Paraxylene production in the complex.

1.3 Basis

The process revamp study for the Platforming unit is done for a single case considering R-334 catalyst. Paraxylene production of 460 KMTA and HA Column Bottoms production of 20 KMTA is targeted with 8000 operating hrs/year.

R334 catalyst was recommended for the revamp as it offered better product values by virtue of higher yields and had lesser feed Naphtha requirement. The extent of revamp in the CCR section with R334 is also less as compared to R264. UOP's email communication UOP/IOC/PX-028 dated 24th June has the details on the R334 recommendation.

As IOCL already has R264 catalyst in stock, UOP provided the extent of modifications required in CCR unit if R264 is used in Platforming. The revamp modifications with R264 catalyst are listed in point "o" of section 1.5 of this report. Also the equipment summary of the CCR section lists the revamp status of each equipment with R264 catalyst.

The catalyst recipe and the design cases for the revamp study were fixed with a meeting with IOC on 29th June 15.

A CCR Platforming yield estimate (P065487) was generated for the Feed Case considering the quantity and feed compositions of the Panipat and Mathura streams provided to UOP.
`

The study consisted of the review of all major equipment including Fired Heaters, Reactors, Fractionators, Combined Feed exchanger, major Vessels, major Heat Exchangers, major Pumps and Compressors. The equipment evaluation was performed based on UOP's simulation of the operation of the process units at the revamp case conditions, and the ability of the existing equipment to meet those requirements based on as-built equipment data sheets and drawings provided by IOCL

IOCL provided the existing equipment data for study. These included general vessel arrangement drawings and layout drawings of trays, vendor data sheets and construction drawings for heat exchangers, pump API sheets, pump curves and pump motor data sheets, as well as Piping & Instrument Diagrams for the Unit. Revision 0 of the Basic Engineering Design Questionnaire ("BEDQ") information for the dated July 2, 2015, was used for utility system values.

The overall basis, scope and deliverables in this report are in accordance with the Engineering Agreement numbered 15A0075 dated March 10, 2015 between IOCL and UOP LLC.

1.4 Scorecard and Results

The following equipment list shows the major equipment considered for the operating case and indicates the status of each item. Please note that minor equipment such as instruments, pressure relief valves, injection pumps and utilities are not considered a part of this Study.

(OK = acceptable, NG = not good, NA = not applicable)
List of major equipment and equipment status

CCR Platforming Unit

Fired Heaters				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Duty**	**Revamp Status**
202-F1	Charge Heater	OK	NG	Retubing and adding additional burner is recommended. Addition of Auxiliary heater can also be considered.
202-F2	No 1 Interheater	OK	OK	Adequate.
202-F3	No 2 Interheater	OK	NG	Retubing and adding additional burner is recommended. Addition of Auxiliary heater can also be considered.
202-F4	No 3 Interheater	OK	NG	Retubing and adding additional burner is recommended. Addition of Auxiliary heater can also be considered.

Reactor			
Equipment Name	**Design Temp / Pressure**	**Flow Distribution/ Pressure drop**	**Revamp Status**
Reactor No 1	OK	OK	Adequate
Reactor No 2	OK	OK	Adequate
Reactor No 3	OK	OK	Adequate
Reactor No 4	OK	OK	Adequate

Column				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Trays**	**Revamp Status**
202-C1	Debutanizer	OK	NG	Bottom section re-traying required

Vessel					
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Residence Time**	**Vapor/ Liquid Separation**	**Revamp Status**
202-V1	Separator	OK	OK	OK	Adequate
202-V2	Recontact Drum No 1	OK	OK	OK	Adequate
202-V3	Recontact Drum No 2	OK	OK	OK	Adequate
202-V4	Discharge Drum	OK	OK	OK	Adequate
202-V5A/B	Fuel Gas Chloride Treaters	OK	OK	OK	Adequate
202-V6A/B	Net Gas Chloride Treaters	OK	OK	OK	Adequate
202-V7	Debutanizer Receiver	OK	OK	OK	Adequate
202-V8A/B	LPG Chloride Treaters	OK	OK	OK	Adequate

Exchanger				
Item Number	**Equipment Name**	**Design Temp. / Pressure**	**Duty/ Pressure drop**	**Revamp Status**
202-E1	Combined Feed Exchanger	OK	OK	Adequate
202-E2	Reactor Purge Exchanger	OK	OK	Exchanger duty maintained same as the original design.

202-E3	Recontact Cooler No 1	OK	NG	Add one more identical shell in parallel
202-E4	Recontact Cooler No 2	NA	NA	Not Required
202-E5	Discharge Cooler	OK	NG	Add one more identical shell in parallel
202-E6	Debutanizer Feed-Bottoms Exchanger	OK	OK	Adequate
202-E7	Debutanizer Reboiler Steam Generator	OK	NG	Modification to convection coil required.
202-E8	Debutanizer Condenser	OK	OK	Adequate
202-E15	Net Gas Exchanger			New
202-E16	Economizer			New
202-E17	Net Gas Chiller			New
202-AC1	Products Condenser	OK	NG	Add additional bay in parallel to the existing 2 bays.
202-AC2	Net Gas Cooler	OK	OK	Adequate (Datasheet Not provided. Original duty was used for evaluation)

Pump					
Item Number	**Equipment Name**	**Capacity**	**Head**	**Seal Status**	**Revamp Status**
202-P1A/B	Separator Pumps	NG	OK	OK	Impeller replacement to maximum impeller size required.
202-P2A/B	Recontact Drum No 1 Pumps	OK	OK	OK	Adequate
202-P3A/B	Debutanizer Reboiler pumps	OK	OK	OK	Adequate

202-P4A/B	Debutanizer Overhead pumps	OK	OK	OK	Adequate

Compressor				
Item Number	**Equipment Name**	**Capacity**	**Discharge Temperature**	**Revamp Status**
202-K1	Recycle Compressor	OK	OK	Both turbine and compressor adequate
202-K2A/B	Net Gas Compressors	NG	OK	Parallel compressor required.

CCR Regenerator Section			
Item Number	**Equipment Name**	**Revamp Status (R334)**	**Revamp Status (R264)**
203-H1	Reduction Gas Heater No 1	New	New
203-H2	Reduction Gas Heater No 2	Adequate	Adequate
203-H3	Regeneration Heater	Adequate	Adequate
203-H4	Air Heater	New	New
203-V1	Catalyst Addition Funnel No 1	Adequate	Adequate
203-V2	Catalyst Addition Lock Hopper No 1	Adequate	Adequate
203-V3	Disengaging Hopper	Replace Elutriation Tube	Replace Elutriation Tube
203-V4	Fines Collection Pot	Adequate	Adequate
203-V5	Catalyst Addition Funnel No 2	Adequate	Adequate
203-V6	Catalyst Addition Lock Hopper No 2	Adequate	Adequate
203-V7	Nitrogen Seal Drum	Re-rate for higher operating temperature	Re-rate for higher operating temperature

203-V8	Lock Hopper	New	New
203-V9	Vent Gas Wash Tower	Replace the Packing	Replace the Packing
203-V10	Caustic Break Tank	Adequate	Adequate
203-V11	Water Break Tank	Adequate	Adequate
203-E1	Reduction Gas Exchanger	Adequate	New
203-E2	Booster Gas Heater	Adequate	New
203-E3	Caustic Cooler	New	New
203-AC1	Regeneration Cooler	New	New
203-P1A/B/C	Organic Chloride Injection Pumps	New	New
203-P2A/B	Caustic Circulation Pumps	Adequate	Adequate
203-P3A/B	Caustic Injection Pumps	Adequate	Adequate
203-V4A/B	Water Injection Pumps	Adequate	Adequate
203-K1	Fines Removal Blower	Adequate	New
203-K2	Lift Gas Blower	Adequate	Adequate
203-K3	Regeneration Blower	Replace the existing driver with a variable speed drive	New
203-K4	Regeneration Cooler Blower	Adequate	Adequate
203-R1	Regeneration Tower	Regeneration Tower internal modifications	Regeneration Tower internal modifications

1.5 Recommendations

UOP can assist IOCL Panipat in conducting a Revamp Schedule A package to further develop more detailed engineering.

The Study results indicate that the objectives of the Study as outlined in Section 1.2 can be achieved by equipment additions and modifications and adjusting the process parameters as elaborated below.

a) The existing Reactor Section Fired Heater (Charge Heater, No 2 Interheater and No 3 Interheater) modification is required. The revamp objective can be attained by either revamping existing Fired Heater or adding Auxiliary Fired Heater to attain the duty requirement for the revamp operating condition. The revamp includes retubing or replacing existing heater radiant coil. Number of staged fuel gas burner also needs to be increased for Charge Heater, No 2 Interheater and No 3 Interheater.
b) The Products Condenser requires additional bay of parallel condenser to the existing two bays to meet revamp duty requirement.
c) Separator Pump impeller replacement to maximum size is required.
d) Existing Recontact Cooler No 1 is not adequate for the revamp condition. One additional shell in parallel is required to the existing shell to meet revamp requirement.
e) Existing Discharge Cooler is not adequate for the revamp condition. One additional shell in parallel is required to the existing shell to meet revamp requirement.
f) The Net Gas Compressors are not adequate for revamp requirement. Both the compressors need to be operated in parallel to meet revamp flow. Therefore a third compressor is required with equal capacity to the existing two compressors. In the revamp operating scenario, two compressors would be running and a third compressor would remain as stand-by.
g) Addition of a Net Gas Chiller, Economizer and Net Gas Exchanger has been suggested for the revamp operating condition. This is to maximize the light ends recovery for the revamp condition.
h) Debutanizer bottom trays are inadequate for the revamp condition. Re-traying is required for this section. UOP recommends trays with sloped downcomer for the revamp operation.
i) Presently Debutanizer reboiler convection coil design has three rows of bare tube and on top one row of fin section. UOP suggests removing the top fin section and one row of bare tube from the convection section to achieve required approach temperature for Debutanizer Reboiler Steam Generator. This fin section and one row of bare tube would be utilized for HP steam generation.
j) For CCR Regenerator Section, internal modification is required for Regeneration Tower. This modification includes reducing the inner screen diameter and modifying the baffles in the Chlorination Zone,

Drying Zone and the Cooling Zone.

k) New electric heaters (Reduction Gas Heater No 1 and Air Heater) are required for the revamp conditions of CCR Regenerator Section.
l) New Lock Hopper is required instead of existing Lock Hopper in CCR Regenerator Section. New Lock Hopper will have three vessels with valves in-between for isolation. The total tangent length of new vessels will be same as the tangent length of existing Lock Hopper so as to fit them in the existing structure.
m) New packing is required for the increased flow in revamp to the Vent Gas Wash Tower. New Caustic Cooler is required for the increased revamp duty.
n) Due to increased catalyst circulation rate and coke burning in the Regeneration Tower, new Regeneration Cooler is required to cool the burn zone circulation gas.
o) If R-264 catalyst is used for this revamp, the following new equipment are required in addition to what is discussed in points j) to n) above.
    a. Reduction Gas Exchanger
    b. Booster Gas Heater
    c. Regeneration Blower
    d. Fines Removal Blower
p) If CCR revamp is carried out for R-334 catalyst and R-264 catalyst is in operation, the expected PX production would be approximately 440 KMTA.

2. Design Basis and Scope

2.1 Background

Indian Oil Corporation Limited ("IOCL") is operating a UOP designed and Licensed Aromatics Complex at its refinery located in Panipat, Haryana, India. The original design objective of the Aromatics Complex was to process 500 KMTA of a heart cut naphtha derived from Bombay High or Bonny Light Crude, while producing 360 KMTA of Paraxylene and Benzene commensurate to the heart cut naphtha feed composition. The Aromatics Complex was commissioned between May to August, 2006 and has been operating since then. IOCL would like UOP to assess the feasibility of processing C8 Aromatics recovered from the Pygas Splitter along with the Naphtha Feed in the Aromatics Complex for incremental PX production of 460 KMTA @ 8000 hrs annual operating.

This report provides the details of the revamp modifications required in the CCR Platforming and CCR Regenerator unit to achieve 460 KMTA Paraxylene production in the complex.

2.2 Objectives and Constrains

The main objective of the project is to target 460 KMTA PX production from available current Naphtha feed and additional feeds from the Pygas Splitter. The Study will identify the major modifications required to achieve the increased capacity by targeting minimum possible revamp capital cost.

For the increased capacity under the revamp scenario, heater duty increase might warrant modifications. In such a scenario, "helper" heater addition is preferable since these can be easily hooked up to the existing system with minimum downtime requirement.

Column shell needs to be retained. However, if needed, existing trays may be replaced with high capacity trays.

The replacement of impellers for existing pumps is IOCL's preferred option. Providing an additional pump, or two new pumps/drivers, may also be considered. The availability of plot space will govern the addition of a pump.

Provision of new heat exchangers and air cooled exchangers shall be based on plot space availability as determined during the site visit. Actual implementation of such with respect to foundation capability and crane access will be assessed by IOCL.

2.3 Basis / Scope of Work

UOP Yield Estimate P065487 served as the basis for the Platforming unit study. The yield estimate is based on R-334 catalyst and Separator pressure has been increased to 2.81 kg/cm2g form the existing 2.46 kg/cm2g so that existing Recycle Compressor is adequate for the revamp condition.

2.4 Deliverables

- A report containing a discussion of the results of the study
- A summary of the operating conditions, yields and product specifications. The operating conditions are provided in the discussion sections of the individual unit.
- UOP will supply budgetary equipment costs for new major equipment.
- These costs will be estimated on a US Gulf Coast basis.
- UOP will supply scaled utility requirements for the revamp operation.
- This information is provided in the discussion sections of the individual unit.
- UOP will supply the PEDS of the new equipment required to meet revamp objective.
- UOP will provide the flagged PFD.

3. Process Description

Hydrotreated naphtha comes to the unit from the Stripper bottoms in NHT section and mixes with the hydrogen rich recycle gas stream. The mixture enters the Combined Feed Exchanger (Packinox), where it picks up heat and vaporizes. The combined feed is further heated in the Charge Heater before it enters the Reactor No 1. The Reactor No 1 effluent passes through a series of interheaters and reactors in the following sequence: No 1 Interheater, Reactor No 2, No 2 Interheater, Reactor No 3, No 3 Interheater and Reactor No 4. The reactor effluent is cooled and condensed in the Packinox followed by Products Condenser before entering the Separator. Hydrocarbon liquid from Separator is pumped to recontact section.

Hydrogen rich gas from the Separator enters the suction of the steam turbine driven Recycle Compressor, where it is compressed and one part is combined with the feed as recycle gas. Another part is sent to the Net Gas Cooler in the Recontact section. After Net Gas Cooler, the net gas mixes with Recontact Drum No 2 liquid and Debutanizer Receiver off gas and the combined stream is cooled in the Recontact Cooler No 1 before sending it to the Recontact Drum No 1. A part of the vapor from the Recontact Drum No 1 mixes with the Stripper net overhead vapor from the Tatoray Unit and enters the Recovery Plus section. A part of the Recontact Drum No 1 liquid also enters the Recovery Plus. From the Recovery Plus section, the rich gas goes to the Fuel Gas Chloride Treaters (in series) and thereafter to the fuel gas header. The liquid from Recovery Plus mixes with Recontact Drum No 1 liquid and goes to Debutanizer Feed-Bottoms Exchanger. Remaining net gas from Recontact Drum No 1 goes to the First Stage Net Gas Compressor. From Net Gas Compressor Discharge, this stream mixes with Separator liquid, is cooled by Economizer followed by Net Gas Chiller. From Net Gas Chiller, the cooled and condensed stream enters Recontact Drum No 2. From Recontact Drum No 2, the liquid stream is heated in the Economizer before sending it to Recontact Drum No 1. The net gas from Recontact drum No 2 is heated by Net Gas Exchanger and goes to the suction of Second Stage Net Gas Compressor. Second Stage Net Gas Compressor discharge is cooled in the Discharge Cooler and sent to the Discharge Drum where vapor and liquid is separated. A part of net gas from Discharge Drum goes as Booster gas to the CCR Regenerator section. The remaining gas is sent to the Net Gas Chloride Treaters for removal of chloride. From Net Gas Chloride Treater, make up gas is taken for NHT and Tatoray unit. Remaining net gas is sent to the PSA header. The liquid from Discharge Drum mixes with liquid from Recontact Drum No 1 and sent to Debutanizer Feed-Bottoms exchanger. From Debutanizer Feed-Bottoms Exchanger, the feed enters the Debutanizer column.

The Debutanizer column overhead is cooled and condensed in the Debutanizer Condenser and goes to Debutanizer Receiver. The vapor from Debutanizer Receiver goes back to the recontact section. The liquid from the Debutanizer receiver is pumped to the LPG Chloride Treater for removal of chlorides. The treated LPG is sent for storage. The Bottoms from Debutanizer section splits into reboiler stream and net bottoms stream. The reboiler stream is pumped to the Debutanizer Reboiler Steam generator where 16 ATA steam is generated. From there it goes to the convection section and returns to the column. The net bottoms from the Debutanizer is cooled in the Debutanizer Feed-Bottoms Exchanger and sent to the Reformate Splitter at Xylene Fractionation Unit.

3.1. Operating Conditions and Material Balance

3.1.1 Cases

Only one feed case was considered for the Study. The feed case is based on R-334 catalyst. The Separator pressure has been increased to 2.81 kg/cm2g from existing 2.46 kg/cm2g so that the existing Recycle Gas Compressor is adequate for the revamp operating condition. The H2/HC ratio considered is 2.1. Two stage recontact system has been used to attain battery limit pressure requirement of 35.2 kg/cm2g. The Debutanizer column is operated at 14.5 kg/cm2g operating pressure. The Column specification includes 1 mol% maximum C4- in the bottoms and 1 mol% maximum C5+ at the overhead.

3.1.2 Material Balance

The overall material balance for the Platforming unit is given in the following table:

	Units	Case
Fresh Feed	kg/hr	78,592
Gas from CCR Regenerator Section	kg/hr	372
To Fuel Gas Header	kg/hr	4,846
Tatoray Stripper Net Overhead Vapor to Recovery Plus	kg/hr	1,590
Booster Gas to CCR Regenerator Section	kg/hr	534
Makeup Gas to Naphtha Hydrotreating Unit	kg/hr	146
Net Gas to PSA header	kg/hr	2,369
Makeup Gas to Tatoray Unit	kg/hr	1,230
LPG from Debutanizer overhead	kg/hr	1,960
Reformate to Reformate Splitter	kg/hr	79,666
Tatoray Stripper Net Overhead Liquid	kg/hr	10,021
Lock Hopper Gas	kg/hr	176

3.1.3 Product Specifications

The reformate from the Debutanizer bottoms contain maximum 1 mol% C4-. The Debutanizer overhead contain maximum 1 mol% C5+.

6. Utility Summary

Below is the estimated utility summary for the revamp operating condition:

Item Number	Service	
Fuel, MMKcal/hr		
	Flue Gas to Fuel Gas Header	- 62.0 (Generated)
202-F1	Charge Heater	12.5
202-F2	No 1 Interheater	13.7
202-F3	No 2 Interheater	10.0
202-F4	No 3 Interheater	8.3
	Total Fuel, MMKcal/hr	-17.5 (Generated)
Electricity, kW		
202-P1A/B	Separator Pumps	60
202-P2A/B	Recontact Drum No 1 Pumps	56
202-P3A/B	Debutanizer Reboiler Pumps	36
202-P4A/B	Debutanizer Overhead Pumps	39
202-K2A/B	Net Gas Compressors	2420
202-AC1	Products Condenser	217
202-AC2	Net Gas Cooler	7
	Recovery Plus	390
	Total Electricity, kW	3225
HP Steam, ton/hr		
202-K1	Recycle Compressor	67.4
	Heater Steam Generation HP Steam	-7.5 (Generated)
	Total HP Steam, ton/hr	59.9
MP Steam, ton/hr		
202-K1	Recycle Compressor	-67.4 (Generated)
202-E7	Debutanizer Reboiler Steam Generator	0.9
	Total MP Steam, ton/hr	-66.5 (Generated)
Boiler Feed Water, ton/hr		
	Heater Steam Generation HP Steam	7.7
202-E7	Debutanizer Reboiler Steam Generator	0.6
	Total Boiler Feed Water, ton/hr	8.3
Cooling Water, m^3/hr		
202-P3A/B	Debutanizer Reboiler Pumps	2.3
202-K1	Recycle Compressor	5.2
202-K2A/B	Net Gas Compressors	62.0
202-E3	Recontact Cooler No 1	140
202-E5	Discharge Cooler	121
202-E8	Debutanizer Condenser	144
	Recovery Plus	123
	Total Cooling Water, m3/hr	597

7 CCR Regenerator Section

7.1 Process Description

Process Flow Diagrams (PFD's) for these units are in Appendices 10.3. The new or modified equipments for this unit are highlighted in bold text.

The CCR Regenerator section consists of a system of integrated equipment that is separate from, but still connected to the CCR Platforming unit. The CCR Regenerator section regenerates spent catalyst from the CCR Platforming section. Proper operation of the CCR Regenerator section ensures that the CCR Platforming section operates economically at the required severity with optimum catalyst performance. The CCR Regenerator section consists of two sections- catalyst regeneration section and catalyst transfer section.

7.2 Catalyst Regeneration

Catalyst regeneration consists of the following four steps:

- Coke Burning
- Chlorination/Oxidation
- Drying
- Reduction

The first three steps occur in the Regeneration Tower. The fourth step occurs in the Reduction Zone above the Platforming Reactors. A fifth step, catalyst cooling, occurs in the bottom of the Regeneration Tower and is not part of actual catalyst regeneration, but is required for proper catalyst transfer. Overall catalyst flow is controlled by cycling the Lock Hopper using the Catalyst Regeneration Control System.

Coke is removed from the catalyst in the Burn Zone of the Regeneration tower through the combustion process, a highly exothermic reaction with oxygen being consumed, and water and CO_2 are generated. The catalyst can easily be damaged within this zone because of the high temperature and moisture content. As a result, the burning process must be carefully controlled by controlling the oxygen content in the regeneration gas.

The second step takes place in the Chlorination Zone of the Regeneration tower. The purpose of this step is to restore the proper metal and acid functions of the catalyst. During this step, chloride content in the catalyst is adjusted; metal is oxidized and dispersed evenly throughout the catalyst surface.

The third step is designed to remove excess moisture on the catalyst, one of the by-products of the coke burning process, and is favored by high temperature, adequate drying time and drying gas flow rate. This step occurs in the Drying Zone of the Regeneration tower. The drier the catalyst is prior to entering the Reduction Zone, the more effective is the reduction step.

As mentioned previously, there is an additional step, catalyst cooling, which also takes place in the Cooling Zone of the Regeneration Tower. This step is not part of the regeneration but is desirable for catalyst transfer and to eliminate the metallurgy upgrade requirement for downstream piping.

7.3 Catalyst Circulation

Spent Catalyst from the Catalyst Collector at the bottom of the Reactor Stack is transported with nitrogen lift gas via the Spent Catalyst L-Valve Assembly and lift line to the Disengaging Hopper, located above the Regeneration Tower. Also, from the L-Valve, a small stream of nitrogen flows upward to the Catalyst Collector, counter-current to the downward flow of catalyst. This nitrogen flow isolates the hydrogen/hydrocarbons in the reactor from the nitrogen lift gas. Differential pressure control systems maintain a higher nitrogen pressure at the Spent Catalyst L-Valve than at the bottom of the Catalyst Collector. This produces what is called a "nitrogen bubble".

Gas from the Disengaging Hopper flows to a Dust Collector, where catalyst fines are removed from the gas, then the gas flows to the Lift Gas Blower. The gas from the lift gas blower discharge split to lift gas and elutriation gas. The lift gas flows to the Spent Catalyst L-Valve assembly for lifting the spent catalyst to Disengaging Hopper. The Elutriation gas flows to the Disengaging Hopper for removing fines from the circulating catalyst. The nitrogen system is isolated from the Regeneration Tower by a small stream of nitrogen, which flows down the catalyst pipe from the Disengaging Hopper to the Regeneration Tower.

Catalyst flows by gravity from the Disengaging Hopper to and through the space between concentric screens in the Burn Zone of the Regeneration Tower. Coke burning occurs as the catalyst flows downward through the annular space while hot combustion gas, with controlled oxygen content, passes radially through the catalyst bed from outside to inside. An oxygen-rich net gas stream flowing upward from the Chlorination Zone immediately below supplies oxygen for coke burning. This stream combines with the effluent from the burn zone, mixes within the inner screen, and the total exits the Regeneration Tower at the top, flows through the Regeneration Blower, Regeneration Cooler, and the Regeneration Heater back to the Regeneration Tower.

From the Regeneration Blower, a portion of the circulating gas is routed to the Reheat Zone, a radial flow section within the Burn Zone concentric screens located immediately below the Burn Zone, for the purpose of re-heating the catalyst before it enters the Chlorination Zone below. An Oxygen Analyzer/Controller on the circulating combustion gas to the Regeneration Tower maintains the required oxygen level in the circulating gas stream by adjusting the venting of air from the drying zone. The electric Regeneration Heater does not normally operate and is only used for start-up. The combustion product vent from Burn Zone is controlled out on differential pressure control.

The net vent gas from the Regeneration Tower it treated with 1 weight% Caustic solution in the Vent Gas Wash Tower for the removal of acid gases before sending to the atmosphere.

From the Burn/Reheat Zone, carbon-free catalyst flows by gravity downward into the Chlorination Zone, where the metals on the catalyst are oxidized and dispersed and the catalyst chloride level is adjusted. A net stream of hot combustion air flows upward from the Drying Zone, located directly below, and mixes with vaporized organic chloride, injected through a distributor, and passes upward into the catalyst bed from the bottom. From the Chlorination Zone catalyst flows by gravity into the Drying Zone, where drying occurs by way of counter-current contacting with dry, heated air flowing upward through the bed. Gas from the Drying Zone flows internally upward into the Chlorination Zone as required by the Burn Zone oxygen consumption.

Catalyst leaving the bottom of the Drying Zone is hot and heat is exchanged with the incoming air, which is at ambient temperature. This heat exchange occurs in the Cooling Zone, where hot catalyst from the Drying Zone is cooled by the counter-current flow of air from the Air Drier. The heated gas from the Cooling Zone flows through the electric Air Heater to bring it to the appropriate temperature for drying. Flow through the Air Heater is caused by the restriction of the catalyst bed in the outlet distributor section of the Drying Zone.

Oxidized catalyst from the bottom of the Regeneration Tower passes through the Nitrogen Seal Drum to the Disengaging Zone Lock Hopper. From the Nitrogen Seal Drum a small stream of nitrogen flows upward to the Regenerator. From the Nitrogen Seal Drum a small stream of nitrogen flows downward to the Disengaging Zone of the Lock Hopper. This isolates the oxygen-containing equipment (Regeneration Tower) from the hydrogen-containing equipment (Lock Hopper and Reactors) and creates a "nitrogen bubble" on the regenerated catalyst section. Catalyst is transferred through the Lock Hopper in small batches, then out as a continuously flowing stream to the Regenerated Catalyst L-Valve and to the Reduction Zone at the top of the Platforming Reactors. A controlled amount of heated Booster Gas flows continuously to the Lock Hopper Surge Zone. The Catalyst Regeneration Control System (CRCS) allows catalyst flow batch-wise through the Lock Hopper by varying the Lock Hopper Zone pressure between that of the Regeneration Tower and that of the Regenerated Catalyst L-Valve Assembly. Off-gas from the Lock Hopper flows continuously to the Product Condenser in the Platforming Unit.

9. Utility Summary

Below is the estimated utility summary for the revamp operating condition:

Item Number	Service	
Electricity, kW		
203-P1A/B/C	Organic Chloride Injection Pumps	0.0002
203-P2A/B	Caustic Circulation Pumps	17.4
203-P3A/B	Caustic Injection Pumps	0.0001
203-P4A/B	Water Injection Pumps	1.2
203-H1	Reduction Gas Heater No 1	227.9
203-H2	Reduction Gas Heater No 2	86.1
203-H4	Air Heater	91.3
203-K1	Fines Removal Blower	4.7
203-K2	Lift Gas Blower	3.6
203-K3	Regeneration Blower	46.9
203-K4	Regeneration Cooler Blower	24.3
	Total Electricity, kW	503.4
MP Steam, ton/hr		
203-E2	Booster Gas Heater	0.1
	Total MP Steam, ton/hr	0.1
Cooling Water, m^3/hr		
203-E3	Caustic Cooler	56.4
	Total Cooling Water, m3/hr	56.4

INVISTA
Technologies S.à R.L.

PTA DEBOTTLENECKING FEASIBILITY STUDY

CLIENT: INDIAN OIL CORPORATION LIMITED

INVISTA Proprietary Information

This document has been disclosed to Indian Oil Corporation Limited under the terms of Work Order for the job "Prefeasibility Study for Expansion of PTA Complex at Panipat" 18019370 and the subsequent amendments between Indian Oil Corporation Limited ("IOCL") and INVISTA Technologies S.à r.l. ("INVISTA"). Disclosure of this document to any third party is prohibited, without the prior written permission of INVISTA.

REPORT: DEBOTTLENECKING FEASIBILITY STUDY

CLIENT: INDIAN OIL CORPORATION LIMITED

DATE: December 2015, updated May 2016

CONTENTS:

1.0 Executive Summary

This report describes a feasibility study carried out to assess the potential for debottlenecking the IOCL PTA Plant (ISBL).

Only the main items of equipment have been reviewed. Current plant performance was assessed using data provided by IOCL and gathered by INVISTA during the operations review. The output of this study should be regarded as reflecting the optimum scope of MPI changes required to debottleneck the ISBL plant to achieve a target production rate of 92.1 t/h PTA. However, such a debottleneck is not without technical risk which is discussed in section 11.

At the request of the IOCL Project team INVISTA have considered and provided details, including capital cost estimates for the following two primary options:

A) Air based debottleneck
B) Oxygen Enrichment based debottleneck

Furthermore, each primary option has two sub-options in the Purification Plant separation area:

i) An additional Atmospheric Centrifuge option.
ii) A Rotary Pressure Filter replacing existing centrifuges.

Finally, we have also presented details of the additional marginal cost for:

C) INVISTA proprietary E^2R technology which offers improved paraxylene consumption, reduced energy consumption and reduced effluent generation.
D) Replacement of 21-E1-1607A-D with a shell & tube heat exchanger 21-E1-1607.
E) PAC Suction Chilling Option
F) Replacement cost of the CTA Drier
G) Replacement cost of PAC Intercoolers

Further discussion will be required to identify IOCL's preferred expansion option based on the cases we have been requested to study.

Please note that the INVISTA Proposal for future work related to this project is issued separately to this feasibility study.

2.0 Design Basis

Objectives of this feasibility study are to identify the optimum scope of changes required to debottleneck the IOCL PTA Plant, and to undertake a preliminary assessment of the capital cost and effect on variable cost and effluents.

The scope of the study has been fixed at a targeted production rate of 92.1 t/h PTA.

The operating hours per annum is considered to be 7600 Hrs

At this stage no attempt has been made to assess the effect of the debottlenecking changes on uptime, or plot plan, although INVISTA do not expect any of these effects to have a major impact on project economics.

INVISTA would expect the uprated plant to be capable of manufacturing PTA within the current specification.

INVISTA has not made a detailed assessment of the length of overhaul required to install the debottlenecking project.

The INVISTA scope for this feasibility study covers the ISBL portion of plant only. No assessment has been made of the impact of the debottlenecking project on the OSBL facilities.

During the Revamp Kick-Off Meeting, IOCL asked INVISTA to consider means for reducing acetic acid consumption as part of the scope of the revamp. The 'best in class' performance of the generation of PTA technology operated by IOCL is [DELETED]. This represents a significant opportunity to improve performance versus current performance without capital expenditure. This issue should be discussed in more detail during the next Customer Service Visit (CSV).

Causes of higher acetic acid usage during normal operation can include:

- Operation at reduced 4-CBA in Oxidation Plant
- Non-optimal catalyst composition
- Insufficient scrubbing on HP Absorber
- Loss of acetic acid via residues

A more detailed discussion with IOCL during the CSV would identify the full list of areas.

To improve beyond [DELETED] requires the Oxidation Reactors' temperature to be dropped significantly [DELETED]. The trade-off in variable cost is that a reduction in temperature inevitably reduces the energy efficiency of the plant (increases power consumption per tonne of PTA). INVISTA estimates that the overall benefit in variable cost may at best only be $1-2/t PTA using the equipment configuration operated at IOCL, but with a very severe capital cost implication.

Changing to reduced temperature operation of the Oxidation Reactors requires at least the following significant equipment changes:

- An increase in reactor slurry residence time, achieved by either de-rating the capacity of the plant, or installation of additional Oxidation Reactor.
- An increase in Oxidation Reactor overheads heat exchange area due to reduced vapour temperature exit the Oxidation Reactor, or de-rate of capacity of the plant.
- Significant modifications to the PAC. For example, potential removal of the final stage of compression (for efficiency reasons), new expander, possibly a new steam turbine due to different split in levels of steam raised.
- New HP Absorber due to lower operating pressure.

It is INVISTA's opinion that the return on capital for such a project would be extremely poor, and in the context of a debottleneck would significantly increase the capital expenditure and complexity of the project.

3.0 Process Description

3.1 Oxidation Plant

Unless specifically mentioned in the following sections, main plant items within the Oxidation Plant are not believed to require modification in order to achieve the debottleneck rate.

3.1.1 Area100

3.1.1.1 Air Based Uprate

Process Air Compressor

The Process Air Compressor (PAC) purchased under project no. 10060 should have sufficient capacity for the debottleneck if performance can be restored to the design values. INVISTA believe the most likely cause of underperformance is that the intercoolers are badly fouled on the air side.

IOCL make the observation that the maximum ambient temperature can be, and machine design is for and RH but this is not a realistic combination of ambient conditions and is the worst case temperature paired with the worst case humidity, and in reality never occurs. A more realistic scenario would be a hot day in May / June (for example [DELETED] ambient, with dewpoint of [DELETED]) or a humid day in August (for example [DELETED]ambient with dewpoint of [DELETED]).

The other possibility is that the compressor is very dirty and the contour rings have corroded and the clearances increased. INVISTA has previously recommended that IOCL re-calibrate the field instruments on the PAC and fit differential pressure measurement across the three intercoolers in order to better diagnose this issue. In the absence of feedback from IOCL, for the purposes of this study it has been assumed that performance can be restored to the design intent without additional capital expenditure.

Therefore, INVISTA believe based on a realistic range of coincident ambient conditions that provided the PAC is restored to it's original as built condition and operated at it most efficient point for the conditions, that the PAC has sufficient capacity for the debottleneck capacity for all but a very few days of a typical year. INVISTA consider this period to be no more than 10-20 days per year.

However, based on IOCL's request, INVISTA has also shown an option to include a Suction Chiller package, 21-X1-112. It is expected that the filter within this package will replace the existing inlet air filter 21-G1-112. In INVISTA's opinion addition of the Suction Chiller package alone will not be sufficient to enable PAC operation at the debottleneck capacity without further maintenance work on the PAC (e.g. : Intercooler bundle cleaning/replacement), and as previously stated capital expenditure on the Suction Chiller should not be necessary if the PAC is restored to its original design condition.

In INVISTA's opinion, the combination of maximum ambient temperature with maximum RH is not a realistic combination, either for determining ultimate PAC performance, or for the Suction Chiller design basis., therefore two scenarios have been considered: firstly a hot day in May / June ([DELETED] ambient, with dewpoint of [DELETED]) and a humid day in August (°C ambient with dewpoint of [DELETED]).

Offgas Expander

The existing expander is believed to be able to accept the additional offgas flow without modification.

Steam Turbine

The existing steam turbine is operating at its limit. The majority of the additional steam generated is expected to be generated at the ELP level. Additional ELP steam generated at higher rates plus additional site LP steam is therefore proposed to be supplied to a new stand-alone steam turbine coupled to a generator. In the event of there being any additional LP steam that cannot be consumed by the existing turbine, this would be letdown to ELP steam prior to being fed to the new turbine.

Offgas Scrubber

Packing in the Offgas Scrubber 21-C1-172 is replaced with higher efficiency packing in order to handle the increased offgas rate within the existing Offgas Scrubber diameter. It is proposed to modify offgas desuperheating arrangement to INVISTA's latest design which is 21-C1-172.

3.1.1.2 Oxygen Based Uprate

Oxygen Supply

The debottleneck could alternatively be achieved by enrichment of the existing air supply with oxygen. Gaseous oxygen would be supplied through a new pipeline to ISBL.

Air will be enriched to a level of [DELETED] v/v oxygen. Due to the presence of enriched air, a number of the 'soft' materials such as gland packing, gaskets and seals in a number of valves and instruments (in the process air lines) will need to be upgraded.

Oxygen Injector 21-MX1-181 introduces oxygen into the process air-line exit the PAC. It comprises a monel oxygen diffuser and a short stainless steel process air pipe section. The monel and stainless steel provide necessary ignition resistance and cleanliness.

Offgas Scrubber

No changes are required to the packing in 21-C1-172. It is proposed to modify offgas desuperheating arrangement to INVISTA's latest design which is 21-C1-172.

3.1.2 Area 300

Oxidation Reaction

For both the air and oxygen debottleneck options it is proposed to install equipment to remove some of the incremental heat of reaction, by cooling of the reflux to the Oxidation Reactors from 21-E1-304A/B. This option minimises the increase in pressure (and hence temperature) of the Oxidation Reactors required in order to remain within entrainment factor limitations. Reflux cooling will allow operation of the Oxidation Reactors at [DELETED] post debottleneck.

Reflux Cooling

INVISTA has experience of reflux cooling on another license plant, where steam is raised by cooling of 21-E1-304A/B reflux. Because of the elevation of 21-E304A/B, a pumped system is required. Each Oxidation Reactor system therefore includes a suction vessel (21-V1-312A/B), pumps (21-P1-315A/B) and steam raisers (21-E1-308A/B).

HP Absorber

Based on preliminary calculations by INVISTA, the existing trayed section of this column does not have sufficient capacity for the rates required for the debottleneck. For the purposes of cost estimating, it has been assumed that the trayed section will need to be replaced with trays with higher capacity for example SUPERFRAC® trays from Koch-Glitsch. The feasibility of such a change to achieve the required rates would need to be discussed with the vendors in due course.

It has also been assumed in the capital cost estimate that the demister will be replaced with a higher capacity demister, but this assumption should be re-confirmed during the design phase.

3.1.3 Area 400

Rotary Vacuum Filters

An additional parallel line (line C) of feed pumps, Rotary Vacuum Filter and associated vacuum system is installed in parallel with the existing A/B filters. An additional Vacuum Pump Knockout Drum (21-V1-417C) is installed to service the new line on the assumption that layout will not allow use of the existing vessel (21-V1-417).

CTA Drier

CTA Drier 21-DR1-423 is very similar size to driers on other license plants that are achieving the rate required for the debottleneck. It is assumed that performance of 21-DR-423 can be restored without further capital expenditure. It should be noted that adding an additional Rotary Vacuum Filter may reduce the moisture in CTA fed to 21-DR1-423 with associated benefit in 21-423 loading.

3.1.4 Area 600

Non-E²R Option

The existing Recovery Column 21-V1-631 will need to be replaced with a larger diameter column. The existing Recovery Column Condenser 21-E1-631 is retained. A new Vent Condenser 21-E1-639 is installed to handle the additional condensing heat load arising from the debottleneck.

INVISTA believe that the existing Solvent Dehydration Column 21-C1-601 is adequate for the debottleneck rate of 92.1 t/h PTA based on performance of this diameter of column and type of packing at other plants, Nevertheless INVISTA have no objection to IOCL's proposal to replace the bottom bed of packing with higher efficiency M352Y.

E²R Option

[DELETED]

3.1.5 Area 900

Rotary Valves

The existing valves, 21-B1-902A/B may have sufficient capacity. This could be tested in a plant trial. For the purposes of this report, it is assumed that larger valves are required. The existing conveying lines are retained, the conveying gas rate is increased slightly and the pressure in the conveying line allowed to rise.

3.2 Purification Plant

Unless specifically mentioned in the following sections, main plant items within the Purification Plant are not believed to require modification in order to achieve the debottleneck rate.

3.2.1 Area 1200

.

LP Dissolver Feed Pump 21-P1-1207A/B is uprated by the replacement of existing impellors with larger impellors.

An additional HP Dissolver Feed Pump (21-P1-1209E) of the same size as current pumps is added in parallel with the current four pumps. 21-P1-1209 pumps will be run on a four working, one spare basis.

The preheat train is modified to include a third hot oil preheater (21-E1-1211/C). Hot oil will be used in series on 21-E1-1211A/B/C. The existing hot oil system is believed to have sufficient capacity to meet the additional heat duty arising from the debottleneck.

3.2.2 Area 1300

A modification to the downcomer in the Hydrogenation Reactor is specified in order to enhance mass transfer capability. The Purification Plant catalyst will need to be renewed on a more frequent basis than currently.

3.2.3 Area 1400

PTA Crystallisers

With the exception of First PTA Crystalliser 21-CR1-1401 the existing crystallisers are large enough for the debottleneck. 21-CR1-1401 needs to be replaced with a larger diameter vessel in order to achieve the debottleneck rate of 92.1 t/h PTA. A new agitator (21-A1-1401) is also required.

Separation (Centrifuge Option)

Additional atmospheric separation capacity is required in order to maintain the variable cost of the plant and ensure the PTA Drier has sufficient capacity. A third atmospheric centrifuge (21-CF1-1421C) of the current size is therefore installed on the plant. Atmospheric Centrifuge Feed Pump 21-P1-1420A/B impellors are replaced in order to handle the increased slurry flowrate. A new Atmospheric Centrifuge Discharge Screw (21-B1-1422C) is required to feed the PTA Drier from the new centrifuge.

PTA Mother Liquor Drum 21-V1-1415 is replaced with a larger vessel to handle the higher mother liquor flow and in conjunction with PTA Mother Liquor Flash Drum 21-V1-1601 provide buffering of mother flow during cycle periods when PTA Mother Liquor Filter 21-G1-1603 is not filtering. A new PTA Mother Liquor Drum Scrubber (21-V1-1416) is installed to scrub flash steam prior to use on 21-C1-631.

Separation (Filter Option)

As an alternative to the centrifuge option, the existing pressure and atmospheric centrifuges could be replaced with single stage separation. The simplest implementation (and hence likely lowest capital cost) is achieved with a hydraulic filter. A single hydraulic filter would have sufficient capacity for the debottleneck rate.

PTA slurry from 21-CR1-1405 is delivered to Rotary Pressure Filter 21-RF1-1440 from a circulating loop fed by new RPF Feed Pump 21-G2-1410A/B. Hot Process water from existing Second Reslurry Water Heater 21-E2-1418B is fed to 21-RF1-1440 where it is used to wash the wet cake which forms in the pockets of the filter drum.

Mother liquor and some wash liquor from these stages are collected in new Mother Liquor Filtrate Receiver 21-V1-1432. Cake is then de-liquored using inert gas. Gas and liquor are collected in Drying Gas Filtrate Receiver 21-V1-1433 before passing to Vent Scrubber 21-V1-1615.

PTA cake is discharged into new RPF Cake Discharge Screw 21-B1-1439 and then PTA Drier Cross Screw 21-B1-1441, the pressure having been reduced to approximately atmospheric pressure as it passes through the sections of the filter.

A cloth wash is applied to 21-RF1-1440 following cake discharge. Cloth wash water is collected in new RPF Drainings Receiver 21-V1-1434 before being pumped back to the first wash stage by new RPF Drainings Recycle Pump 21-P1-1435A/B.

Mother liquor from 21-V1-1432 is flashed into new PTA Mother Liquor Flash Drum 21-V1-1437. Process steam generated in this vessel is scrubbed in PTA Flash Drum Scrubber 21-V1-1436 and fed to 21-C1-631. Remaining mother liquor is fed to 21-V1-1601.

PTA Drier

PTA Drier 21-DR1-1423 is very similar size to driers on other license plants that are achieving the rate required for the debottleneck. It is assumed that performance of 21-DR-1423 can be restored without further capital outlay. It should be noted that the Filter Option will significantly debottleneck 21-DR1-1423 due to the reduction in moisture in PTA fed to 21-DR1-1423. Adding

a third Atmospheric Centrifuge would also be expected to help somewhat in reducing PTA feed moisture.

3.2.4 Area 1600

PTA Mother Liquor Filter Feed Pumps

Impellors in the existing pumps 21-P1-1602A/B are replaced with larger versions. Larger motors are required to meet the new duty.

Mother Liquor Filter

Based on preliminary calculations it is believed that the increased mother liquor flow can be handled by the existing PTA Mother Liquor Filter 21-G1-1603. Note any modifications to increase the filter area will be expensive and therefore difficult to justify.

Mother Liquor Cooler

The existing duty is met by four units operated in parallel. A fifth unit (21-E1-1607E) is added for the debottleneck.

Alternatively, the complete duty of 21-E1-1607A-E could be met be replacement with a single shell and tube heat exchanger.

Vent Scrubber

No changes are proposed. A small plume is likely from 21-V1-1615 vent stack if the additional flash steam generated cannot be utilised in the feed preheat section.

3.2.5 Area 1900

Rotary Valve

The existing valve 21-B1-1902 may have sufficient capacity. This could be tested in a plant trial. For the purposes of this report, it is assumed that a larger valve is required. The existing conveying line is retained, the conveying gas rate is increased slightly and the pressure in the conveying line allowed to rise.

3.2.6 Paratoluic Acid Purging

IOCL has requested that the ‘p-Tol Management’ scheme as presented during a previous Customer Service Visit is incorporated into the debottleneck proposal. This has not been done explicitly, however INVISTA note that many of the features of this scheme are incorporated into sub-option (ii) (replacement Purification Rotary Pressure Filters), which INVISTA believes is the most sensible option for achieving the debottleneck Purification rate.

3.3 Additional Onplots

The capacity of Additional Onplot systems is believed to be sufficient to support the debottleneck rate. Some modification is required to the steam system in order to accommodate new Reflux Coolers 21-E1-308A/B and new 2nd Steam Turbine 21-KT1-150.

21-KT1-150 will be supplied with surplus steam, as well as having the facility to accept up to 50 t/h of Site LP Steam.

3.4 Large/Significant (i.e. expensive) Control Valves

Review of significant control valves has not formed part of this study. However, based on experience on other plants it is likely that a number of control valves such as level control valves, pressure control valves and flow control valves will need to be either modified or replace. Detailed checks would be carried out as part of the design phase.

4.0 Process Flow Diagrams

A full set of process diagrams provided in the PEP with project number 10060 has been updated

5.0 Preliminary Equipment List

The preliminary equipment list details new and modified equipment required for the debottleneck. Equipment not specified on the equipment list is therefore believed to be suitable for the debottleneck.

This drawing is a private and proprietary communication and the property of INVISTA. It must not be copied or loaned without the consent of INVISTA and must be returned.

INVISTA Performance Technologies The Wilton Centre, Wilton Redcar TS10 4RF	DOC	**SECTION** **20610/00/15/00100 A**	REV **Z2**	NOTE THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	INPUT BY CHJ CHECKED BY RMS APPROVED BY GMcC	NAME / DATE 10/12/2015
		PLANT	LOCATION	PLANT SECTION		
PRELIM. EQUIPMENT LIST		**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**OXIDATION - Process Air Compression & Offgas Treatment (OPTION A)**		

EQUIP		SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES** **(DegC)**	PRESS **DES** **(kg/cm^2g)**		DIMENSIONS (mm)	**REMARKS**
21 KT1	**150**		2nd STEAM TURBINE	1							AIR BASED UPRATE - Package includes condenser, condensate pumps and generator.
21 C1	**172**		OFFGAS SCRUBBER	1							AIR BASED UPRATE -Replace packing with high efficiency packing. Also, relocate desuperheating within C1-172 by installation of new sprays.

INVISTA Performance Technologies The Wilton Centre, Wilton Redcar TS10 4RF	DOC	**SECTION** **20610/00/15/00100 A(SC)**	REV **Z2**	NOTE THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	NAME INPUT BY CHJ CHECKED BY RMS APPROVED BY GMcC	DATE 10/12/2015

	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**OXIDATION - Process Air Compression & Offgas Treatment (OPTION A(SC))**

EQUIP		SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES (DegC)**	PRESS **DES** **(kg/cm^2g)**	DIMENSIONS (mm)	**REMARKS**	
21 X1	**112**		SUCTION CHILLER PACKAGE	1						AIR BASED UPRATE - Package includes replacement inlet air filter, and all vendor equipment required for suction chilling.	P4
21 KT1	**150**		2nd STEAM TURBINE	1						AIR BASED UPRATE - Package includes condenser, condensate pumps and generator.	
21 C1	**172**		OFFGAS SCRUBBER	1						AIR BASED UPRATE -Replace packing with high efficiency packing. Also, relocate desuperheating within C1-172 by installation of new sprays.	

INVISTA Performance Technologies
The Wilton Centre, Wilton
Redcar
TS10 4RF

DOC	SECTION	REV	NOTE		NAME	DATE
PRELIM. EQUIPMENT LIST	**20610/00/15/00100 B**	**Z2**	THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	INPUT BY	CHJ	05/08/2015
				CHECKED BY	RMS	24/09/2015
				APPROVED BY	GMcC	28/09/2015

PLANT	LOCATION	PLANT SECTION
PTA PANIPAT UPRATE 92.1te/h PTA	**INDIA**	**OXIDATION - Process Air Compression & Offgas Treatment (OPTION B)**

EQUIP		SUB SET	EQUIPMENT TITLE	NO OFF	MOC	CAPACITY EACH	TEMP DES (DegC)	PRESS DES (kg/cm^2g)	DIMENSIONS (mm)	REMARKS
21 KT1	**150**		2nd STEAM TURBINE	1						OXYGEN BASED UPRATE - Package includes condenser, condensate pumps and generator.
21 C1	**172**		OFFGAS SCRUBBER	1						OXYGEN BASED UPRATE -Relocate desuperheating within C1-172 by installation of new sprays.
21 MX1	**181**	PROCESS OXYGEN	OXYGEN INJECTOR	1						OXYGEN BASED UPRATE

This drawing is a private and proprietary communication and the property of INVISTA. It must not be copied or loaned without the consent of INVISTA and must be returned.

INVISTA Performance Technologies The Wilton Centre, Wilton Redcar TS10 4RF	DOC	**SECTION** **20610/00/15/00300**	REV **Z2**	NOTE THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	NAME / DATE INPUT BY CHJ 05/08/2015 CHECKED BY RMS 24/09/2015 APPROVED BY GMcC 28/09/2015

	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**OXIDATION - Reaction & Reactor Overheads**

EQUIP		SUB SET	EQUIPMENT TITLE	NO OFF	MOC	CAPACITY EACH	TEMP DES (DegC)	PRESS DES (kg/cm^2g)	DIMENSIONS (mm)	REMARKS	
21 E1	308A/B	SHELL	REFLUX COOLER	2							P2
		TUBE									
21 C1	310		HP ABSORBER	1						NO CHANGE TO COLUMN - ONLY REPLACEMENT OF 10 TRAYS IN UPPER SECTION WITH HIGH CAPACITY TRAYS eg. SUPERFRAC TRAYS	P3
21 V1	312A/B		REFLUX DRUM	2							
21 P1	315A/B		REFLUX PUMP	2							P2

INVISTA Performance Technologies
The Wilton Centre, Wilton
Redcar
TS10 4RF

DOC	SECTION	REV	Z2	NOTE		NAME	DATE
	20610/00/15/00400			THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	INPUT BY	CHJ	05/08/2015
					CHECKED BY	RMS	24/09/2015
					APPROVED BY	GMcC	28/09/2015

	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**OXIDATION - Crystallisation & Separation**

EQUIP		SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES (DegC)**	PRESS **DES (kg/cm^{2}g)**	DIMENSIONS (mm)	**REMARKS**	
21 RF	**410C**		CTA ROTARY VACUUM FILTER	1 (See remarks)						Additional equipment to be identical to two already installed.	P3
21 V1	**411C**		CTA MOTHER LIQUOR SEPARATOR	1 (See remarks)						Additional equipment to be identical to two already installed.	
21 P1	**412C**		CTA FILTRATE PUMP	1 (See remarks)						Additional equipment to be identical to two already installed.	
21 E1	**415C**	SHELL	CTA VACUUM VAPOUR CONDENSER	1 (See remarks)						Additional equipment to be identical to two already installed.	
		TUBE									
21 P1	**416C**		CTA VACUUM PUMP	1 (See remarks)						Additional equipment to be identical to two already installed.	
21 V1	**417C**		CTA VACUUM PUMP KNOCKOUT DRUM	1 (See remarks)						Additional KO Drum for 3rd ROVAC.	
21 P1	**418C**		CTA SEAL FLUID PUMP	1 (See remarks)						Additional equipment to be identical to two already installed.	
21 E1	**419C**	COLD	SEAL FLUID COOLER	1 (See remarks)						Additional equipment to be identical to two already installed.	
		HOT									

INVISTA Performance Technologies The Wilton Centre, Wilton Redcar TS10 4RF	DOC	**SECTION** **20610/00/15/00400**	REV **Z2**	NOTE THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	NAME / DATE INPUT BY CHJ 05/08/2015 CHECKED BY RMS 24/09/2015 APPROVED BY GMcC 28/09/2015
PRELIM. EQUIPMENT LIST	PLANT **PTA PANIPAT UPRATE 92.1te/h PTA**		LOCATION **INDIA**	PLANT SECTION **OXIDATION - Crystallisation & Separation**	

EQUIP		SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES (DegC)**	PRESS **DES (kg/cm^2g)**	DIMENSIONS (mm)	**REMARKS**
21 B1	**420C**		CTA CAKE DISCHARGE SCREW	1 (See remarks)						Additional equipment to be identical to two already installed.
21 B1	**421C**		CTA CROSS SCREW CONVEYOR	1 (See remarks)						Additional equipment to be identical to two already installed.

INVISTA Performance Technologies
The Wilton Centre, Wilton
Redcar
TS10 4RF

DOC	SECTION	REV	NOTE		NAME	DATE
	20610/00/15/00600 A	**Z2**	THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	INPUT BY	CHJ	05/08/2015
				CHECKED BY	RMS	24/09/2015
				APPROVED BY	GMcC	28/09/2015

	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**OXIDATION - Solvent Treatment (OPTION A)**

EQUIP		SUB SET	EQUIPMENT TITLE	NO OFF	MOC	CAPACITY EACH	TEMP DES (DegC)	PRESS DES (kg/cm^2g)	DIMENSIONS (mm)	REMARKS
21 C1	**631**	TOP	RECOVERY COLUMN	1						Repalcement for existing 21C1-631
21 C1	**631**	BOTTOM	RECOVERY COLUMN	1						Repalcement for existing 21C1-631
21 E1	**639**	SHELL	VENT CONDENSER	1						TEMA type BEM
		TUBE								

INVISTA Performance Technologies
The Wilton Centre, Wilton
Redcar
TS10 4RF

DOC	SECTION	REV	NOTE		NAME	DATE
	20610/00/15/00600 B	**Z2**	THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	INPUT BY	CHJ	05/08/2015
				CHECKED BY	RMS	24/09/2015
				APPROVED BY	GMcC	28/09/2015

	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**OXIDATION - SOLVENT TREATMENT & PPML EXTRACTION (OPTION B)**

EQUIP	SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES (DegC)**	PRESS **DES (kg/cm^2g)**	DIMENSIONS (mm)	**REMARKS**	
									INFORMATION DELETED	P2
										P2

INVISTA Performance Technologies
The Wilton Centre, Wilton
Redcar
TS10 4RF

DOC	SECTION	REV	NOTE		NAME	DATE
	20610/00/15/00900	**Z2**	THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	INPUT BY	CHJ	05/08/2015
				CHECKED BY	RMS	24/09/2015
				APPROVED BY	GMcC	28/09/2015

	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**OXIDATION - CTA Handling & Offgas Drying**

EQUIP		SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES (DegC)**	PRESS **DES** **(kg/cm^2g)**		DIMENSIONS (mm)	**REMARKS**
21 B1	**902A/B**	BODY	CTA DRIER EXIT ROTARY VALVE	1W/1S							
		JACKET									

INVISTA Performance Technologies
The Wilton Centre, Wilton
Redcar
TS10 4RF

DOC	SECTION	REV	NOTE		NAME	DATE
		Z2	THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	INPUT BY	CHJ	05/08/2015
	20610/00/15/01200			CHECKED BY	RMS	24/09/2015
				APPROVED BY	GMcC	28/09/2015

DOC	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**PURIFICATION - Feed Preparation**

EQUIP		SUB SET	EQUIPMENT TITLE	NO OFF	MOC	CAPACITY EACH	TEMP DES (DegC)	PRESS DES (kg/cm^2g)	DIMENSIONS (mm)	REMARKS
21 P1	1207A/B		LP DISSOLVER FEED PUMP	1W/1S						New impellors for existing pumps ONLY. Existing motors and casings appear adequate.
21 P1	1209E		HP DISSOLVER FEED PUMP	1 (See remarks)						New pump to be operated in parallel with existing pumps on 4W/1S basis.
21 E1	1211C	SHELL	THIRD PREHEATER	1						
		TUBE								

INVISTA Performance Technologies
The Wilton Centre, Wilton
Redcar
TS10 4RF

DOC	**SECTION** 20610/00/15/01300	REV **Z2**	NOTE THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	

	NAME	DATE
INPUT BY	CHJ	05/08/2015
CHECKED BY	RMS	24/09/2015
APPROVED BY	GMcC	28/09/2015

	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**PURIFICATION - Reaction**

EQUIP	SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES (DegC)**	PRESS **DES** **(kg/cm^2g)**		DIMENSIONS (mm)	**REMARKS**
21 R1 1301		DISSOLVER/ REACTOR	1							NOTE - Modifications to downcomer to increase area available for mass transfer. No change to exisiting pressure vessel.

This drawing is a private and proprietary communication and the property of INVISTA. It must not be copied or loaned without the consent of INVISTA and must be returned.

INVISTA Performance Technologies The Wilton Centre, Wilton Redcar TS10 4RF	DOC	**SECTION** **20610/00/15/01400 A**	REV	**Z2**	NOTE THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN		NAME	DATE
						INPUT BY	CHJ	05/08/2015
						CHECKED BY	RMS	24/09/2015
						APPROVED BY	GMcC	28/09/2015

	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**PURIFICATION - Crystallisation & Pressure Separation (OPTION A)**

EQUIP		SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES (DegC)**	PRESS **DES (kg/cm²g)**	DIMENSIONS (mm)	**REMARKS**
21 A1	**1401**		FIRST PTA CRYSTALLISER AGITATOR	1						
21 CR1	**1401**		FIRST PTA CRYSTALLISER	1						
21 V1	**1415**		PTA MOTHER LIQUOR DRUM	1						Additional Atmospheric Centrifuge option only.
21 V1	**1416**		PTA MOTHER LIQUOR DRUM SCRUBBER	1						Additional Atmospheric Centrifuge option only.
21 P1	**1420A/B**		ATMOSPHERIC CENTRIFUGE FEED PUMP	1W/1S						Additional Atmospheric Centrifuge option only. New impellors for existing pumps.
21 CF1	**1421C**		ATMOSPHERIC CENTRIFUGE	1						Additional Atmospheric Centrifuge option only. Horizontal decanter.
21 B1	**1422C**		ATMOSPHERIC CENTRIFUGE DISCHARGE SCREW	1						Additional Atmospheric Centrifuge option only.

INVISTA Performance Technologies The Wilton Centre, Wilton Redcar TS10 4RF	DOC	**SECTION** **20610/00/15/01400 B**	REV **Z2**	NOTE THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	NAME / DATE INPUT BY CHJ 05/08/2015 CHECKED BY RMS 24/09/2015 APPROVED BY GMcC 28/09/2015

	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**PURIFICATION - Crystallisation & Pressure Separation (OPTION B)**

EQUIP		SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES (DegC)**	PRESS **DES (kg/cm^2g)**	DIMENSIONS (mm)	**REMARKS**
21 A1	**1401**		FIRST PTA CRYSTALLISER AGITATOR	1						
21 CR1	**1401**		FIRST PTA CRYSTALLISER	1						
21 P1	**1420A/B**		RPF FEED PUMP	1W/1S						Hydraulic RPF option only. Replacement pumps and motors.

INVISTA Performance Technologies The Wilton Centre, Wilton Redcar TS10 4RF	DOC	**SECTION** **20610/00/15/01400 B**	REV **Z2**	NOTE THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	NAME	DATE
				INPUT BY	CHJ	05/08/2015
				CHECKED BY	RMS	24/09/2015
				APPROVED BY	GMcC	28/09/2015

	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**PURIFICATION - Crystallisation & Pressure Separation (OPTION B)**

EQUIP		SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES (DegC)**	PRESS **DES** **(kg/cm^{2}g)**	DIMENSIONS (mm)	**REMARKS**	
21 V1	**1432**		MOTHER LIQUOR FILTRATE RECEIVER	1						Hydraulic RPF option only.	
21 V1	**1433**		DRYING GAS FILTRATE RECEIVER	1						Hydraulic RPF option only.	
21 V1	**1434**		RPF DRAININGS RECEIVER	1						Hydraulic RPF option only.	
21 P1	**1435A/B**		RPF DRAININGS RECYCLE PUMP	1W/1S						Hydraulic RPF option only.	
21 V1	**1436**		PTA FLASH DRUM SCRUBBER	1						Hydraulic RPF option only.	
21 A1	**1437**		PTA MOTHER LIQUOR FLASH DRUM AGITATOR	1						Hydraulic RPF option only.	P2
21 V1	**1437**		PTA MOTHER LIQUOR FLASH DRUM	1						Hydraulic RPF option only.	
21 B1	**1439**		RPF CAKE DISCHARGE	1						Hydraulic RPF option only.	

INVISTA Performance Technologies The Wilton Centre, Wilton Redcar TS10 4RF	DOC	**SECTION** **20610/00/15/01400 B**	REV **Z2**	NOTE THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN		NAME / DATE INPUT BY CHJ 05/08/2015 CHECKED BY RMS 24/09/2015 APPROVED BY GMcC 28/09/2015
PRELIM. EQUIPMENT LIST	PLANT	**PTA PANIPAT UPRATE 92.1te/h PTA**	LOCATION	**INDIA**	PLANT SECTION	**PURIFICATION - Crystallisation & Pressure Separation (OPTION B)**

EQUIP		SUB SET	EQUIPMENT TITLE	NO OFF	MOC	CAPACITY EACH	TEMP DES (DegC)	PRESS DES (kg/cm^2g)	DIMENSIONS (mm)	REMARKS
21 RF1	**1440**		ROTARY PRESSURE FILTER	1						Hydraulic RPF option only
21 B1	**1441**		PTA DRIER CROSS-SCREW	1						Hydraulic RPF option only

INVISTA Performance Technologies
The Wilton Centre, Wilton
Redcar
TS10 4RF

DOC	**SECTION**	REV	**Z2**	NOTE	
PRELIM. EQUIPMENT LIST	**20610/00/15/01600**			THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN	

	NAME	DATE
INPUT BY	CHJ	05/08/2015
CHECKED BY	RMS	24/09/2015
APPROVED BY	GMcC	28/09/2015

PLANT	LOCATION	PLANT SECTION
PTA PANIPAT UPRATE 92.1te/h PTA	**INDIA**	**PURIFICATION - Mother Liquor Treatment & Vents Treatment**

EQUIP		SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES (DegC)**	PRESS **DES (kg/cm^2g)**	DIMENSIONS (mm)	**REMARKS**	
21 P1	**1602**		PTA MOTHER LIQUOR FILTER FEED PUMPS	1W/1S						New impellors and motors for existing pumps.	
21 E1	**1607E**	PROCESS	PTA MOTHER LIQUOR COOLER	1 (See Remarks)						OPTION A - Single additional spiral cooler, same size as existing (4) units	P2
		CW									
21 E1	**1607**	SHELL	PTA MOTHER LIQUOR COOLER	1						OPTION B - Alternative to spiral coolers. Sized for total E1-1607 duty to replace 21E1-1607A-E.	P2
		TUBE									
21 V1	**1608**		PTA MOTHER LIQUOR BUFFER VESSEL	1							
21 P1	**1609A/B**		PTA MOTHER LIQUOR TRANSFER PUMP	1W/1S							

INVISTA Performance Technologies
The Wilton Centre, Wilton
Redcar
TS10 4RF

DOC	**SECTION**	REV	**Z2**
	20610/00/15/01900		

NOTE

THE INFORMATION GIVEN ON THIS SHEET IS PRELIMINARY AND MAY BE SUBJECT TO CHANGE DURING FEEP DESIGN

	NAME	DATE
INPUT BY	CHJ	05/08/2015
CHECKED BY	RMS	24/09/2015
APPROVED BY	GMcC	28/09/2015

	PLANT	LOCATION	PLANT SECTION
PRELIM. EQUIPMENT LIST	**PTA PANIPAT UPRATE 92.1te/h PTA**	**INDIA**	**PURIFICATION - Product Handling**

EQUIP		SUB SET	**EQUIPMENT TITLE**	NO OFF	**MOC**	CAPACITY EACH	TEMP **DES (DegC)**	PRESS **DES (kg/cm^2g)**		DIMENSIONS (mm)	**REMARKS**	
21 B1	**1902**	BODY	PTA DRIER EXIT ROTARY VALVE	1								P3
		JACKET										

6.0 Estimated Impact on Utility Requirements

Debottleneck of the PTA Plant to 92.1 t/h PTA is expected to only have a minor impact on the consumption of utilities per tonne of PTA product.

6.1 Electricity

Excluding the Process Air Compressor, the consumption of electricity post-debottleneck is expected to be essentially neutral. The duties on the main motor drives on the plant will largely pro-rate with the increased operating rate. Agitator powers will remain as current, hence there will be a slight fall in the consumption of power per tonne of PTA at the debottleneck rate.

Power consumption by the Process Air Compressor differs for the air and oxygen based debottlenecks. It should be noted that is estimated that the under-performance (inefficiency) of the PAC in its current state could be worth in excess of 1 MW of power.

If it is assumed that the PAC can be restored to its expected efficiency, the expected additional power consumption by the PAC is estimated at ~ 5.8 MW for the air debottleneck case. The gain in power from the Offgas Expander for the air debottleneck case of ~ 3 MW. Hence the overall impact is ~ 2.8 MW.

Both the air and oxygen debottlenecks result in the generation of additional ELP steam which is fed to a new stand-alone Steam Turbine-Generator (21-KT1-150). The estimated power generation from 21-KT1-150 based solely on the additional steam generated by the debottleneck (and excluding voluntary Site LP Steam) is ~ 2.7 MW.

Summarising:

	PTA Plant	Existing PAC	New 21-KT1-150	Overall
Air based debottleneck	Neutral to marginal gain	+2.8 MW	-2.7 MW	Neutral to marginal gain
Oxygen based debottleneck	Neutral to marginal gain	Approximately neutral	-2.7 MW	-2.7 MW

Impact of Options on Electricity Usage

Replacing the centrifuges on the Purification Plant with a single stage Rotary Pressure Filter may result in a reduction in ISBL power consumption of ~ 1MW at 92.1 t/h PTA rate.

E^2R technology is estimated to be neutral in ISBL power consumption. Additional power consumption of new and replacement pumps is offset by additional power from 21-KT1-150 arising from reduced LP steam consumption in Area 600.

The Suction Chiller package, if installed (and depending on the final suction temperature selected) will consume 5-10 t/h LP steam, which (together with the power penalty resulting arising from the suction chiller pressure drop) is equivalent to 0.6-1.2 MW of additional 'lost' power recovery in the Steam Turbine.

6.2 Site Steam

The debottleneck is expected to have only a minor impact on steam consumption from OSBL. Site HP Steam consumption is expected to increase pro-rata with plant rate (from current ~1.2 t/h to ~1.5 t/h). No change in site MP steam consumption is expected based on improved performance of the CTA Drier with additional upstream Rotary Vacuum Filter capacity. 21-KT1-150 will be sized to accept up to 50 t/h of Site LP Steam.

6.3 Cooling Water

The flow to existing users is unchanged by the debottleneck, although the temperature difference between supply and return will rise approximately pro-rata with the debottleneck rate increase.

The following new users of cooling water are specified in the debottleneck equipment list.

	Approximate Cooling Water Flow (t/h)	Comments
2nd Steam Turbine Condenser (21-KT1-151)		
Vacuum Vapour Condenser (21-E1-415C)		
Vent Condenser (21-E1-639)		
PTA Mother Liquor Cooler (21-E1-1607E)		
Suction Chiller package (21-X1-112)		
TOTAL		

6.4 Demineralised Water

For both the additional Atmospheric Centrifuge and single stage Rotary Pressure Filter options the demineralised water consumption will increase pro-rata with the plant rate increase.

6.5 Hot Oil

The total heat duty on 21-E1-1211A/B/C is estimated to increase to [DELETED] MW (absorbed heat) at the debottleneck rate. This is believed to be within the capacity of the Hot Oil system (~[DELETED] MW).

6.6 Others

Usages of raw water, nitrogen and instrument air are expected to be unchanged by the debottleneck. The usage of inert gas will marginally increase, but will still be within the capacity of the current Offgas Driers.

7.0 Estimated Impact on Raw Material Efficiencies

	Expected Change	Comment
Paraxylene	.	Impact of increased reactor temperature / pressure in order to stay within entrainment factor known band of operation is offset by reduction in Oxidation Reactor residence time. Assumes that product 4CBA can be maintained as current when at debottleneck rate. There may be some slight deterioration due to less efficient Oxidation reaction, or worse performance of M/L filter. Note: The option of using INVISTA's E²R technology would reduce paraxylene usage.
Acetic		Comments as for paraxylene.
Cobalt, Manganese, Bromide, Oxalic Acid		
Entrainer	No change	
Hydrogen	No change	
Oxygen (1)		

Notes:

(1) Oxygen is imported to ISBL by pipeline and shall meeting the following typical specification:

Oxygen concentration	99.5 mole% min.
Supply pressure	18.2 kg/cm^2g
Design pressure	21.4 kg/cm^2g
Supply temperature	Ambient
Design temperature	[DELETED]

8.0 Yield Estimates & Equipment Requiring Modification

As stated in section 6, the yield of paraxylene to Pure Terephthalic Acid is expected to be unchanged as result of the debottleneck. Based on the best performance of PTA plants of this generation running at rates close to the debottleneck rate, a paraxylene consumption of [DELETED] is believed to be achievable with all equipment operating at their design intent. The inclusion of INVISTA's E^2R technology would reduce this paraxylene consumption by [DELETED].

Equipment requiring modification are described in sections 3, 4 and 5.

9.0 Estimate of Incremental Effluents

9.1 Gaseous Emissions

The offgas flow will increase but this will all continue to be treated through the HPCCU and Offgas Scrubber. It is expected that concentrations of components emitted will not increase significantly but the total mass flow of gaseous emissions will increase proportionate to the plant rate. The total emission from 21-C1-172 will increase to approximately 175,000 Nm^3/h for the air based uprate.

If the vent from First CTA Crystalliser 21-CR1-401 continues to be fed to Atmospheric Scrubber 21-C1-508, flows from this vent will increase proportional to the plant rate increase. If the 21-CR1-401 vent is fed to 21-C1-310, the flow from 21-C1-508 will reduce by approximately 4900 Nm^3/h at debottleneck rates.

There will be an increase in flow from the Feed Hopper Vent Gas De-duster (21-G1-1202), estimated at ~ 2000 Nm^3/h. Concentration of components should be unchanged by the increase in flow.

There will also be an increase in flow from the PTA Batch Tanks. This increase in flow is estimated at 500 Nm^3/h. Concentration of components should be unchanged by the increase in flow.

Emissions of steam from the Vent Scrubber will increase pro-rata with the increase in plant rate. Use of flash steam on 21-C1-631 would reduce the flow by approximately 5 t/h. Emissions of steam from the PTA Drier should remain approximately constant, on the assumption that the performance of the PTA Drier can be maintained by the installation of additional upstream PTA separating capacity.

Flows from the PTA Product Silos should be unchanged by the debottleneck.

9.2 Liquid Effluent

Total liquid effluent flow and total COD will increase in proportion to the plant rate. This assumes that the PTA Mother Liquor Filter 21-G1-1603 is capable of meeting current performance with the higher liquid flow-rate. See section 8.

Implementation of INVISTA's E^2R technology would reduce liquid effluent [DELETED] at debottleneck rates.

9.3 Solid Effluent

The amount of solid effluent arising from the Oxidation Plant will increase in proportion to the plant rate.

10.0 Summary of Revamp Operating Conditions, Product Specification and Overall Material Balance

10.1 Summary of Revamp Operating Conditions

It is expected that the PTA Plant will operate within its current operating envelope following the debottleneck. The increase in plant rate will necessitate an increase in the minimum Oxidation Reactor operating pressure. The minimum operating pressure will increase to [DELETED]. As stated in section 10, yield of paraxylene to Pure Terephthalic Acid should be [DELETED].

Operating conditions through the rest of Oxidation Plant will remain as current, with the exception of 21-CR-401 where the operating pressure will need to be raised to approximately [DELETED] in order to allow introduction of vent gas to 21-C1-310.

Operation conditions in Purification Plant (including inlet temperature to 21-R1-1301 and PTA Crystalliser operating pressures) will remain as current operation.

10.2 Product Specification

Product specification will be unchanged by the debottleneck:

Appearance	White, dry free-flowing crystals	
Acid number	675+/-2	mg KOH/g
4-Carboxybenzaldehyde	25	ppm wt. max
Paratoluic Acid	125 +/- 45	ppm wt.
Ash	6	ppm wt. max
Total significant heavy metals (Ti, Cr, Fe, Co, Ni, Mo, Mn)	6	ppm wt. max
Moisture	0.2	% w/w max
Colour (b* value, CIE lab 1976)	1.4	max
Colour (in 2N KOH solution)	10	Hazen units max

10.3 Overall Material Balance

The following is a preliminary overall material balance [DELETED] for an air based debottleneck to 92.1 t/h PTA rate. All numbers are subject to confirmation in any process design phase.

Purification Vents	60.9t/h
Reactor Off Gas and Other Oxidation Vents	226.8t/h
Process Effluent	168t/h
Oxidation residues	1t/h

11.0 Assessment of Technical Uncertainty

The following items of technical uncertainty remain at the end of this feasibility study stage. Overall, as INVISTA has experience of this generation of INVISTA PTA technology operating at the proposed debottleneck rate, INVISTA has a high degree of confidence in where the equipment bottlenecks lie, and has addressed these bottlenecks in the proposed modifications.

11.1 Oxidation Plant

- The PAC is currently underperforming significantly versus its design performance curves. INVISTA believes that the most likely cause is that the intercoolers are badly fouled on the air side. The other possibility is that the compressor is very dirty and the contour rings have corroded and the clearances increased.

  INVISTA has already proposed that IOCL correctly calibrate all the field instruments on the PAC (since the shared pressure and temperature readings are not thermodynamically possible) and the fitting of differential pressure measurement across the three intercoolers.

  Cleaning intercooler bundles on the air side is very difficult (whereas cleaning the compressor and the contour rings is not too difficult) and may require replacement of the worst affected bundles to return the compressor to its original condition.

  It is estimated that replacing the 21-E1-116 bundle would bring a power improvement averaging around 450kW through the year but more importantly would allow the PAC to run at higher rate in summer. Estimated intercooler replacement costs are in Section 12. Payback on energy to replace the 21-E1-116 bundle would be around 2 years. The payback on energy alone for 21-E1-117 and 21-E1-118 is greater than 5 years.

  INVISTA recommend:-

  - Regardless of whether the uprate project is sanctioned IOCL should clean the intercooler bundles and plan to replace the 21-E1-116 bundle if cleaning is not effective. This will payback in energy terms in around 2 years and allow the plant to run at higher rate in summer.
  - Inspect the PAC at the first opportunity to check for any others causes of pressure drop; in particular at the suction and for any other restrictions. Implement regular water wash of at least stages 1 and 2 following inspection.
  - Collect full performance data for the PAC under different ambient conditions and analyse performance. It is important to also record the local ambient pressure and either dewpoint or humidity to enable suction volume to be confirmed.
  - Include intercooler bundle replacement and if required suction chilling in the uprate package. It may be possible to remove intercooler bundles for the project scope if cleaning is successful.
  - Check the venturi flowmeter is correctly calibrated and verify against other plant flowmeters (eg CCU flowmeter when running on 100% bypass air at plant start-up).

  Ignoring the PAC condition and adding oxygen addition will not improve the power performance and the PAC will likely continue to deteriorate.

  As stated previously, installation of the Suction Chiller package will not return the PAC to its original performance.

- The Offgas Expander is predicted to be at its power output limit for the air debottleneck case. Discussions with Man Turbo would be needed as part of any further work phase to see if there is any scope to raise this limit.

- Based on modelling of the Oxidation Reactor overheads for the air debottleneck case, the temperature exit 21-E1-307 is predicted to rise to [DELETED]. A consequence of such a rise would be to increase the slippage of methyl acetate and acetic acid from 21-C1-310 by a small amount resulting in additional variable cost. In practice, INVISTA believe that a rise to [DELETED] is unlikely.

- The feasibility study has included a modification to route the vent gas from 21-CR1-401 to 21-C1-310. The pressure in 21-CR1-401 is estimated to need to rise to an operating pressure of [DELETED]. The min. burst pressure of the 21-CR1-401 bursting discs will need review. Worst case would see a reversion to current mode of operation i.e. venting of 21-CR1-401 via 21-C1-508.

  The CTA Drier is currently underperforming. It has been assumed that performance can be restored without further capital expenditure to meet the needs of the debottleneck. The cost of replacing the CTA drier would be significant. The estimated marginal increase of all options is provided in Section 12.0 Option F – Marginal Cost of replacing CTA Drier.

  Existing CTA Drier Exit Rotary Valve 21-B1-902A/B, may have sufficient capacity for the debottleneck case. A plant modification followed by plant trials would be required to confirm the capacity of the existing valves. The trial carries some risk to the CTA Drier seal. For this reason INVISTA has assumed that new, larger valves will be installed.

- The capacity of 21-C1-601 is believed to be sufficient based on performance of this diameter of column and type of packing at other plants at the debottleneck rate. The issue would be explored further with the packing vendor as part of any further work phase.

- At current rates, DH Column Reboiler 21-E1-602 over-performs based on modelling of this heat exchanger in HTRI. This over-performance forms the basis for INVISTA's belief that the existing unit is sufficient for the debottleneck.

- 21-C1-631 is only marginally too small for the debottleneck, necessitating replacement. Further discussion with the column internals vendor as part of a future phase may allow retention of the existing column.

11.2 Purification Plant

- NPSH requirements of modified impellors in 21-P1-1207A/B will need further investigation as part of any future phase of work.

- Existing 21-CR1-1401 has too small a diameter based on INVISTA's experience of acceptable operation at the debottleneck rate. Replacing 21-CR1-1401 has a huge impact on the capital cost of the debottleneck project, but risk of not replacing is believed to be too high to avoid this replacement.

- The PTA Drier is currently underperforming. It has been assumed that performance can be restored without further capital expenditure to meet the needs of the debottleneck.

- 21-V1-1601 is relatively small compared to the size on other plants operating at the debottleneck rate, however observations during the November 2015 Customer Service Visit have highlighted a potential issue with the level indication, and it is likely that the existing vessel capacity is sufficient, even at the debottleneck capacity. Due to the cost of replacement, INVISTA has not included a new vessel on the debottleneck equipment list. The existing vessel may overflow more often if operated at the debottleneck rate. A larger 21-V1-1415 installed as part of the project mitigates this risk somewhat.

- Based on sizing calculations, it is believed that 21-G1-1603 should have sufficient capacity for the debottleneck. A consequence of insufficient capacity would be a small reduction in paraxylene yield. The capital cost of an additional filter is believed to far exceed potential variable cost benefits.

11.3 Additional On-plot

Additional on-plot systems are believed to be capable of meeting debottleneck requirements. This would be rechecked as part of any future phase of work. Technical uncertainty in these systems is considered to be low.

12.0 ISBL Capital Cost Estimate

[DELETED]

Attachment 1
[DELETED]

INDIAN OIL CORPORATION LTD

Panipat Petrochemical complex, Panipat, India

Pre-Feasibility Report

for

Expansion of Petrochemical Plants

and

New Catalyst Plant

JUNE 2016

CONTENTS

INTRODUCTION

Indian Oil Corporation Ltd. (IOCL) intends to expand the Naphtha Cracker Unit including associated units from the current 800 KTA ethylene @ 8000 hr/yr to 947 KTA ethylene (including ethylene from OSBL recycle streams) @8000 hr/yr. CBI Lummus is the licensor of the plant. Part of this expansion is to consider additional feed from an ethane/ethylene recovery unit. Downstream of NCU, associated units, MEG unit and polymer plants exist.

The petrochemical complex consists of the following existing plants & related utilities which are in operation.

- Naphtha Cracker Unit (NCU) including associated units
- High Density Polyethylene (HDPE) Unit
- Swing (LLDPE/HDPE) Unit
- Ethylene Glycol (EG) Plant
- Polypropylene (PP) Unit
- Utilities including Captive Power Plant

IOCL is currently planning expansions of its production capacities for NCU, MEG, HDPE and PP plant.

Proposed Expansion Capacities

Plant	Existing (KTA)	Proposed (KTA)
NCU (in terms of ethylene)	800	947
MEG (in terms of product)	300	425.5
HDPE (in terms of product)	300	351
PP (in terms of product)	600	780

IOC also intends to setup a Catalyst plant for production of FCC (Fluidized Catalytic Cracker) additives and hydro-treating (DHDS/DHDT) Diesel Hydro De-Sulfurisation/ Diesel Hydro Treater catalyst which will be situated in current PNC complex.

PROJECT INFORMATION

Indian Oil Corporation Ltd. (IOCL) is proposing to expand the existing units which includes NCU, MEG, HDPE and PP. Also IOCL is proposing to put up a catalyst manufacturing plant. The existing complex is spread over 306 HA hectares and there is sufficient vacant space to accommodate these proposed expansion projects.

Proposed Projects Justification

India, with a population exceeding 1.2 billion, and a large and steadily growing middle class with rising disposable income, is forecast to be among the world's five largest consumer markets by 2025. Polymer consumption is closely linked to GDP growth and India is the major economy with the second-fastest growth rate after China. In the competitive scenario of petrochemical business, profitability of petrochemical complex depends on cost of production of olefins. It is observed that cost of production of olefins decrease with increase in capacity. It appears from the market study, that Domestic demand is expected to grow leading to substantial gap in supply/demand. In view of the above, we should seize the opportunity of NCU, MEG, HDPE and PP revamp at present for production of additional olefins to penetrate the deficit market for improving margin.

The following feature also justifies the PNC complex as the best alternative as regards to the project site, infrastructure, connectivity and market potential..

- The products manufactured meets market specification.
- Adequate safety systems are in design to handle to any operational upset.
- Availability of the required land for the proposed expansion.
- Availability of infrastructure facilities
- Availability of Electric Power through captive power plant.
- Environmental management:

Existing plants already have robust management systems for quality, environment and occupational health and safety which are certified against the standards ISO 9001, ISO 14001 and OHSAS 18001 respectively.

- The wastewater from proposed projects will be treated in the existing ETP.
- Greenbelt development: There exists a well-established greenbelt area, which is adequate for attenuation of air emissions and noise levels.

IOC R&D over the years has developed significant expertise in FCC catalysts & additives as well as DHDS and DHDT catalysts. These formulations are currently commercially offered through SCIL (Sud Chemie India Ltd) by a licensing agreement. Considering IOC R&D's proven capability/ expertise and as a strategic initiative IOCL has decided to put-up a catalyst manufacturing plant on its own for captive consumption.

PROJECT AREA

Area within Fencing	= 306 HA
Green Belt	= 55 HA
Area within Compound Wall	= 251 HA
Pet coke Area	= 40HA
Balance Area for PNCP	=211 HA

FUTURE EXPANSION AREA

Total Area within Fencing	= 102.5 HA (253 ACRE)

PROCESS DESCRIPTION

Naphtha Cracker Unit :

Naphtha Cracker Unit is the mother unit of the entire complex .The associated units with NCU at Panipat is C4 hydrogenation, Pyrolysis gasoline hydrogenation unit, Butadiene extraction unit and Benzene extraction unit. In NCU, low aromatic naphtha cracks into lighter hydrocarbons in cracking heaters, which are then individually separated by fractionation to produce mainly polymer grade ethylene and polymer grade propylene. It also produces hydrogen, methane off gasses ,pyrolysis fuel oil and other products like raw mix C4 ,raw pyrolysis gasoline that are further processed in Associated Units . Ethane and propane, produced in the process, are recycled back to cracking heaters.

Naphtha feed is received from storage tanks outside battery limits, filtered and mixed with hydrogenated C4, C5 and C6 recycle streams, prior to being sent to the liquid cracking heaters. All the SRT heaters operate using fuel gas only. The primary fuel gas is methane rich off gas produced in the naphtha cracker unit. Make-up/ back-up fuel is C3/C4 LPG supplied from adjacent refinery or RLNG vapor.

In the Gasoline Fractionator, the cracked effluent gases are further cooled, pyrolysis fuel oil (PFO) is separated as a bottoms product, a side stream product-pyrolysis gas oil (PGO) is withdrawn from the Fractionator, gasoline and lighter materials are taken as an overhead vapor. Pyrolysis fuel oil product is mixed with heavy C9+ cut from the return tower of the Pyrolysis Gasoline Hydrogenation unit and is cooled prior to being sent to OSBL storage. A side stream product withdrawn from the Fractionator is sent to the PGO Stripper where it is steam stabilized .A portion of the stabilized PGO is filtered and used as purge oil for instruments, the rest is blended with fuel oil and cooled before being sent to OSBL storage.

Overhead vapor from the Gasoline Fractionator is cooled and partially condensed in the Quench Tower. The overhead vapor from the Quench Tower is sent to the Charge Gas Compressor. Recycle streams from downstream swing PE plant, HDPE plant and PP plant and off spec ethylene vapors from HP ethylene storage as well as internal recycle streams of the naphtha cracker are also reprocessed in the Charge Gas Compressor system. The quench tower overhead vapors are compressed in a five-stage centrifugal compressor with interstage cooling and is dried in a two bed molecular sieve drying system .The dry charge gas is progressively chilled against the process and propylene and ethylene refrigeration.

The condensed liquids from the charge gas chilling train along with the vent gas from ethylene fractionation and propylene fractionation, and light gas recycle from PP plant are sent to the appropriate feed locations of the demethanizer. Here, the residual gas is hydrogen of approximately 75 mol% purity. The hydrogen rich stream is further upgraded to 95+ mol% purity in an adiabatic heat exchange system. The raw hydrogen generated is used in the two primary processing steps: Methanation of CO in hydrogen to methane and water, as CO is a catalyst poison and in Drying of hydrogen; required for the hydrogenation reactors water is a poison to these catalysts.

Provision for reprocessing off spec ethylene is also provided in the demethanizer. Reflux to this tower is provided by an open loop refrigeration system which utilizes a motor driven centrifugal compressor. The demethanizer bottoms product feeds the deethanizer which also processes FCC dry gas C2s and FCC C3 streams received from Panipat refinery. Acetylene is removed from the net deethanizer overhead product by selective hydrogenation to ethylene and ethane, which is sent to Ethylene Fractionator. The ethylene products handled are: One stream is chilled and delivered to low pressure OSBL cryogenic storage.MP ethylene, HP Ethylene & Ethane.

The purpose of the depropaniser is to make a sharp separation between C3 components and heavier components in the Deethanizer bottoms and condensate stripper bottoms. Methyl acetylene and propadiene contained in the deprpopanizer overhead are removed by selective hydrogenation to Propylene and Propane in a single bed reactor. The net C3 product is sent to Propylene Fractionator where Polymer grade propylene is the distillate with bottom product containing primarily Propane.

The bottom product from the depropanizer flows to the debutanizer where the C4s product is separated. The debutanizer net overhead product, consisting of mixed C4s is pumped to the butadiene extraction unit, C4 hydrogenation unit , or to OSBL storage for further processing. The bottom product is combined with gasoline from Gasoline stripper bottom to make up the total raw pyrolysis gasoline product and is sent to the Pyrolysis gasoline hydrogenation unit.

Butadiene extraction unit recovers 1,3 butadiene from the raw mixed C4 stream produced in the NCU. BD raffinate is sent to hydrogenation unit.

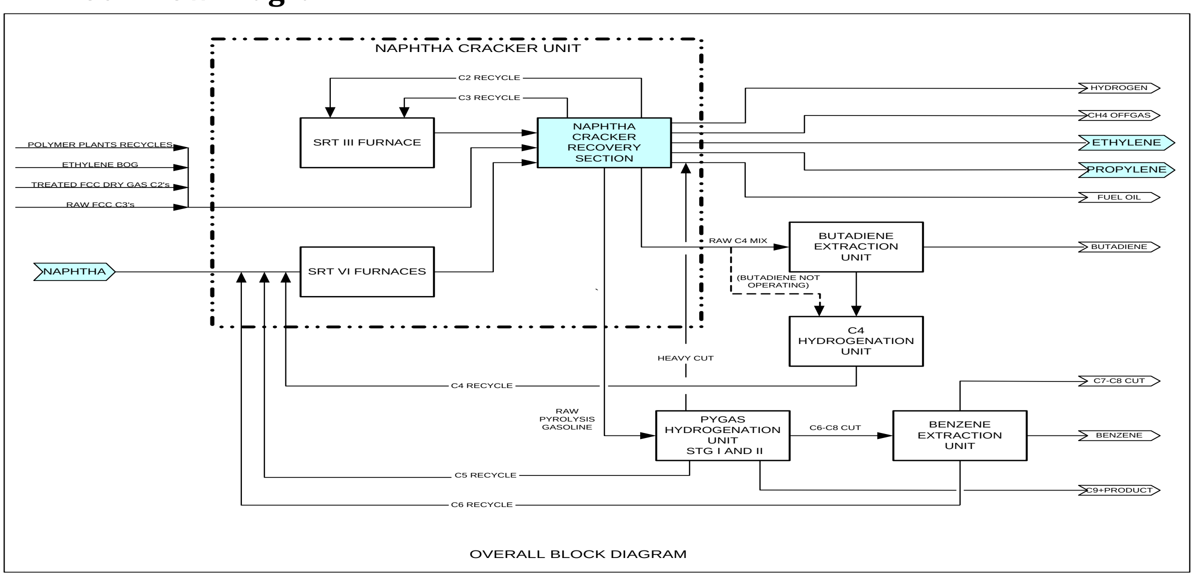
C4 hydrogenation unit is designed to fully hydrogenate Butadiene raffinate in normal operation. Hydrogenated C4s is recycled to the NCU cracking heaters. In alternate operation, when BDEU is not operating, C4HU will process raw C4 mix. Butadiene and a portion of Butenes are hydrogenated in this mode of the operation.

Pyrolysis Gasoline Hydrogenation Unit processes raw pyrolysis gasoline from NCU in a two stage hydrotreating unit to produce C6-C8 heart cut that is sent to Benzene extraction Unit for benzene recovery, a fully hydrogenated C5 cut that is recycled to the cracking heaters and a partially hydrogenated c9+ product.

Proposed Modification:

Panipat Refinery Delayed Coker unit (DCU) and Fluidized Catalytic Cracking (FCC) units off gases are presently routed to Refinery Fuel gas , contain useful Ethylene, Ethane, Propylene & Propane which can be recovered and sent to Naphtha cracker unit .In this project, integration of ERU (ethylene recovery unit) with the existing system will take place and by this cracker capacity will be augmented. To support this capacity enhancement, modifications / additions in exchangers, pumps, vessels, column internals is being done. One furnace will be added in this project.

Block Flow Diagram :



OVERALL BLOCK DIAGRAM

Mono Ethyl Glycol Unit :

IOCL has installed a MEG unit at Panipat Naphtha Cracker Complex, having a capacity to produce 300KTA fiber grade MEG. DEG and TEG are also produced from the unit.

In the process ethylene is partially oxidized by oxygen in the presence of silver catalyst to make ethylene oxide (EO) in an exothermic reaction. The EO produced in the reactor is recovered by lean cycle water from reactor affluent stream in high pressure scrubber column and then it is again separated from rich cycle water by steam stripping in stripper. Then EO rich gaseous stream is absorbed with recycle water in low pressure reabsorber. Then 10 wt. % EO is hydrolyzed in glycol reactor at an elevated temperature and pressure by adiabatic, non-catalytic reaction to form MEG, DEG and TEG. So formed dilute glycol is concentrated in series of seven forward feed evaporators to form Crude glycol and then distilled in MEG, DEG and TEG column respectively to produce fiber grade MEG as products and DEG, TEG and HG as by-Products.

Proposed Modification:

Modifications proposed are based on a review of plant hydraulic load data, as-built equipment efficiency, preliminary process calculations and computer simulations performed by SD. To achieve the mentioned capacities, several equipments shall undergo some modifications; although no modifications have been envisaged in major equipments like recycle gas compressor and EO reactors. Some new equipments shall be added to the unit to cater to higher load. Few of the new equipments envisaged are:

1) Additional Oxygen filters – To cater to higher oxygen flow rate.
2) An ethylene recovery unit – To minimize ethylene loss from argon purge from cycle gas flow.
3) Regenerator trim condenser – To achieve design outlet temperature of 60 Deg.C
4) Cycle Water Exchanger – To avoid leakage
5) Cycle Water Chiller – For better absorption of EO in cycle water in the scrubber

High Density Polyethylene Unit (HDPE) :

IOCL has installed a HDPE unit at Panipat Naphtha Cracker Complex based on M/s Lyondell Basell's Hostalen technology, having a capacity to produce 300,000 MTPA HDPE.

HDPE is produced by continuous polymerization of monomer (Ethylene sourced from Naphtha Cracker unit) and co-monomer (Butene) in presence of catalyst, co-catalyst and solvent in reactors. After polymerization, the solvent is separated from the polymer by decanters. The solvent is sent to recovery section, wherein wax generated during polymerisation is separated from solvent and solvent is recycled back to process. The polymer separated in the decanters is dried in Drying section and then sent to Extrusion section. The polymer is extruded in Extruder along with additives for stabilization of polymer. The polymer is stored and homogenized in silos and then sent for bagging and packaging.

Proposed Modification:

To achieve the mentioned capacities and to cater for higher load, several equipments shall undergo modifications and some new equipment shall also be added to the unit. Few of the equipments wherein addition/modification are envisaged:

1) Flash vessel
2) Dryer
3) Evaporator
4) Distillation Vessel

Polypropylene Unit (PP):

IOCL Panipat refinery & petrochemicals complex is operating two trains of PP unit (2 x 300,000 MTPA) based on M/s Lyondell Basell's Spheripol technology.

PP is produced by continuous polymerization of monomer (Propylene sourced from Naphtha Cracker Unit) and co-monomer (Ethylene) in presence of catalyst, co-catalyst and selectivity control agent in reactors. After polymerization, the unconverted monomer is separated from polymer in recovery section and is recycled back to process. The polymer separated in the recovery section is dried in Drying section and then sent to Extrusion section. The polymer is extruded in Extruder along with additives for stabilization of polymer. The polymer is stored and homogenized in silos and then sent for bagging and packaging.

Proposed Modification:

To achieve the mentioned capacities and to cater for higher load, several equipments shall be added/modified. Few of the equipments wherein addition/modification are envisaged:

1) Recycle gas compressor
2) Loop reactor coolers and pumps
3) Teal pump
4) Water/organic separator

Catalyst Plant:

The plant is designed for manufacturing Fixed bed and Fluidized bed type catalyst systems used for refinery application. The plant shall be configured to manufacture 500 MTPA of ZSM-5 FCC catalyst additive and 1000 MTPA of DHDS/DHDT catalyst. This plant will also able to produce other FCC additives like Residue Upgradation Additive, CO-combustion promoter additive etc.

The plant shall also be designed to produce key ZSM-5 zeolite and gel alumina required for the manufacture of ZSM-5 additive and alumina support for manufacture of DHDS/DHDT catalyst & FCC additive/ catalyst. Other raw materials like fillers, silicates, hydrated alumina, and acids required are to be directly sourced from market.

SOCIAL BENEFITS

Social Benefits :

These projects, besides general economic desirability, would result in substantial socio-economic benefit to the country in general and more specifically to the region. The socio-economic benefits are described hereinafter.

Social Upliftment of the Region :

This area of the country is undergoing rapid industrialization. Setting up of these projects will be a boon to this region and will bound to improve living conditions and thereby result in further reduction of population below poverty line, which is one of the prime policy objectives of the Government. It is expected that by creation of vast employment potential and industrialization of the area poor/weaker section of the society will see an upliftment in their living conditions.

Employment Generation :

During Construction phase, projects will provide employment to persons, of which significant portion is expected to be drawn from the surrounding areas. On commissioning and achieving successful trial runs, these projects will provide direct employment to local persons and indirect employment in the form of contractors, workers, transporters etc.

ENNVIRONMENTAL PROTECTION

Prevention of Air Pollution :

Emissions from the plant are minimized with the application of nitrogen blanketing for storage tanks, proper selection of pumps as per OSHA standard, proper selection of gaskets, etc.

Prevention of Water Pollution :

As for process waste water, it will be routed to the already existing Effluent treatment plant, which is installed so as to comply as per Haryana State Pollution Control Board (HSPCB) and Central Pollution Control Board (CPCB) guide line.

In order to prevent underground water pollution, the process area is paved with concrete. Oil, acid and rain water falling in the area is collected in a sump for further treatment. This waste water is discharged to the waste water treatment plant for treatment.

Prevention of Noise :

Noise level of working place will be controlled within the limit as specified in the HSPCB standard. In the case that some areas might not satisfy the said standard, suitable countermeasures, e.g. addition of noise insulation, use of PPE etc., will be applied in order to satisfy the said standard. Noise level at plant boundary fence will be controlled to satisfy noise criteria.