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SORESCO

COSTA RICA MOIN Refinery Expansion & Modernization Project

L.P. Enhanced Validation for FSR Case Ten – Phase II

Summary

This report lists findings of WorleyParsons process validation work as of 03/26/2012 for the Soresco Moin refinery brownfield and greenfield studies.

In Phase II of this work, we utilize conversion unit yields from licensor quotes to modify the estimated configuration. There are no real significant changes from the Phase I efforts and nothing at all to change our contention that the choice of a delayed coker with hydrocracker a is the best selection for a "diesel fuel" oriented production scheme.

There were some issues with the potential design specs and initially reported yields for some of the conversion units, especially for Chinese licensors. These issues have been resolved and agreed to by all parties after a weeks' worth of collaboration meetings in Monrovia.

There is still one remaining issue - hydrogen consumption in the HCU (Hydrocracker) and DHT (Distillate Hydrotreater).

Table 1 shows the Phase I estimates for and the Phase II validated refinery unit capacities. The feasible under/overcapacity ratios are 60 /110 % for all units but the DHT which has an under/overcapacity ratio of 60 /120 % per Soresco management request.

WPM concurs with this decision. The ability of the DHT to absorb unit feed swings based on normal feed rate and quality changes is a critical factor in the success of the Soresco project. It is so critical for heavy ends processing that one technology licensor, Chevron, wants to combine the DHT and HCU units because it recognizes the synergy between the two for this plant.

Unit	CASE		Phase I 10A	Phase II 10A
ADU 1	**Feed Crude**	**Name**	**Pennington**	**Pennington**
		API	33.5	33.5
		Rate (BPSD)	25000	25000
		Nominal Rate (kta)	1200	1200
ADU 2	Feed Crude	Name	Vasconia	Leona
		API	27.3	21.5
		Rate (BPSD)	40000	40000
		Nominal Rate (kta)	2000	2000
VDU2	Feed Capacity	BPSD	27900	30200
DHT	Feed Capacity	BPSD	21700	28800
HCU	Feed Capacity	BPSD	22200	24800
DELAYED COKER	Feed Capacity	BPSD	12400	10500
KHT	Feed Capacity	BPSD	3600	2700
NHT	Feed Capacity	BPSD	12000	14600
C5/C6 ISOM	Feed Capacity	BPSD	6600	5100
CCR	Feed Capacity	BPSD	14200	15400
COKE MAKE	Estimated Capacity.	ST/D	511	526
SULFUR MAKE	Estimated Capacity	LT/D	16.3	18.5 x 2units

Table 1 – Phase I & II capacity case comparisons

1. OVERALL CONFIGURATION MODELING ISSUES

New, modified, or extremely important assumptions in our configuration validation models since Phase II are newly presented or repeated as follows:

A. **ADU 1 yields.** We have modified the ADU#1 yields to match our brownfield Hysys v 7.3 model results.

B. **ADU 1 AGO draw.** We have validated the viability of using the ADU#1 AGO stripper to recover AGO from the atmospheric column. We chose to keep this function in ADU#! for the following reasons.

   a. The equipment and piping connections from the column to the AGO product pump exist and are seemingly adequate for the use we propose to put them to. It costs very little to have this option as opposed to a brand new unit.
   b. Taking AGO from the crude unit takes a capacity and duty load off the downstream vacuum unit, in this case, VDU-2
   c. Our preferred disposition of the AGO material is as HCU feed. We can send it hot from the product pumps, or we can selectively use the stream to maintain preheat train temperatures before sending it across unit boundaries at a reduced temperature. The amount of AGO is less than 1000 BPD and whether or not it comes at 650F or 400 F should make little difference to the HCU operational capacity.
   d. A fraction (0-70%) of this AGO, while nominally like a VGO material, could be sent to the DHT and end up yielding more percentage of diesel material than what hydrocracking of that stock would yield. The total amount of the ADU#1 AGO is less than 5% of the other DHT feed stock rates and given the cut point decrease of 10-15C expected in the DHT, would allow this material to blend in to the ULSD pool without messing end point and gravity characteristics of the fuel. This will be an operations optimization opportunity by Soresco but our refinery blending experience give us reason to expect some of this potential revenue enhancement to be available after startup

C. **ADU 2 AGO draw.** We are neutral as to whether or not WPC uses an AGO draw in the atmospheric column. **It is a matter for FEED optimization.** If an atmospheric AGO stream is not taken, it is possible a small amount of diesel can be recovered from the vacuum tower overhead and/or LVGO pump-around stream. We call it out and estimate it for the possibility it could be available as a DHT feed. Again, using this draw strategy is a matter for FEED optimization to determine if the benefits are worth the additional unit capital cost.

D. **VDU-2** flash zone is estimated at approximately 700 F at 25-35 mm pressure. We assume a 986F HVGO cut point as we feel the metals and asphaltenes content of this swing to 1050+ cut point material is too high for proper hydrocracker catalyst performance, which is where this material is mapped to for normal processing. If the HCU licensor concurs, this swing material will add to the hydrocracker feed at the expense of the coker feed. It should be noted that swing gasoil metals and asphaltene content for one crude may be OK and that of another crude may not. The hydrocracker licensor needs to be aware of all potential crude gasoils that will be fed to the unit.

   Taking the 986-1050F VGO swing cut will require a 710-720 F flash zone at 10-15 mm Hg pressure. It will also add to the ultimate amount of hydrogen needed for both the Hydrocracker to process the heavier amount of heavy gas oil feed.

E. **KHT feed capacity.** Case10A requires the KHT to handle 3600 BPD of potential jet fuel. Preliminary analysis of the hydrotreater shows that this is within the capacity of the unit. We have not finished and accepted revamp proposal to raise the capacity, though we believe we can get to 5000 + BPD capacity. We need more time to engineer this solution.

   Meanwhile, all ADU#2 heart cut kero will go to the DHT for eventual inclusion into diesel product, currently a more profitable operating mode. ADU#2 swing cut N/K material will go to the NHT and eventually to the CCR.

F. **All Heavy SR naphtha, swing naphtha, and Heavy Hydrocrackate (HHC) go to the new CCR**. In all but maximum jet make, we can't allow too much of the NK swing material to blend into jet because of 10% distillation limits. This material is low octane and represents a substantial downgrade loss if blended to gasoline or sold independently. The HHC RON can be as high as 80-82 but we are now thinking it probably should be a CCR feed, even if it is this high and is not as good a feed as straight run for RON increase. Soresco needs all the octane-bbl. it can get. Licensor clarification will probably confirm this concern. Sometimes the licensors can blend catalyst or condition the catalyst to provide a bit of naphtha isomerization. Even if the HHC gets a lower RON base-delta boost, it would definitely help the overall octane balance without loading up on a high severity, high-aromatics content reformate. The cut point of the HHC depends upon what the CCR vendor will allow into their reactor. We would like to cut it at around 200 C but may have to keep it lower with the tails going to HCK. We have experience in reforming such material to get more hydrogen make as a valuable by-product of that unit. as much as for the octane boost. The biggest threat to utilizing the stock is an increase in the required regeneration rate - the R part of the CCR. Coke laydown on the CCR catalyst is the primary problem here.

G. **All straight run Heavy Naphtha & Coker Naphtha will go to combined new NHT,** in accordance with item F above, which may be integrated with the CCR operations. This assumes the olefins and di-olefins make in the latter feed is low enough to meet gasoline specs. If not, a selective hydrogenation unit may be needed for saturation of the double bond items. We feel you will be OK as this naphtha in native mode is no more than 5 % of the total naphtha pool. We assume that the DHT naphtha will first have to be treated in the DHT, then go to the NHT for ultimate splitting into light and medium fractions.

H. **Hydrogen Balance**. We have redone our initial simple hydrogen balance in terms of supply needs to each conversion reactor based on licensor needs. FEED project WORK will determine the detailed balance as we do not get this sort of info from the LP other than what we might guess as an input. We have assumed the CCR H2 is a typical 83-85mol% purity. It could be as low as 75%. The licensor will have to verity this information... We have left a high-conservative estimate of pure hydrogen makeup needs regardless of whether it is delivered from a CCR or a PSA (typ. 99+% purity). Again, the final configuration here will be the responsibility of the FEED project work. We cover more discussion of this topic in Section 2.0 of this report

I. **Sulfur Balance.** The initial sulfur plant balance in Phase I assumed that 80% of the feed crude sulfur will come out in the hydrotreater and hydrocracker vent gases since it is not really allowed to come out in the liquid products. We have changed this estimate to 85% just to up the SRU plant design basis to cover all reasonable potential scenarios. Typically, 80% is a more common recovery factor, but we wanted to be conservative.

J. **All Light SR naphtha goes to the new Isom unit.** The cut point for this material is 85C for both LN1 and LN2. This keeps most of the benzene and precursors away from the CCR. We are not certain but that a pre-feed BTX column in the CCR unit may be needed. The Isom catalyst vendor will indicate if this is necessary to keep aromatics away from its catalyst. Most tolerate 5-8% benzene in the isom reactor feed. Light Hydrocrackate and hydrotreated DCU Naphtha are also traditional gasoline blend stocks and could benefit as ISOM feedstocks. There is some marginal value to doing so and as stated above, as many low-aromatics octane-barrels as can be made are necessary for keeping the gasoline blender operations as flexible as possible. We also identify HCU debutanizer bottoms as a source of significant c5+ material from the HCU separator system – material that doesn't come off the main HCU fractionator. There could be several sets of Debutanizer columns in the plant-one for each hydrotreater as well. Each will have a recoverable, if small, amount of light naphtha of very low sulfur and aromatics content. Collection of such material is a FEED project optimization problem but it should be considered as a source of ISOM feed.

K. **Isom Unit Definition**. It is possible that the CCR&Isom licensor(s) will want some control on the NHT unit, probably needing very low sulfur on the CCR feed. Some newer cat reforming catalysts can hold up to 10-20 ppm sulfur but most CCR/reformers still want <1ppm. If the NHT is not linked to the CCR, then there will likely be an added pre-CCR hydrotreater as a part of the cat reformer unit package. Typical ISOM units have a sulfur guard bed followed by a DeCyclohexanizer unit to keep cyclic C6's an lighter going to the isom unit instead of the CCR with the heavy stream from this column. The DCH column would typically function as a naphtha splitter and not require one in the NHT. Please note that the ISOM unit may also have an aromatics saturation unit as well to handle benzene if the final reformate has excessive benzene or total aromatics make. The Isom licensor will decide what parts of the system are required for the Moin Refinery. If Coker naphtha components are present in the Isom feed, a catalytic diene removal (selective hydrogenation of diolefins) may be needed as well.

L. **All Straight Run Diesel, VDU & Coker LGO go to the new DHT to make ULSD stocks.** About 20 F is cut off the end point of the composite feed material (so you can handle a bit more Coker LGO) and you will crack a small but significant amount, enough to generate some naphtha material. The reactor hydrogen partial pressure will likely be over 1200 psig and the catalyst needs to be able to saturate olefins as well as sulfur and nitrogen. It also will have to do a bit of light chain aromatics saturation to make cetane and smoke point specs. Thus it is a much more severe hydrotreater than either the NHT or the KHT. We modified the Phase I model to more closely agree with CEI DHT yield estimates.

M. **All SR LVGO and HVGO go to the new Hydrocracker along with Coker HGO.** . We originally picked the DCU model cases to maximize liquid products. Item D above discusses the cut point issues for the straight run cuts. Depending upon crude feed source and the operations of the coker itself (temp and soak time), the yield of heavy Coker GO may be as high as 30% of whole feedstock with potential properties of 1.1~1.6wt% CCR, 600~640℃ FBP and 400~800wppm asphalt. If this is the case, it may be necessary to use more recycle and restrict the cut point of the HCGO draw. There would thus be more coke make and gas make with less fuel make. It is a consideration to take up with the licensor before getting too far into fixing the FEED stage design. We modified the Phase I model to more closely agree with CEI HCU and DCU yield estimates from their licensor contacts.

N. **Coker model yields.** We initially assumed a 1.05 Recycle ratio to maximize liquid yields and estimate the max feed rate to the HCU. We believed the coke price we were instructed to use is too high to be sustained and that liquids to the HCU will give better return. Licensor information, however, is the key modeling priority for our work here. Therefore, we modified the Phase I model to more closely agree with CEI DCU yield estimates from their licensor contacts. They are reasonable since all mid and heavy distillates will be processed in either the DHT or HCU, ultimately making mostly diesel product anyway. It takes a great change in the light to heavy DCU gasoil product ratio to greatly affect the amount of high value product for sale in the end.

O. **UPR & URG blends.** We changed little in our use of oxygenates, estimating both MTBE and ethanol blend characteristics. Soresco management backed off on the 35% aromatics content in the gasoline after Phase I work showed that this level could not effectively be made with MTBE and only approached by using Ethanol. Both octane grades and oxygenate blends now must meet only the old 40% aromatics standard.

We attempted to maximize the amount of premium grade sales product that would be made from the refinery conversions and assumed that Soresco would buy the appropriate stocks on the market to meet regular gasoline demand. We did not include this added purchase program in the overall refinery economics. Soresco would be wise to look at this buy/make blending situation as another We did however, identify an opportunity to reduce octane giveaway (and trim aromatics a bit) by blending a small amount of low octane naphtha into both UPR and URG to close the octane giveaway from 0.3-0.5 RON down to 0.05-0.2 RON

With the choice of the DCU-HCU configuration, Soresco creates a bit of a problem for maximizing gasoline octane while keeping aromatics within the proposed 35% limits. Do you set the reformer to a single "one size fits all" severity and blend it with Isom and with non-enhanced low octane cuts such as HHC. If so, you will likely not be able to meet both octane minimum and aromatics maximum limits for your premium gasoline. Or you will limit your premium make to whatever Isom you can balance the blend with and hope your residual stocks can blend with the high severity reformate in at least a 2:1 ratio. An alternative option is to blend the low value stocks into the straight run CCR feed and run a lower overall RON product, say 87-90, but get more octane-bbl. overall into the gasoline pool.

Soresco can make all the URG product it would want to both octane and aromatics specs even with MTBE. In order to make premium, however, some amount of higher-severity (100-102) reformate is needed to blend with 89-90 octane ISOM in a roughly a 1:2 ratio to keep aromatics in spec – that is, below 35%. This means, also, in order to sell UPR, the refinery must not use all of its ISOM making URG. This is the optimization problem one faces. Ethanol usage (10%) helps the aromatics levels (and octane) but in all cases, it is going to be difficult to make reformate blend aromatics limits with a single severity operation.

Blend ratios for maximizing 95 RON production (70-80% of total gasoline pool) show that the aromatics content can just make the current 40% limit, but are far off from the proposed 35% limit. The ethanol blend scheme gets the aromatics level down to 36% which might indicate possible compliance within the error of the modeling data, but again, a solution at this ration here would just barely hit within the margins. Even with dual severity modes, Soresco will have a more difficult time making as much UPR with 4% MTBE as the oxygenate.

P. No regular production of LRG, AVG, HSD, heating oil or Fuel oils as normal significant sales. These are one-off or irregular types of batch processing projects and not considered for normal processing configuration analysis. This assumption has not changed.

Q. Overall Gasoline Conversion Strategy. The refinery also needs every octane-bbl. it can make that is low in aromatics. For the proposed Moin refinery configuration a C5/C6 Isomerization unit is correctly used for this function. Initial estimates by HQCEC assumed an 84-85 RON production with straight-run naphtha feed. Worleyparsons believes that a quality licensor unit with recycle will allow production of 89-90 RON blend stock with typically 5% or less aromatics. Maximizing the amount of this stream allows use of more high-severity CCR product. Octane-bbl. production can be enhanced by upgrading coker naphtha and light hydrocrackate from low to high 80's values. . The implications for the Isom Unit are discussed later in this report.

R. **Overall MidDistillate Conversion Strategy.** Current market conditions would drive the LP to maximize USD above Jet and Jet above gasoline, subject to product specs, of course. Optimal design strategy, however, would tend to ensure the ability of the refinery to meet domestic consumption and be able to sell more Jet if the market dictates. We have artificially (though perfectly maintaining product specs) about 3000 BPD of kero into diesel that would have otherwise gone to jet fuel. We have attempted to provide a design philosophy that would all other potential planning opportunites for Soresco to "upgrade" blend-stocks by mere operational decisions.

S. **ADU#1 Satgas vs. ADU#2 Sats/Unsat gas plant.** We initially recommended two gas plants, an existing revamped ADU1 and a new unsats gas plant needed for both net Greenfield unit light ends and coker olefins processing. In reality, the utilization of existing ADU1 equipment beyond the atmospheric tower is an issue needed for FEED activities in both brownfield and greenfield units. Our initial Brownfield FEED analysis in this area shows that we could handle only about half of the ADU#2 naphtha make in our debutanizer and naphtha splitters, both of which may need a reboiler and tower tray/packing revamp to do so. The same is true for the LPG/light ends material fed to the absorber/stripper. All of the ADU#2 LPG/LE material would overwhelm the Gascon #1 capacity. In addition, WPM believe that sending hydrogen rich vent streams to the Absorber may not be the best way to handle effective economic hydrogen recovery. For these reasons, we would suggest Soresco use Gascon#1 in a capacity offload scheme with Gascon#2 designed for the full load of the Greenfield vent and naphtha makes.

T. **Hydrogen Unit Design Assumptions.** : With more defined vendor liquid product yields and hydrogen makeup requirements, we are able to form an initial hydrogen balance for the Moin refinery model.

CEI presented a good idea for recovery of 99.9% Hydrogen from the CCR Separator off gas with a PSA unit. Combining this recovery with a combination of a steam-reforming plant and a 99.9% H2 from PSA unit and feeding all hydrogen needs from a PSA hydrogen header would seem to be an economical and operationally beneficial scheme which we endorse.

Since the Moin refinery has no outside natural gas network to feed from, it must generate all of its own hydrocarbon fuel or buy fuel oil to make up the difference in fuel duty availability. Thus the refinery RFG and LPG material makes necessarily form the feed stock to the hydrogen reformer. We have estimated amounts of hydrogen, water, and CO2 resulting from either fuel gas or LPG usage in the model.

U. Assumed Refinery Operating Costs for the LP. Soresco has indicated that the typical refinery operations cost they are responsible for is about $ 5 /bbl. RECOPE corporate costs are not covered by this number.

KBC analysis of year 2016 and beyond crude-product prices indicate that the refinery may be expected at about $17-19/bbl under the circumstances predicted by the KBC study. The refinery is expected to generate about 80% of its fuel needs for fired heaters and import 30MW every operating day. In addition, we assume that about 500 BPD of fuel oil per day is imported to make up the 20% shortfall. An alternative option was to use the coker naphtha as supplemental fuel gas in order to avoid any need to import oil. This option actually ends up costing $21/net MM BTU (basis; lost LPG sales revenue) which is nearly $4.5 higher than for simply using fuel oil. We chose the former fuel scheme.

The final values show for operating expense fixed and variable operations expenses are shown in Table 2 below.

	Cost (2012)		Consumption		
Variable cost	Value	Unit	Value	Unit	% COST
Imported fuel oil***	500	BSPD	175.000	bbl/Y	14%
Fuel Gas***	0,00	US$/MMBTU	1.024.800	MMBTU/Y	0%
Electricity***	251.244.000	kWyear	26.631.864	US$/Y	24%
Catalysts & Chemicals**	3.067.500	US$/Y	3.067.500	US$/Y	3%
Make-up water from river**	0,00012	US$/m³	2.418.994	m3/Y	0%
Potable water**	0,5	US$/m³	54.600	m3/Y	0%
Sewage Charge**	197	US$/Y	197	US$/Y	0%
Total Variable Cost					40%
Fixed cost					
Employee expenses****	20	U$/man-hour	69.333	man-hour/month	15%
Maintenance **	3,50%	CAPEX	46.350.675	US$/YEAR	41%
Insurance & Local taxes**	0,15	US$/bbl	10062	US$/YEAR	3%
Total Fixed Cost					60%
Total Cost per year					100%
		ENERGY	Maintenance**	Employee expenses	
Total Cost per barrel with maintenance =	$5,25	$1,98	$2,18	$0,78	

RECOPE INFORMATION****			** maintenance global factor			
Employees:	INVESTMENT**	UNIT	FUEL	CONSUMPTION		
400	1.324.305.000	US $	FUEL OIL	500	BSPD	
400	67.080	bbl/day	FUEL GAS	1024800	MMBTU	

**FSR INFORMATION
***ESTIMATION WPM
****RECOPE INFO.

Table 2 – Fixed and Variable Operating Costs Predicted for the Moin Refinery

V. Fuel Balances. Enough information was gleaned in the last month to allow WPM to give reasonable estimates of energy duties, fuel make yields and power consumptions needed.

Once the reformer/H2 Unit needs were tightened, it was possible to determine the fuel makes available to run the refinery as well as the need for supplementary fuel oil purchases.

The fuel LHV supply was calculated in the model as:

FUEL GAS MAKES

Stream	Description.	Comments	LHV BTU/SCF		KTA	MW
GG1	GCU #1 Clean Fuel Gas to H2U	12.6 MOL % h2	lhv=	1,109.0	0.0	20.3
CH4	VENT GAS TO FUEL	5.6 WT% h2	lhv=	1060	26.5	20.3
GG2	Clean Fuel Gas to Fuel Gas header	10MOL % purity	lhv=	1,109.0	41.6	30.5
CHD	H2U PSA VENT GAS TO FUEL	9.3 WT% h2	lhv=	265.0	44.0	20.3
						3.2

RFG	**Refinery Fuel Gas-High BTU**	Comb. Rate	To Fuel Gas	kta=	68.1	Note: CO2 removed from H2U PSA vent gas not included
mw	lhv - btu/scf	194.56	MT/D			
24.4	1098	17866.76	LB/Hr			
		732.85	MOLS/HR			
		263384.53	scf/hr			
		289.2	**MMBTU/HR**	available		
RFG	**H2U PSA VENT GAS – Low Btu**	Rate	To Fuel Gas	kta=	44.0	Low BTU fuel gas
mw	lhv - btu/scf	125.80	MT/D			
3.3	264	11552.80	LB/Hr			
		3500.85	MOLS/HR			
		1258204.54	scf/hr			
		332.2	**MMBTU/HR**	available		

LPG	**COKER NAPHTHA TO FUEL**			
	Available Coker LPG	23.80	kta	
	LHV	19600	BTU/lb	
		67.98680914	Mt/d	
		149842.9274	lb/d	
		122.371724	**MM BTU/hr**	**potential duty**

Total RFG (Hi&Low BTU) = 734 **MM BTU/hr Potential duty**

The total fired duty needs of the refinery These were taken from licensor data, some rigorous crude simulation models, and fill in with FSR data shows that the maximum fired duty is between 670 and 750 MM BTU/hr

It is possible that the Moin refinery can supply nearly all of its fired duty by fuel gas if it is willing to utilize the low BTU gas derived from the H2U PSA vent gas. This is predicated upon a 50.5 kta net PSA hydrogen stream taken from a steam reformer of high conversion and with relatively efficient CO_2 removal – presumably from amine. If the reformer CO_2 is substantially removed, the low BTU-gas will have an LHV of about 265 BTU/SCF. If the CO_2 is not removed, the LHV will be about 175 BTU/SCF but the total volumetric make will increase by 50%. The net available energy will not significantly decrease but the gas volume increase for the same duty delivery will result in larger delivery headers, more complicated valving and controls, and more expensive custom designed burners with less design flexibility.

This scenario also requires the use of coker-derived LPG as fuel. which can provide approximately 120 MM BTU/hr or about 15-20% of the plant needs. While there are some good reasons to do this, the option actually ends up costing $21/net MM BTU (basis; lost LPG sales revenue) which is nearly $4.5 higher than for simply using fuel oil. We chose the former fuel scheme. If the coker LPG could be used in place of fuel gas for the hydrogen reformer, instead, the economics might go against fuel oil usage

In another scenario, , consolidated blended fuel gas (Hi and Low mixed) would have a low heating value of about 820 BTU/SCF, which is a workable level, especially if supplemented with some vaporized LPG to get the rate up to natural gas levels at 900+BTU/SCF.

These sorts of final configuration decisions are properly a part of FEED optimization for the Greenfield units.

Fuel Oil usage and Power Needs

The FSR shows total electrical power needs of 30 MW which we believe is a reasonable value for a limited steam turbine utilization and only needing a 50 kta steam boiler. At a 40% generation efficiency at the nearby Moin Power plant, this translates into a total fuel oil use of 45 BPH or 1080 BPD.

1 bpd of fuel oil/ MM BTU fired duty will be needed to make up refinery heating needs that can't be met by one or another fuel gas. Fuel Oil equivalent is roughly 1 bbl per 6 MM BTU of fired duty.

W. Product Makes After most of the licensor information has been incorporated into the validation model, the final Case10A feed purchases and product yields are:

Feedstock Purchases	Units	DAY	YEAR
Pennington	BSPD	25.000	8.750.000
Vasconia	BSPD	40.000	14.000.000
Ethanol	BSPD	2080	728.000
Total Purchases	BSPD	67.080	23.478.000
Product Sales	**Units**	**DAY**	**YEAR**
LPG	BSPD	749	262,150
Unleaded Premium	BSPD	14.480	5.068.000
Unleaded Regular	BSPD	5.600	1.960.000
Kero/Jet	BSPD	9.070	3.174.500
Diesel	BSPD	35,490	12.425.500
Coke	MTONS*	526	184.100
Sulfur	MTONS*	53	18.515
Total liquid	BSPD	65389	22.886.150

Table 3 – Feed Purchase and Product Makes for Case 10A

2.0 HYDROGEN CONSUMPTION FOR HIGH SEVERITY HYDROPROCESSING

WPM's original evaluation of the hydrogen consumption needs of the Greenfield HCU and DHT units were derived from the perceived need to provide the following makeup rates

Unit	Description	Design rate (SCF H2/BBL Fresh Feed)
DHT	Distillate Hydrotreater – Processes SR Diesel range and Coker Midbbl or light Gas oil material into Euro V type ULSD. Cetane >52, S<20 ppm, Aromatics <20 vol% This is a very severe service with considerable aromatics saturation.	700 SCF/BBL
HCU	Hvy Gasoil Hydrocracker – Processes 650-1000 F end point SR VGO and Coker HGO. 98+% conversion to diesel, kero, and naphtha blend stocks. Must meet Euro V+ ULSD needs Cetane >60, S<20 ppm, Aromatics <20 vol% This is a very severe service with considerable aromatics Saturation and paraffin/naphthene chain cracking	2500 SCF/BBL

Upon further evaluation of typical licensor data for various types of severe LGO hydrotreating and VGO Hydrocracking, WPM proposes to use the values above as expected consumption (chemical and other basis) and to change the design basis to 1000 and 2800 SCF/Bbl of Fresh Feed respectively.

Case10AValidationCasesrEV9.xlsx was used to calculate the effect of this increase on the design capacity of the hydrogen/PSA unit. CCR vent gas hydrogen recovery is still retained as per CEI's suggestion. Even so, a total of 56.5 kta of 99.9% pure hydrogen is needed from new generation in a steam-HC reformer/PSA unit. WPM would suggest that two (2) 33 Nm^3/h units be installed, each with a nominal capacity of 30 kta of pure hydrogen. CEI/WPC has claimed that only one such unit will be required in addition to the CCR recovery scheme PSA.

The reason for our higher hydrogen consumptions is explained below.

INITIAL CONSUMPTION VALUATION

Typical consumption values for various projects using UOP catalyst indicate that the 700 and 2500 values represent reasonable values to use for our type of units, but that the final value should be based on a kinetic and thermo analysis of what reactions will actually take place. For example, almost total sulfur removal is easy to calculate. Similar levels of nitrogen and oxygen removal are also easy to find, although the kinetic removal of nitrogen is accomplished with more difficulty than sulfur removal. If the catalyst type and activity does not convert embedded nitrogen with any significance, the H2 consumption could be a bit lower but the resulting retention of nitro compounds in the products would not be very desirable. This retention can also be a by-product of hydrogen starvation.

The most critical component of the hydrogen consumption for the DHT will be aromatics saturation. For hydrocracking it will be aromatic saturation and also considerable paraffinic and naphthenic chain cracking. These reactions consume several times as much hydrogen per reaction site than simple HDS and HDS reactions.

Three Unocal pilot plant tests (NesteOy, Petronas, and Scanreff proposals' info) tallies up the **chemical** hydrogen consumption for a series of similar (to Moin feeds) heavy gasoil hydrocrackers and light gasoil hydrotreaters that were required to meet ULSD-like conditions for the most part. In some cases, the final aromatics content was considerably above 20 vol % and in others, it was not. All were required to remove S to 50 ppm, or essentially all the sulfur starting from 1-2 wt% in the feed. Although this data was derived from rigorous pilot plant data 20+ years ago, the chemistry is still valid. Catalyst activity affects temperature, H2 partial pressure, and space velocity needs, but the chemical consumption is still essentially the same.

In addition, the Unocal IROM design manual extract shows that the true hydrogen consumption, as opposed to pure chemical consumption, can be any-where from 10-40% above the former usage rate. This is our concern that we are low at least on design capacity of hydrogen.

We have recently validated this information informally with UOP and Haldor-Topsoe staff that WPM engineers have personally worked with in the past. Three WPM engineers have considerable HCU refinery operations and R&D licensing sales of hydrocrackers and hydrotreaters.

One final reference point is on a recent project in Saudi Arabia. The hydrogen consumption for a full conversion heavy gasoil HCU was set at 2800 SCF/BBL. We believe that this ought to govern at least the design of the hydrogen generation capacity. Using only the total 48 kta (max 18 from CCR gas) of hydrogen proposed by CEI will definitely be detrimental to the processing of any heavier, more sour crudes in the Moin refinery.

Reference files mentioned above(and others) for the basis of WPM DHT & HCU hydrogen consumption calues will be found on the Soresco Sharepoint site along with the proper validation model Rev9, a BFD and this report.

3.0 USE OF HEAVIER FEED CRUDES TO ADU#2

In Phase 1 an analysis was performed on use of Leona 22 and Roncador 18 gravity crudes instead of Vasconia. As set of yields based on earlier versions of the model, predicted capacities of the Refinery conversion and processing units. The percentage increase/decrease of capacities would still be expected to hold reasonably well for the capacities of Phase II presented in Table 1. We have put the Phase II capacities in Table 2 along with the relative capacity percentages for the two heavier crude cases.

Both the Leona and Roncador crudes are considerably heavier than the Vasconia crude. Their yields get shifted to the heavier products accordingly. The Leona crude has a high yield of diesel and light gas oil material, so the diesel yield at an overall refinery rate of 65000 BPSD Is highest. In contrast, Roncador crude has a high heavy gas oil fraction and resid fraction with a high CCR and asphaltene production. The resulting coker feed rate is highest of the three crude cases with a significant net hydrocracker feed as well.

The results in Table 2 can be interpreted in a couple of ways. If the new refinery design strictly matches Case 10A design capacities, then the other two crude case runs will need to have feed rate cuts to the level that VDU, DHT, DCU and HCU rates are within about 10% (the planned design rate contingency) above the Case 10A capacities shown.

Alternately, with some preplanning, use of the heavier crudes could allow Soresco to make some known but limited products. For example, use of the Roncador feed could allow a diversion of some VDU/DCU feed to the new Asphalt Column, VDU1. Use of the Leona crude may present an opportunity to divert some high sulfur diesel blend stock to storage, reducing the normally large DHT feed by enough capacity to stay within tentative Case 10A capacities.

Unit	CASE		10A(New)	10B	10C
			VOL BASIS	% BASIS	% BASIS
ADU 1	**Feed Crude**	**Name**	**Pennington**	**Pennington**	**Pennington**
		API	27.3	27.3	27.3
		Rate (BPSD)	25000	100	100
ADU 2	Feed Crude	Name	Vasconia	Leona	Roncador
		API	25.3	21.5	17.94
		Rate (BPSD)	40000	100	100
VDU2	Feed Capacity	BPSD	28000	108.24	128.67
DHT	Feed Capacity	BPSD	28000	132.72	84.33
HCU	Feed Capacity	BPSD	21350	111.71	106.31
DELAYED COKER	Feed Capacity	BPSD	13000	84.68	125.81
KHT	Feed Capacity	BPSD	3600	38.57	51.43
NHT	Feed Capacity	BPSD	17500	121.67	89.17
C5/C6 ISOM	Feed Capacity	BPSD	5100	92.42	96.97
CCR	Feed Capacity	BPSD	14400	108.45	85.92
GASOLINE BLENDER	Prod Capacity	BPSD	22000 MAX	105.91	88.89
JET BLENDER	Prod Capacity	BPSD	15000 MAX	92.17	92.17
ULSD BLENDER	Prod Capacity	BPSD	36000 MAX	96.27	101.12
COKE MAKE	Estimated Capacity.	ST/D	526	97.06	241.68
SULFUR MAKE	Estimated Capacity.	LT/D	18.5	186.50	69.33

Table 2 – Three heavy crude case (% of Case 10 A Yield basis)

4.0 ATTACHED MATERIAL, FUEL, HYDROGEN AND OCTANE BALANCE SPREADSHEETS

The attached spreadsheet, *Case10A_ValidationREV9.xls* holds the Phase II material balances contains trial refinery product stream hydrogen and fuel balances based upon the Phase I model modified with licensor information.

Within the spreadsheet, there is a revised volume balance worksheet where we mostly match good data in terms of volume yields. For each unit and stream, we match the volume balance to a weight rate and calculate a corresponding API gravity of each cut. This allows us to check on reasonability of the values. For example, light naphtha are expected to have an API value of 69-77, heavy naphthas 53-64, kero, 42-46, diesel 32-39, etc. We then adjusted volume and weight rates so that the unit balanced on weight exactly and the volume balance offset was minor. In fact, most hydrotreater models increase the volume yield of product by adding hydrogen and thus lightening the product gravity some. In hydrocrackers, this volume gain is even more pronounced.

In each spreadsheet, the refinery units where we are satisfied that the feed streams and yield products are reasonably well balanced have their title blocks outlined in yellow. Items of interest on this spreadsheet include:

ADU-1 Existing atmospheric preflash and crude tower for light sweet crudes (Cases 10A,B,and C use Pennington). Treated as one unit operation with CPF#1.

ADU-2 New atmospheric preflash and crude tower for heavier, more sour crudes (Case 10 uses Vasconia, 10 B uses Leona21, and 10C uses Roncador18). Treated as one unit operation with CPF#2.

VDU-2 New Vacuum Unit

DCU - New Delayed Coker Unit

ISOM – New C5/C6 Isomerization Unit

KHT – Existing Kerosene Hydrotreater assuming it is mechanically sound for re-use, which it appears to be from our limited initial examination. A more thorough inspection needs to be done to validate this assumption

NHT – New consolidated Naphtha Hydrotreater, which also serves as the naphtha desulfurizer for the feed to the Isom and CCR Units.

DHT – New consolidated Distillate Hydrotreater, which reduces the sulfur content of the feed to the diesel blenders to 15 ppm, max

HCU – New Hydrocracker Unit with a 98+% conversion of feed to diesel or lighter products.

The spreadsheet also has worksheets illustrating the limits of UPR vs URG for 4% MTBE and 10% Ethanol gasoline blends, as well as for fuel gas estimates in support of the hydrogen plant needs and the total refinery fired heater duty.

Spreadsheets will be found on the official project Soresco Sharepoint site.

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CASE 1B EVALUATION CASE
FLOW BALANCE WORKSHEET
[illegible] CRUDE TO GREENFIELD
[illegible] CRUDE TO BROWNFIELD

UNIT CAPACITIES

BROWNFIELD UNIT

GREENFIELD UNIT

REFINERY BALANCE SUMMARY

TOTAL GASOLINE	17,600	BPD	[illegible]	
TOTAL KERO/JET	[illegible]	BPD	[illegible]	KTA
TOTAL DIESEL	[illegible]	BPD	[illegible]	KTA
TOTAL LPG to sales	[illegible]	BPD	[illegible]	KTA
TOTAL LPG to fuel	[illegible]	BPD	0	KTA
TOTAL LIQUIDS	[illegible]	BPD	[illegible]	KTA
PROD YIELD On Crude	[illegible]	%	[illegible]	KTA

[illegible]

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MOIN REFINERY APP PAN
EXPANSION & MODERNIZATION PROJECT CASE 10 A VALIDATION CASE
ETHANOL BLEND SHEET

DUAL OCTANE POOL WITH ethanol

URG Naphtha Pool

Stream	bpd	RON	BPD*RON	ARO	bpd*ARO
RF90	4300	90	387,000.0	50.0	215,000.0
RF97	0	97	0.0	68.5	0.0
ISOM	600	90	54,000.0	1.3	804.0
SRBP	150	65	9,750.0	9.0	1,350.0
Sum hc	5050		450,750.0		217,154.0
ETOH (10.(	550.0	111	61,050.0	0.0	0.0
SUM TOT	5600.0		511,800.0		217,154.0
AVG		91.393		38.778	

UPR Naphtha Pool

Stream	bpd	RON	BPD*RON	ARO	BPD*ARO
RF90	0	90	0.0	50.0	0.0
RF97	8050	97	780,850.0	68.5	551,425.0
ISOM	4600	89	409,400.0	1.3	6,164.0
SRBP	300	65	19,500.0	9.0	2,700.0
Sum hc	12950		1209750	0.0	560289
ETOH (10.(	1530.0	111	169,830.0	0.0	0.0
SUM TOT	14480.0		1,379,580.0		560,289.0
AVG		95.275		38.694	

	Gaso Blender	
R89	CCR Reformate (89 RON A	4100
SRBP	ST RUN BYPASS OR PUR(	450
R97	CCR Reformate (98 RON A	8050
IN1	ISOM Isomerate to Gasoline	5100
ETHANOL	ETOH @ 10%	2080.0
		19780

Total Ethanol	2080.0 BPD
Total UPR	14480.0
Total URG	5600.0
	20080.0

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FUEL GAS	per FSR Table 9.4-2								n+n	N	N+n+M/2			
COMP	VOLFRAC reported	VOLFRAC recalc	MW	xpr	n	m	2N	m/2	prod make h20	prod make co2	prod make H2	LHV	xpr	
H2	0.1246	0.148404	2.016	0.299182	0	2	0	1			0.148404	274	40.6627	
CH4	0.421	0.501429	16.02	8.032897	1	4	2	2	1.00286	0.501429	2.005717	909	455.7992	
C2H4	0.0202	0.024059	28.04	0.674616	2	4	4	2	0.09624	0.048118	0.144354	1499	36.06455	
C2H6	0.181	0.215579	30.06	6.4803	2	6	4	3	0.86232	0.431158	1.509052	1619	349.0222	
C3H8	0.0403	0.047999	44.06	2.114838	3	8	6	4	0.28799	0.143997	0.47999	2315	111.1178	
C3H6	0.0235	0.02799	42.04	1.176679	3	6	6	3	0.16794	0.083969	0.251906	2182	61.07313	
C4H10	0.006	0.007146	58.1	0.415198	4	10	8	5	0.05717	0.028585	0.092901	3005	21.47451	
C4H8	0.01	0.01191	58.1	0.691996	4	8	8	4	0.09528	0.047642	0.142925	2875	34.2425	
CO2	0.0023	0.002739	44.02	0.120588	1	0				0.0023			0	
H2O	0.003	0.003573	18.02	0.064388	0	2			0.00715				0	
N2	0.0077	0.009171	28.01	0.256881	0	0	0	0					0	
	0.8396	1	20.32756	20.32756					2.576941	1.287198	4.77525		1109.457	BTU/SCF

bad form ! Have to normalize   Not unreasonable FG ratios however
One more reason to distrust FSR

LPG	1	1	51.08		3.5	9	7	4.5	7.00000	3.5	11.5		19600	btu/LB-Liq GPSA DATABOOK
50% c3-50% c4														

									LEFT SIDE	RIGHT SIDE
20.32756 kta FG	plus	46.43648 kta h2o	yields	56.66244 kta CO2	plus	9.626904 kta H2			66.76405	66.28934
51.08 kta FG	plus	126.14 kta h2o	yields	154.07 kta CO2	plus	23.184 kta H2			177.22	177.254

from FG	3.261057	kta H2O per kta FG
	2.787468	kta CO2 per kta FG
	0.473589	kta H2 per kta FG
from LPG	3.470125	kta H2O per kta FG
	3.016249	kta CO2 per kta FG
	0.453876	kta H2 per kta FG

FUEL VALUE OF LPG

Available Coker LPG	23.8	kta
LHV	19600	BTU/lb
	68	Mt/d
	149872	lb/d
	122.3955	MM BTU/hpotential duty

LHV CALCS

H2U OUTLET		xpr		
MPCTG H2	0.946	1.907136	58.85285	wt% H2
MPCTG CO	0.015	0.6603		
mPCTG ch4	0.015	0.2406		
Mpctg h2o	0.024	0.43248		
	1	3.240516		

CCR PSA O	before psa	after psa	lhv	
MPCTG H2	0.85	0.06	274	16.44
MPCTG ch4	0.09	0.09	909	81.81
mPCTG c2h	0.02	0.02	1619	32.38
Mpctg c3h	0.04	0.04	2315	92.6
	1	0.21		1063

H2U PSA V	before psa	after psa	lhv	
MPCTG H2	0.946	0.06622	274	18.14428
MPCTG CO	0.015	0.015	0	0
mPCTG ch4	0.015	0.015	909	13.635
Mpctg h2o	0.024	0.024	0	0
	1	0.12022		264.3427

MW CALCS

H2U OUTLET		xpr	lhv	
MPCTG H2	0.946	1.907136	2.016	1.907136
MPCTG CO	0.015	0.6603	44.02	0.6603
mPCTG ch4	0.015	0.2406	16.02	0.2403
Mpctg h2o	0.024	0.43248	18.02	0.43248
	1	3.240516		3.240216

CCR PSA O	before psa	after psa	lhv	
MPCTG H2	0.85	0.06	2.016	0.12096
MPCTG ch4	0.09	0.09	44.02	3.9618
mPCTG c2h	0.02	0.02	28	0.56
Mpctg c3h	0.04	0.04	44	1.76
	1	0.21		30.48933

H2U PSA V	before psa	after psa	lhv	
MPCTG H2	0.946	0.06622	2.016	0.1335
MPCTG CO	0.015	0.015	44.02	0.6603
mPCTG ch4	0.015	0.015	16.02	0.2403
Mpctg h2o	0.024	0.024	18.02	0.43248
	1	0.12022		12.19913

Option 1

FUEL GAS EKTA		MW		MOLE FR	LHV	
MAIN REF I	58.4	20.3	2.876847	0.186109	1109	206.3945
CCR PSA VI	26.3	30.5	0.862295	0.055783	1060	59.13049
H2U PSA V	37.5	3.2	11.71875	0.758108	265	200.8986
			15.45789	1		466.4236

Option 1

FUEL GAS EKTA		MW			MOLE FR	LHV	
MAIN REF I	58.4	20.3	1185.52	0.596437	0.596437	1109	661.4487
CCR PSA VI	26.3	30.5	802.15	0.403563	0.403563	1060	427.7767
H2U PSA V	0	0	0		0	0	0
	84.7		1987.67	1	1		1089.225

mw= 0.6*20.3+0.4*30.5   24.38

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EXPANSION & MODERNIZATION PROJECT

AS OF: 11-Apr-12
APP PAN

CASE 10 A VALIDATION CASE
ECONOMICS

BUY	UNITS	UNIT COST $USD	MTBE CASE QUANTITY	REVENUE K $/SD		ETOH CASE QUANTITIE	REVENUE K $/SD	
PENNINGT	BPSD	-113.95	25000	-2848.75		25000	-2848.75	
VASCONIA	BPSD	-102.76	40000	-4110.40		40000	-4110.40	
MTBE	BPSD	-138.54	708	-98.09				
ETOH	BPSD	-120.16				2088	-250.89	
SRBP (Note	BPSD	-110	450	-49.50		450	-49.50	
			TOTAL BUY	-7106.74	$K/D	485	-7259.54	
				-109.33	$/BBL Crude		-111.69	/BBL Crude
SELL	**UNITS**	**UNIT COST $USD**						
UPR	BPSD	121.65	13710	1667.82		14480	1761.49	
URG	BPSD	117.39	4580	537.65		5600	657.38	
JET	BPSD	124.99	9070	1133.66		9070	1133.66	
USD	BPSD	127.22	35490	4515.04		35490	4515.04	
COK	Long tons	115.58	526	60.80		526	60.80	
SULFUR	Long tons	46.01	52.9	2.43		52.9	2.43	
LPG	BPSD	84.41	0	749.20		0	749.20	
FUEL GAS (	MM BTU/d	NA	NA	0.00		NA	0.00	
			TOTAL SELL	8666.59	$K/D		8880.00	
			Total Liquid Prod	62850	BBLS		64640	
				133.3322	$/BBL Crude		136.6154	$/BBL Crude

Productio	Amount BPD	Projected Demand 2015	2020
Total Gasol	22.8	19.4	22.1
Jet	13.5	11.5	13.9
USD	26.8	25.4	32.4
LPG	1.23	5.0	6.5

Notes
1. SRBP is straight run CCR feed bypass or purchased outside stock used to trim octane giveaway

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Year	0 2013	1 2014	2 2015	3 2016	4 2017	5 2018	6 2019	7 2020	8 2021	9 2022	10 2023	11 2024	12 2025	13 2026
Initial Investment	[redacted]	[redacted]	[redacted]											
Incomes Product Price in CR														
LPG				23,322,308	24,429,850	25,529,391	26,632,933	27,736,475	28,607,378	29,476,291	30,349,184	31,220,087	32,090,990	33,164,706
Unleaded Regular				242,024,018	253,007,246	264,063,431	275,336,993	286,247,562	294,858,065	303,468,568	312,079,071	320,689,574	329,300,077	340,061,082
Unleaded Premium				647,391,068	675,790,576	704,377,938	733,528,922	761,740,536	784,004,836	806,269,137	828,533,438	850,797,738	873,062,039	900,886,925
Kero/Jet				416,903,406	434,402,268	452,026,147	470,025,078	487,398,923	501,976,709	516,554,495	531,132,281	545,710,067	560,287,853	578,419,257
Diesel				1,659,932,000	1,729,913,666	1,799,693,391	1,873,383,656	1,944,027,548	1,997,785,175	2,054,542,801	2,111,300,428	2,168,058,054	2,224,815,681	2,296,773,358
Coke, KTONS				21,890,604	22,744,927	23,567,446	24,472,792	25,215,928	25,947,187	26,696,174	27,228,501	27,698,070	27,962,756	28,240,550
Sulfur, KTONS				1,147,264	1,141,567	1,135,849	1,130,141	1,124,434	1,118,726	1,113,019	1,107,311	1,101,604	1,095,896	1,090,189
Premium				25,262,027	25,893,578	26,540,918	27,204,440	27,884,551	28,561,665	29,296,207	30,028,612	30,779,327	31,548,841	32,337,531
Total				3,037,672,716	3,167,319,667	3,297,126,151	3,431,714,955	3,556,375,956	3,662,879,741	3,767,358,681	3,871,750,826	3,976,054,821	4,080,164,102	4,210,973,597
Expenses Crude Price in CR (-)														
Pennington				1,041,512,500	1,086,225,000	1,130,937,500	1,175,912,500	1,221,325,000	1,258,162,500	1,295,000,000	1,331,925,000	1,368,850,000	1,405,775,000	1,451,012,500
Vasconia				1,503,564,569	1,571,180,227	1,638,810,468	1,705,822,333	1,773,799,505	1,828,588,652	1,883,393,896	1,938,215,640	1,993,054,295	2,047,910,256	2,116,110,705
Ethanol				93,793,605	98,157,899	102,550,316	106,794,405	110,901,732	114,143,189	117,384,646	120,626,104	123,867,561	127,109,018	131,160,041
Total				2,638,870,674	2,755,563,126	2,872,298,284	2,988,529,238	3,106,026,236	3,200,894,341	3,295,778,542	3,390,766,743	3,485,771,856	3,580,794,304	3,698,283,246
Gross Margin				[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]
Gross Margin per barrel				[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]
Operating costs				[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]
Working capital				[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]
Rescue Value (+)														
Net Margin				[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]
Net Flow US$	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]

Year	14 2027	15 2028	16 2029	17 2030	18 2031	19 2032	20 2033	21 2034	22 2035	23 2036	24 2037	25 2038	26 2039	27 2040
Initial Investment														
Incomes Product Price in CR														
LPG	34,238,423	35,312,139	36,385,855	37,459,571	39,830,615	42,188,098	44,530,630	46,657,568	49,167,993	51,173,379	53,260,558	55,432,866	57,693,774	60,016,896
Unleaded Regular	350,676,271	361,292,459	371,908,148	382,523,837	403,066,053	423,539,799	444,017,028	464,497,012	484,978,998	503,424,539	522,565,404	542,437,262	563,064,797	584,476,747
Unleaded Premium	928,336,062	955,785,200	983,234,338	1,010,683,476	1,063,801,862	1,116,746,014	1,169,702,736	1,222,670,920	1,275,649,420	1,322,900,043	1,371,963,074	1,422,813,463	1,475,548,570	1,530,238,248
Kero/Jet	596,300,026	614,181,995	632,063,364	649,944,734	683,921,745	717,794,465	751,677,625	785,579,695	819,498,897	849,789,850	881,200,675	913,778,680	947,567,922	982,555,865
Diesel	2,366,393,262	2,436,013,166	2,505,633,070	2,575,252,974	2,704,457,517	2,837,764,223	2,977,837,768	3,112,777,608	3,245,497,033	3,364,504,577	3,487,938,151	3,615,657,890	3,748,149,806	3,888,975,995
Coke, KTONS	28,812,911	29,491,220	30,026,875	30,759,800	31,432,276	32,124,301	32,856,027	33,525,331	34,314,699	35,145,542	35,996,502	36,868,066	37,760,732	38,675,013
Sulfur, KTONS	1,084,481	1,078,774	1,073,066	1,067,358	1,061,651	1,055,943	1,050,236	1,044,528	1,038,821	1,051,686	1,064,710	1,077,896	1,091,245	1,104,760
Premium	33,145,969	33,971,616	34,823,984	35,694,583	36,566,948	37,501,622	38,439,162	39,400,141	40,388,145	41,364,773	42,429,613	43,490,384	44,577,643	45,692,084
Total	4,338,988,505	4,467,129,571	4,595,148,701	4,723,386,333	4,964,158,867	5,208,711,469	5,460,111,412	5,706,352,801	5,950,528,006	6,169,444,370	6,396,421,687	6,631,766,507	6,875,784,489	7,126,795,608
Expenses Crude Price in CR (-)														
Pennington	1,496,950,000	1,542,975,000	1,589,000,000	1,635,025,000	1,725,325,000	1,815,537,500	1,905,750,000	1,996,050,000	2,086,437,500	2,165,179,598	2,246,893,420	2,331,691,120	2,419,689,083	2,511,008,087
Vasconia	2,184,645,999	2,253,199,960	2,321,773,056	2,390,365,764	2,526,607,690	2,667,415,616	2,806,155,272	2,944,825,766	3,083,426,276	3,204,354,613	3,327,947,501	3,457,386,368	3,591,859,696	3,731,563,296
Ethanol	135,156,356	139,152,676	143,148,993	147,145,310	154,678,811	162,566,945	170,296,909	178,008,542	185,721,677	191,682,368	199,903,911	207,396,172	215,169,205	223,233,565
Total	3,816,752,357	3,935,327,636	4,053,922,049	4,172,536,075	4,406,611,501	4,645,540,061	4,883,202,181	5,118,884,308	5,355,585,454	5,561,216,579	5,774,744,842	5,996,473,660	6,226,717,984	6,465,804,949
Gross Margin	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]
Gross Margin per barrel	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]
Operating costs	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]
Working capital	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]
Rescue Value (+)														20,524,493
Net Margin	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	4,813,602,858
Net Flow US$	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]	[redacted]

[redacted]

LAC Fee [redacted]