UOP Feasibility Study Review for RECOPE S.A. Refinery Expansion Project

Table of Contents

Executive Summary

Honeywell was asked to assist Recope in evaluating the configuration study work being done for their refinery expansion project. Honeywell enlisted some of their UOP configuration specialists to assist in this review to look at the process modeling used in the study while the Honeywell RPMS experts looked at the validity of the LP model solutions

The review of this study shows serious issues both in the LP modeling and process representation that make the study results suspect. The process modeling of some critical units show yields that are significantly different from the range expected. The model also is missing some key adjustments for changes in feed property which should have changed the balances and properties as crude and configuration changed. This invalidates some of the comparisons between different configurations and crude mixes. The combination of poor yields and lack of yield changes in response to feed properties also are leading to error in the needed unit sizes for some of the processes. Some critical product properties like diesel cetane and octane are shown at unreasonably good levels due to fixed properties at the diesel hydrotreater and reformer respectively. It is doubtful that the products from the given equipment would meet some of the desired specifications.

The LP modeling used also shows significant problems. In some cases the product specifications have been relaxed, making comparisons invalid. This may have been due to the model having trouble meeting them. There are also some intermediate stream property specifications that are causing the LP models provided to give significantly incorrect answers.

These issues should be addressed before conclusions are made on the equipment needed for the refinery expansion. Without improvement, using this study as a basis for unit selection and design could lead to units that are incorrectly sized and designed. This would lead to a larger capital cost for less capability than planned and may be insufficient to make the targeted product specifications. Investment figures seem to be high, presenting the Reformer, Diesel Hydrofining Unit and the Delayed Coker the highest differences. Consultant total investment compared to UOP is ~20% bigger.

A more detailed description of the issues we see is given below. Attached balances include notes on data we feel either questionable or in error.

Study Observations

The given study information centered on a Coker-FCC and Coker-HCU configuration. Of these, the Coker-HCU scheme was recommended based on very slightly better economic performance. UOP has been asked to comment on these results.

There are a variety of significant issues associated with the proposed configurations and these are documented in the following discussion points. Many of these issues are of such magnitude as to have potential to significantly impact material balance, refinery margin, unit size, severity and required capital. Since these factors subsequently drive relative financial performance of the various cases, then making a recommendation regarding case performance is unrealistic until these issues are clarified and resolved.

1) Missing or Incorrect Specifications

A variety of key specifications were missing, altered, or relaxed in the LP analysis as indicated in the following table. The suitability of any configuration is uncertain if the analysis does not consider the fundamental requirements.

	Basis Document Specification	**Used in LP**
Premium Gasoline		
DON, minimum	91	88.1
RVP, maximum	69	72
Benzene, maximum	1%	*Not tracked*
Fuel Oil		
Sulfur, maximum	1%	*Not tracked*
Viscosity, maximum	300 ssf	*Not tracked*
ConCarbon maximum	20	*Not tracked*

2) Gasoline octane

Most cases showed 100% premium gasoline production but if reasonable properties as octane number, aromatic content and RVP are substituted for the blendstocks in the given recipes, the result is that 100% premium production is not possible. Production will be a mix of regular and premium that will vary between cases and change the relative profitability of the respective cases. While premium gasoline was one of the highest value products produced by the refinery, regular gasoline was valued as one of the lowest. So, a simple grade shift will have a very significant impact on refinery margin and possibly unit operating/design objectives.

In the LP runs, the CCR reformer unit was run at very high severity (RON=105.2). Normally reformer units targeted for gasoline operation will not be designed for such high severity. Ethanol RON and DON values were reported as 120 and 107.5 in the study. Ethanol blending octane value is a function of the octane of non-oxygenated blend octane and should change from cases to case. Isomerate RON was assumed as 80 in the study which is quite low. Typically isomerate is about 82-83 RON for once-through operation and 87-88 RON for a recycle unit (Penex-DIH type).

Additional properties of the blendstocks used in Cases 1 through 10 are shown in the attached spreadsheets.

3) Gasoline Aromatics

The Coker-HCU configurations will violate the aromatics specification due to the large amount of reformate in the blends. This condition will likely apply to all gasoline grades. The current study used an aromatics content for reformate of 50% and a much more likely value would be around to 75% for the given octane. FCC gasoline aromatics content was reported as 20, but for 94 RON FCC gasoline, the aromatic content would be about 25 vol%.

	LP Blend Vol%	LP Aromatics Vol%	Realistic Aromatics Vol%
Ethanol	2.7	0	1
Reformate	63.1	50	75
HC L Naphtha	7.8	0.1	
DHT Naphtha	10.8	4	
HC H Naphtha	3.5	10	
Isomerate	12.2	0.1	
Resulting Blend Aromatics v%		32.3%	48%, Violates Specification

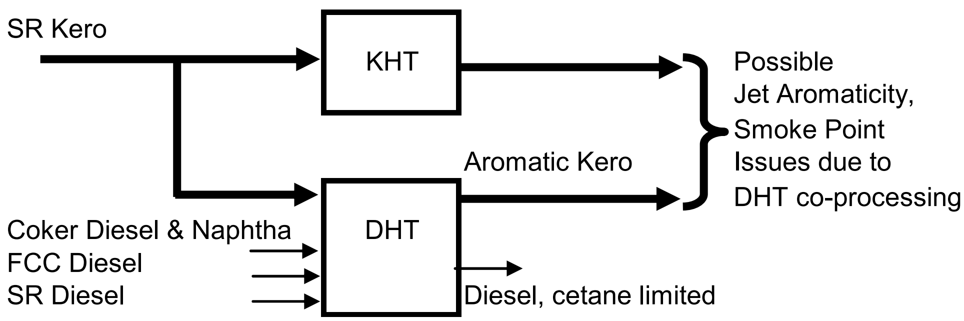
There are methods to adapt the Coker_HCU configuration gasoline aromatics that may include changing the grade mix, using more ethanol for octane as limited by RVP, adjusting reformer severity, naphtha management, increasing non-aromatic octane by adapting the isomerization unit to a deisohexanizer operation, etc. In any case, the current scenario is infeasible, and the remedy will have a significant impact on material balance, cost and financial performance.

4) Gasoline Benzene

Although the study report shows benzene specification of 1.0 vol%, the LP studies did not include benzene specification for any gasoline product. Based on the implicit light naphtha cut point used in the NHT submodel, it is almost certain that all the benzene precursors will go to CCR reformer unit rather than Isomerization unit. Therefore gasoline product will not meet benzene specification regardless of the configuration.

5) Jet Aromatics

It is very likely that at least some of the Coker-FCC scenarios will violate maximum Jet aromatics or smoke point. The current DHT arrangement is to co-process Coker naphtha, straight run kerosene and Coker, FCC and straight run distillates all together at the DHT and fractionate the various products after hydrotreating. The DHT kerosene will be blended with KHT kerosene to form Jet. However, the DHT kerosene will become very aromatic as a result of co-processing with the Coker and FCC distillate. The degree to which this is a concern varies with the aromatics content of the straight run kerosene, but in some cases the contribution of highly aromatic Coker and FCC distillate appears to be as much as nearly one third of the Jet pool.



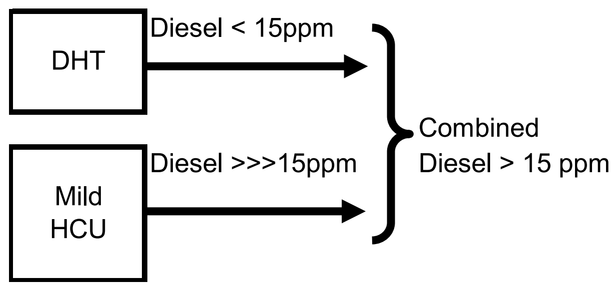
The degree to which Jet/Kero could move to diesel may be limited by flash and cetane. Possibly Jet may be downgraded to kerosene at some penalty to refinery margin for those cases. Possibly a revamp to the KHT may be possible to reduce/eliminate kerosene processing at the DHT.

In the Coker-Hydrocracker cases it may also be difficult to meet 25 mm smoke point in jet product since straight run kerosene smoke point is only 22 mm. In order to meet the smoke point KHT product needs to be treated in an aromatic saturation unit.

6) Diesel Sulfur

The Coker-FCC cases will violate maximum diesel sulfur specification (15 ppm).
The finished diesel is a blend of DHT plus Mild Hydrocracker diesel.
The DHT can be designed to produce 15 ppm product, but the mild hydrocracker diesel will be sufficiently high in sulfur that it cannot be blended to diesel successfully at the expected relative production rates.

It is not practical from a cost standpoint to produce 15 ppm sulfur diesel at the mild hydrocracker. It could be processed at the DHT with subsequent capital impact at that unit.



7) Diesel Cetane & Gravity

The Coker-FCC cases, as given, appear to be infeasible with respect to cetane and/or specific gravity. By experience, a standard Coker-FCC refinery is unable to produce 51 (Euro) cetane diesel due to the quantity of low cetane FCC distillate produced. Most U.S. Coker-FCC refineries can produce 41 cetane diesel. Many property details for the intermediate streams were not completely available, so only generalizations can be made. The straight run diesel by assay data appears to be fairly low as will be the Coker distillate. Upon upgrade across a typical low sulfur DHT, these will likely improve to a cetane in the mid to upper 40's. However, the difficulty is in dealing with the FCC and the mild hydrocracker distillate. Both will have very low cetane. For example, a typical FCC distillate cetane will be about 28 and a Mild Hydrocracker distillate will run less than 40 cetane. Even with upgrade in a typical DHT, the FCC distillate will be significantly below the pool specification and the current plan is to have the Mild Hydrocracker distillate bypass the DHT altogether. Very likely a high pressure DHT designed for cetane/gravity upgrade, in addition to desulfurization, will be required, or possibly a two stage unit designed for aromatics saturation. In any case, the plan should also include processing of the Mild Hydrocracker distillate through the DHT as well.

Even in the Coker-Hydrocracker cases DHT unit performance needs to be reviewed in detail. In the LP study, it was assumed that there will be about 10 number cetane index upgrade across the DHT unit. That may be possible only with very expensive DHT unit (high pressure unit followed by aromatic saturation unit). It was also assumed that coker diesel cetane index was 50 which is very optimistic.

8) Fuel Oil Sales

Most cases produce a fuel oil product (PFO) that is significantly below crude oil value. Fuel oil is also produced as needed to balance the refinery fuel system. Largely Coker gasoil is the main blendstock for these fuel oils. It is recommended that an evaluation be made to eliminate all fuel oil sales, route all coker gasoil to the respective conversion unit, and make refinery fuel oil from some combination of vacuum residue, unconverted oil from the respective conversion unit (HCU or FCC) and lowest value distillate as needed for cutterstock. It is almost never economic to intentionally produce low value fuel oil. Only the minimum required for refinery fuel should be produced and it should be produced from streams having no better distribution (HCU, and FCC unconverted oils), or from streams such as residues that are the most difficult and expensive to convert.

Incremental values for vacuum residue, heavy coker gasoil, and hydrocracker unconverted oil for Cases 5 through 10 are shown in the attached spreadsheet. Heavy coker gasoil has the highest value. Therefore it should be processed in the hydrocracker unit and converted to more valuable products.

9) Refinery Fuel

a) Some cases show very high fuel oil use as a percent of total refinery fuel. For example cases 8 and 10 show over 50% of total refinery fuel comes from fuel oil. Most high conversion U.S. refineries burn no fuel oil at all and some are just barely out of the flare with respect to fuel balance depending on the local situation. Even when burning fuel oil, only select processes can easily do so (Crude and Vacuum Distillation, Boiler House, etc.), so it may be impractical to plan on very high proportions in the overall fuel supply. Using oil for fuel at some processes such as a reformer is technically possible, but very expensive from a capital cost perspective. These considerations suggest a detailed review of the fuel balance is needed.

b) Case 10 shows only the Coker, HCU and DHT vent gases being routed to the fuel treater. The specific fuel system treating requirements were not made clear. However, often all sour vent gases are treated which in this case would include the CDU, NHT, and KHT. If done, this would have an impact on the fuel gas treater sizing and capex.

10) Coker

a) Yields are relatively invariant with feed quality - For example there are cases where the feed concarbon changes by almost 14 numbers with no change in coke yield. Such yields fundamentally mis-state the coker representation and the conversion feed material balance between cases.

Example LP Study Coker Results

	Case 6	Case 7	Case 8	Case 9	Case 10
Coker Feed ConC	19.3	24.5	23.0	34.3	24.2
Product Yields wt%FF					
H2S	1%	1%	1%	1%	1%
LPG	4%	4%	4%	4%	4%
Fuel Gas	4%	4%	4%	4%	4%
Diesel (& Naphtha)	42%	45%	45%	45%	45%
Gasoil	17%	14%	14%	14%	14%
Coke	32%	32%	32%	32%	32%
	100%	100%	100%	100%	100%

Coker yields used in the LP for Cases 1 through 10 are shown in the attached spreadsheets.

b) Unusual yield patterns - Gasoil yield is roughly 14 wt% of feed which is extremely low. Low gasoil yields can be produced by recycling gasoil to feed. Generally this increases distillate yield, but also converts some gasoil to coke and fuel and at the expense of a much larger, more capital intensive unit to handle the recycle. Usually for a transportation fuels refinery, a conventional, low recycle coker is most economic with respect to yield and capital expenditure.

11) FCC

Unusual yield pattern - No slurry oil (CSO) yield is indicated and the distillate yield is very high at ~29 wt%. LP results show FCC distillate SPG > 1.0 implying a CSO gravity. Possibly the FCC distillate yield shown is for combined distillate plus CSO together, but some cases route all FCC distillate to the DHT which would be impossible if CSO were included. If the given FCC distillate yield excludes CSO, then it is doubtful the given yields will carbon/hydrogen balance. These yields appear fundamentally unrealistic.

12) NHT Unit and Light and Heavy Naphtha Splitting

In the LP study it was assumed that NHT product would have 14 wt% light naphtha and 86 wt% heavy naphtha. According to crude oil assay data C5-85 °C light naphtha in total naphtha (C5-165 °C) varies between 20 to 32 wt%. Therefore in the study the light naphtha cut point is much lower than 85 °C (most probably around 70 °C). This means that all the benzene precursors would end up in the heavy naphtha. As a result, it would not possible to control the benzene content in the reformate and thus the refinery would not meet gasoline benzene specification. In order to remove the benzene precursors from treated heavy naphtha, light naphtha cut point should be increased over 100 °C

13) Reformer

a) Yields are basically invariant with feed quality. Some cases show a feed N+2A change of 40% with no effect on reformer yields which is unrealistic. This fails to account for crude changes, HCU naphtha influence, and affects plant H_2 balance and gasoline octane balance, etc.

Example LP Study Reformer Results

WT% Yields	Case 6	Case 7	Case 8	Case 9	Case 10
Feed	100	100	100	100	100
Products					
H2	2.8	***2.8***	***2.8***	2.8	2.8
Fuel	7.7	***7.7***	***7.7***	7.7	7.7
LPG	3.5	***3.5***	***3.5***	3.5	3.5
Reformate	86	***86***	***86***	86	86
	100	100	100	100	100

Feed Qualities	Case 6	Case 7	Case 8	Case 9	Case 10
SPG	0.807	0.765	0.763	0.764	0.750
N2A	60	***51***	***70***	60	58

Reformate Qualities	Case 6	Case 7	Case 8	Case 9	Case 10
SPG	0.82	***0.82***	***0.82***	0.82	0.82
RON	105.2	***105.2***	***105.2***	105.2	105.2
ARO	50	***50***	***50***	50	50

b) Gasoline blends indicate reformate aromatics content of 50% at high severity (RON=105), which is unrealistic. A more reasonable value is 75% which may make the Coker-HCU cases infeasible with respect to gasoline aromatics.

c) Certain study documents indicate a 98 to 100 RON nominal CCR design, yet LP model indicates 105 RON reformate was used. It appears that no variation in operating severity was allowed in the analysis. Normal reformer operation includes severity variation which can significantly impact case results.

14) Light Naphtha Isomerization unit

Isomerization unit hydrogen consumption was 1.5 wt% and isomerate RON was 80. Properly designed Isomerization unit will use much less hydrogen and the product will have higher octane.

15) DHT

a) Relatively constant H2 uptake regardless of significant feed quality changes. For example, between cases 9 and 10 the proportion of straight run to coker distillate changes significantly while H2 uptake remains essentially constant.

Example LP Study DHT Results

	Case 6	Case 7	Case 8	Case 9	Case 10
DHT H2 wt% FF	0.88	0.83	0.82	0.85	0.82
% Olefinic (Coker) Feed to DHT	28	37	41	55	25

b) In the LP study, coker naphtha and diesel were combined into a single product stream at the coker unit. The combined mixture was treated in the DHT unit. However, in the LP cases, the DHT unit naphtha yield is insensitive to the amount of coker material (including significant naphtha) in the DHT feed. This causes improper estimation of the relative amounts of gasoline and distillate products and would have blending affects that could impact the size/severity of the NHT, CCR reformer and Isomerization units.

Example LP Study DHT Results

	Case 6	Case 7	Case 8	Case 9	Case 10
DHT Naphtha Yield wt% FF	7.5	8.9	8.9	8.9	8.0
% Coker Material in DHT Feed (Naphtha + Distillate combined)	28	37	41	55	25

c) Product distillate qualities appear fixed regardless of changing operation and at quite optimistic values (~56 cetane) for a typical low sulfur DHT operation.

16) Mild HCU

a) Relatively constant H2 uptake regardless of significant feed quality changes. For example, between cases 3 and 5 the proportion of straight run to coker gasoil changes significantly while H2 uptake remains essentially constant as does the hydrotreated gasoil yield.

Example LP Study Mild HCU Results

	Case 3	Case 5
H2 Uptake wt% FF	1.54	1.54
% Coker Material in Feed	0.3	10.2

b) Erratic Loss yields – Loss yield varies between 0 to 1.37 wt% between some cases. Normally Loss would be associated with feed nitrogen conversion to ammonia. The loss yield must be above 0%, but the unit feed does not contain sufficient nitrogen to justify a result as high as 1.37%. Such a wide swing in results incorrectly affects the refinery margin and corrupts the case by case comparison. For example, the high Loss (1.37 wt%) cases are incorrectly disadvantaged with respect to refinery margin due to reduced product yield.

Example LP Study Mild HCU Results

	Case 1	Case 2	Case 3	Case 4	Case 5
Loss Yield wt% FF	1.36	0.00	1.37	1.35	0.00

c) Product distillate qualities are very unrealistic at ~56 cetane and 10 ppm sulfur. Such a unit would typically produce a distillate cetane well below 40 at a sulfur level well above 15 ppm. This stream will require processing through the DHT in order to blend into finished low sulfur diesel.

17) Hydrocracker

a) Relatively constant H2 uptake and product slate regardless of significant feed quality changes.

Example LP Study HCU Results

	Case 6	Case 7	Case 8	Case 9	Case 10
H2 Uptake wt% FF	2.80	2.80	2.80	2.80	2.80
Feed API	22	15	20	17	17
Feed Sulfur	1.30	1.26	0.80	1.93	0.95

b) No Loss yield indicated - Loss yield should result and vary mainly with feed nitrogen.

Example LP Study HCU Results

Product Yields wt% FF	**Case 6**	**Case 7**	**Case 8**	**Case 9**	**Case 10**
H2S	1.31%	1.27%	0.81%	1.95%	0.96%
Fuel	2.72%	1.24%	1.24%	1.23%	1.24%
LPG	2.47%	2.47%	2.48%	2.45%	2.48%
L Naphtha	4.44%	4.94%	4.96%	4.91%	4.95%
H Naphtha	14.82%	12.84%	12.90%	12.76%	12.88%
Jet	19.75%	24.70%	24.81%	24.54%	24.77%
Diesel	55.36%	53.40%	53.64%	53.06%	53.56%
Gasoil	1.98%	1.98%	1.98%	1.96%	1.98%
LOSS	**0.00%**	**0.00%**	**0.00%**	**0.00%**	**0.00%**
Feed Nitrogen, ppm	2427	1726	2557	2908	1504

c) Unusual product qualities - Example Case 10 HCU distillate indicated at ~53 cetane versus DHT distillate cetane for the same case is ~56.

18) H2 General

a) H2 uptakes are generally constant regardless of significant feed quality changes. See comments by unit.

b) Handling of solution loss is unclear.

c) Significant H2 imbalances appear and vary by case. Some cases are H2 balanced such as Case 10. However, in some cases such as Case 5, H2 production exceeds consumption by ~27%, with a large excess of H2 to fuel. H2 Plant operation should run to balance H2 needs. Such variation indicates an issue with the case setup. Results are unreliable since refinery margin and H2 plant sizing is incorrect.

Example LP Study H2 Balance Results

Case 5 H2 Balance

		NHT	0.89
		Isom	0.93
		KHT	0.18
Reformer	10.63	DHT	11.22
H2 Plant	26.33	CFH	14.03
Total produced	**36.95**	**Total Consumed**	**27.25**

~27% Imbalance , H2 to Fuel, H2 Plant over-sized

d) Invariant reformer yields as already discussed will impact case results.

19) H2 Plant

Loss yield appears understated and produced fuel yield appears overstated. H2 Plant Loss is usually the yield of water and CO2 since these have no value as fuel. Usually Loss and produced fuel yields are roughly similar magnitudes. However, in the current study the produced fuel yield is nearly two and half times the Loss yield. The H2 Plant produced fuel appears to be handled the same as other refinery fuel gases. That is, it contributes into the refinery fuel system apparently on an equal heating value basis which implies it is on a water and CO2 free basis. This arrangement will overstate the H2 plant fuel yield and incorrectly represent the refinery material balance.

20) Capital Expenditures (Capex)

Capex estimations were made by China Huanqiu Contracting & Engineering Corp. (CHCEC) based on Costa Rica, using their indexes. In the other hand, UOP Capex estimates are made based on US Gulf Coast, which is the standard for this type of estimates. Although bases are not the same and the comparison would be altered, some qualitative comments can be made related to the investment numbers and its magnitude. This comparison was executed only for case 10 configuration.

In general, investment figures seem to be high, presenting the Reformer, Diesel Hydrofining Unit and the Delayed Coker the highest differences. CHCEC total investment is ~20% bigger when compared to UOP.

Suspect Capital estimate is found for almost all the Process Units. In the majority of the cases, values indicate a simple linear relationship relative to feed rate, instead of the Standard scaling practice of the capex based on capacity ratio raised to an exponent, usually 0.6 to 0.7. [$Capex_2 = (Capacity_2 / Capacity_1)^{(0.6 - 0.7)} \times Capex_1$]

Specific comments to process units are:

a) Reformer:
The same linear Capital relation appears to be used for the Coker-FCC and Coker-HCU configurations. Very likely the optimum reformer feed quality and severity combination would be quite different between these cases with subsequent capex impact.

b) DHT:
As in the Reformer case, the same linear capex relation appears to be used for both the Coker-FCC and Coker-HCU configurations. Very likely the optimum DHT design severity will change significantly between a cetane/SPG constrained Coker-FCC configuration versus the Coker-HCU configuration which will be much less constrained. Based on the diesel quality, the DHT capacity/capex in the Coker-FCC configuration should include processing of the mild hydrocracker distillate.

c) Coker Unit:
Coker unit capital is a strong function of both feed rate and con carbon which is directly proportional to coke rate. Coke rate affects the drum size and number which is a significant cost. In the study analysis to the coke rate effect on capital appears to be ignored.