# Comparison of Draft NETL and ENSYS Capital Models For Distillates Deep Hydro-desulfurization

**Draft Final Report**
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**For**
**Energy Information Administration**

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# Comparison of Draft NETL and ENSYS Capital Models
# Summary Findings and Recommendations
# Draft Final Report

This report by EnSys follows a request from EIA to compare capital costs for diesel deep hydro-processing units in the NEMS model, based on the EnSys WORLD/RYM technology database, with those tentatively developed by NETL for EIA. The focus of concern is ultra low sulfur diesel (ULSD) at a sub 10 ppm sulfur target per the recent EPA on-road final rule. To ensure a thorough assessment based on detailed knowledge of the processes involved, and an impartial perspective favoring neither the EnSys nor draft NETL data, EnSys retained the services of Dr. Richard Foley, an acknowledged expert and independent consultant.

The second section of this report contains Dr. Foley's analysis. The first section below provides a summary and EnSys' specific recommendations for immediate and later adjustments to data in the NEMS/WORLD/RYM technology database.

## Summary

### Key Findings of Foley Analysis

#### *DRAFT NETL Capital Cost Model*

The draft NETL capital cost model was found to have been derived from a 1991 Bechtel model for a coal liquefaction synthetic naphtha hydrotreater. While NETL put considerable effort into adjusting reactor size and pressure rating for diesel type feedstocks, their model did not alter the reactor internals or the non-reactor naphtha hydrotreater process facilities representing around 90% of the total onsite cost. As detailed in the Foley report, the naphtha hydrotreater had numerous process facilities either under-rated or absent that are essential in a ULSD unit. As a consequence of this, and of the lack of reactor internals, the draft NETL capital costs are approximately 40% below quoted costs from major vendors such as UOP, IFP and others for similar duty units

#### *EnSys Capital Costs*

The Foley report also compares quoted vendor costs with the EnSys capital costs in NEMS/WORLD/RYM. It is acknowledged the latter were based on "reliable supply" technologies by virtue of the terms of reference of EnSys'

Summer 2000 ULSD study for DOE. This called on EnSys to assess the costs for reliable supply of ULSD. The costs were based on two stage technology and on assessment of the state of the diesel HDT technology as of one year ago. EnSys provided two first stage and two second stage units, respectively HL1/HD1 and HS2/HD2. The performance objective of HL1 and HD1 was to reduce sulfur to approximately 30 ppm. The units were split only for notional purposes to represent different capital cost levels for low sulfur versus high sulfur feed streams. The objective of HS2 and HD2 was sulfur reduction from around 30 ppm to sub 10 ppm. The difference was that HS2 was designed for deep desulfurization with only incidental de-aromatization/cetane/gravity improvement whereas HD2 was designed for deep desulfurization plus substantial cetane/aromatics/gravity improvement. The intent of HD2 was to capture those situations, such as in Europe where extreme low sulfur and aromatics/poly-nuclear aromatics control is required and such as in the US where high severity cracked stock streams needed to be processed for inclusion in ULSD. For these latter blendstocks to be acceptable, it is not sufficient to reduce their sulfur. It is also essential to improve their cetane and gravity. EnSys anchored the costs of the HD2 unit based on confidential data for a refinery actually processing these "worst" cracked stocks.

The Foley analysis states EnSys' total costs for getting to <10 ppm as around 20% above "straight" vendor costs for straight run components and around 50% above for cracked stocks. Further analysis of this result showed that:

(a) EnSys had exercised conservatism through anchoring its costs to two-stage processes, but

(b) had also overstated its capital costs through a misallocation between on-sites and total costs. This was especially the case for the severe cracked stocks.

A further factor in the large cost variation was that the Foley vendor costs for cracked stock streams were for sulfur reduction only (not sulfur plus cetane improvement).

In summary, the Foley analysis, which we believe is a sound representation of the costs for new units purchased in the 2002-2006 timeframe, supports a capital cost range, including 20% contingency per NPC and CRA, of around $1,600-$2,500/bbl/sd (ISBL) for a sub 10 ppm sulfur target with only incidental cetane/aromatics/gravity improvement. The cost level depends on feed stream type and quality. For processing severe (cracked stock) streams to sub 10 ppm with the combined target of cetane/aromatics/gravity improvement, the stated cost is up to $3,500+/bbl/sd. These figures are based on vendor cost estimates adjusted upward with a 20% contingency factor and to a 25,000 bpsd size base. (This is the size base in the NEMS/WORLD/RYM database and is designed to represent the average size of major new units in US refineries. Costs can be adjusted to say a 30,000 bpsd basis by applying a 0.6 power factor.) This cost range is broadly consistent with the CRA cost basis – in turn understood to have been taken from the NPC 2000 report. While CRA gives little detail on feedstocks, their quoted (average) cost is $1780 bbl/sd, including a 20% contingency factor. By way of comparison, the draft NETL ISBL costs at 25,000 bpsd were around $950-1050/bbl/sd.

Applying the revised EnSys capital costs to our DOE PADD3 Summer 2000 ULSD study would reduce the ULSD total costs by 1.1-1.2 c/gal at 10-8 ppm (base cost projection, 0% revamp, data adjusted post-optimally i.e. no case re-optimization).

### *Revamp Potential & Costs*

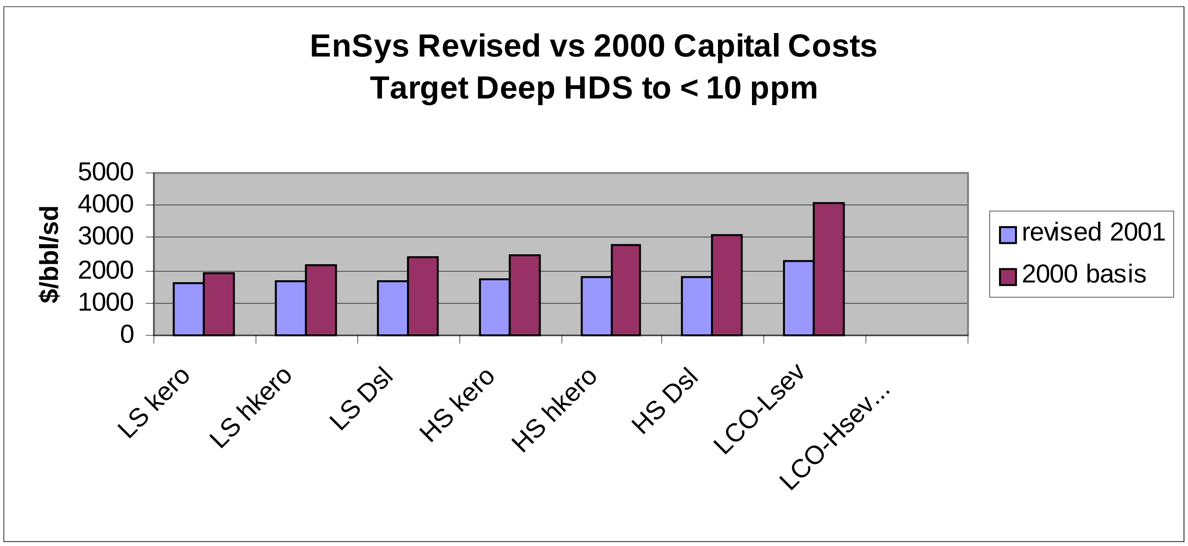
The Foley report also points to recent papers which detail and confirm the ability to revamp existing high and low pressure HDS units that are already producing 500 ppm diesel. For existing units with high pressure reactors (800 psig), the estimated ISBL revamp cost is around $700/bbl/sd including (20%) contingency. For lower pressure units (400-600 psig), the cost is approximately $1200/bbl/sd ISBL including contingency. This clarification is significant, as the majority of USA refineries are current producers of 500 ppm LSD.

Scope of revamp potential must be treated with some caution however. As stated in the NPC report, the potential to revamp units processing streams containing significant amounts of cracked stocks is much more limited than potential for units processing mainly straight run blendstocks. In addition, older units in particular may have mechanical or other limits that render them unattractive for revamping, especially if a significant increase in capacity is sought, e.g. if off-road diesel moves to ULSD quality.

## Recommendations for Adjusting NEMS/WORLD/RYM Capital Costs

Based upon the Foley report and on corrections to the existing model cost basis, EnSys proposes revising the ISBL capital cost figures and capacity factors to be entered into Table INVUNT. Broadly, through corrections to ISBL versus total costs, and through allowance for greater confidence today versus a year ago in single stage processes to reach 10 ppm sulfur target with incidental cetane/aromatics/gravity improvement, the capital cost reductions are approximately 30% versus the values EnSys used for its DOE work in 2000. These apply to sulfur reduction, not sulfur plus cetane/gravity improvement. For this latter category, our average capital cost reductions are 12%. We have still anchored the high end of our capital costs, i.e. sulfur reduction plus cetane/gravity improvement on severe cracked stocks, to confidential industry data for actual units. Table 1 and the two Figures below summarize the revised versus original capital costs expressed as ISBL $/bbl/sd. Arguably, NEMS/RYM/WORLD modeling should force a certain proportion of heavily processed (severe) cracked stocks into diesel to reflect that certain refiners, notably in PADDs 3 and 5, operate in this mode.

| **Table 1: Original and Revised Capital Costs $/bbl/sd ISBL** | | | | |
|---|---|---|---|---|
| Target < 10 ppm | **deep HDS** | deep HDS | **deep HDS** | deep HDS + HAD/cetane/API |
| | | | | |
| | ***revised*** | *original* | ***revised*** | *original* |
| | ***2001*** | *2000* | ***2001*** | *2000* |
| LS kero | **1600** | 1900 | **2100** | 2080 |
| LS hkero | **1650** | 2150 | **2260** | 2350 |
| LS Dsl | **1680** | 2380 | **2390** | 2610 |
| HS kero | **1730** | 2490 | **2230** | 2670 |
| HS hkero | **1780** | 2800 | **2390** | 3000 |
| HS Dsl | **1820** | 3110 | **2530** | 3340 |
| LCO-Lsev | **2270** | 4050 | **3290** | 4340 |
| LCO-Hsev/CGO | **n.a.** | n.a. | **3800** | 5010 |



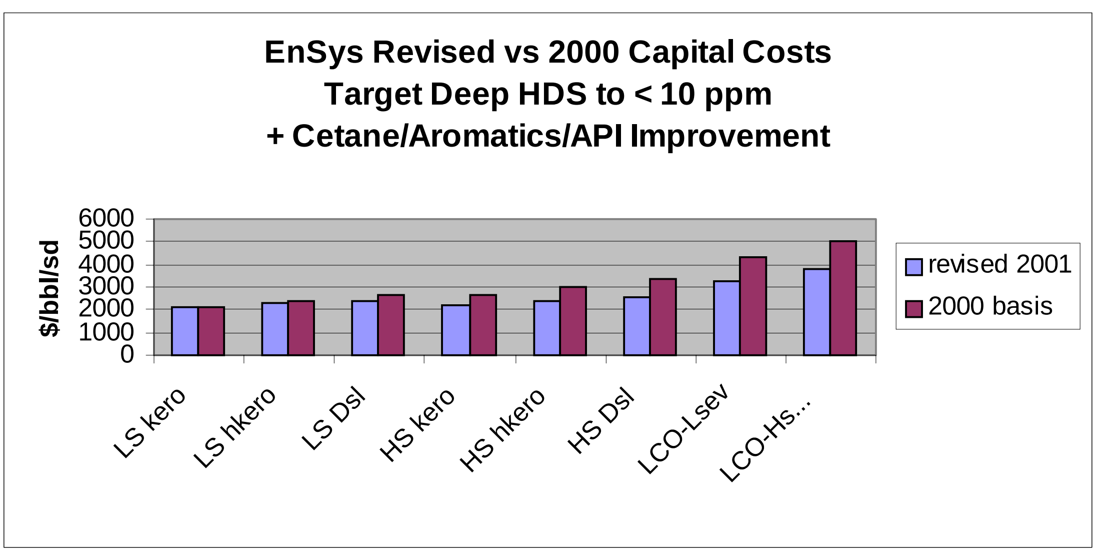



Table 2 sets out the proposed revised entries into Table INVUNT. In this corrected and updated formulation, the cost of unit HS2 effectively represents the add-on cost of taking HL1 or HD1 to a sub 10 ppm sulfur target as distinct from sub 50 ppm. The labor (LAB) entry in Table INVUNT should therefore be set at zero for unit HS2. The still high cost for unit HD2, and the high CAP factors for severe cracked stocks on unit HD1 represent the much higher costs of processing these streams for diesel use, i.e. costs that entail cetane/aromatics/gravity improvement as well as sulfur reduction.

In the future, vectors for severe cracked stocks could be added to unit HS2. However these would only likely be relevant in a scenario where say all off-road diesel and heating oil were being reduced to 15 ppm. Also, unit HL1 could be consolidated into unit HD1. (Original EnSys costs had understated the costs of HL1 streams relative to HD1.) Consolidation would entail moving the HL1 vectors into HD1 and multiplying the HL1 vector CAP factors listed below by 1240/1380 = 0.90.

| **Table 2** | | | INVUNT | CAP | effective |
|---|---|---|---|---|---|
| | | | Cap Cost | factor | Cap Costs |
| | | | New Unit | | New |
| | | | DslB feed | Tables HL1 | |
| Unit | prod S | Duty | $mmISBL | etc. | $mmISBL |
| | | | | | |
| | | | 1.2 | | |
| | | | | | |
| HL1 | 30 | LS kero | | 0.95 | 1180 |
| | | LS hkero | | 0.98 | 1220 |
| | | LS Dsl | 1240 | 1 | 1240 |
| | | | | | |
| HD1 | 30 | HS kero | | 0.95 | 1310 |
| | | HS hkero | | 0.98 | 1350 |
| | | HS Dsl | 1380 | 1 | 1380 |
| | | LCO-Lsev | | 1.3 | 1790 |
| | | LCO-Hsev/CGO | | 1.5 | 2070 |
| | | | | | |
| HS2 | 7 | SR kero | | 0.95 | 420 |
| | | SR Hkero | | 0.98 | 430 |
| 30% HDa | | SR DslB | 440 | 1 | 440 |
| | | LCO (5) | | 1.1 | 480 |
| | | LCO (6) | no vector | | |
| | | | | | |
| HD2 | 7 | SR kero | | 0.8 | 920 |
| | | SR hkero | | 0.9 | 1040 |
| 70% HDa | | SR DslB | 1150 | 1 | 1150 |
| | | LCO (5) | | 1.3 | 1500 |
| | | LC6/CGO | | 1.5 | 1730 |

# Report by Dr. Richard Foley PhD, Consultant

**Review of Activities April 10 to 19, 2001**

Draft NETL Model:

a) Draft NETL spread sheet was provided by ENSYS, and the structure of the model could be determined
b) Requested information on basis for numbers in draft NETL model from John Hackworth (Independent consultant who was employed by Gulf in their Pittsburgh R&D Lab prior to Chevron merger)
c) NETL employees Howard McIlvred (also former Gulf employee), and Jared Ciferno, who had developed the model provided details on the basis for the model on April 18. Several pages of the Bechtel report, including a process description and cost estimate summary were provided

ENSYS Model and other Information Sources

a) Full ENSYS model information that had been used in their study for DOE was provided by EnSys, and had been available to the public from earlier publications.
b) Information was also accessed from:
   i) MathPro study for the EMA
   ii) NPC Study
   iii) Charles River study for the API
c) NPRA and Other Publications (See references)

Comparisons with draft NETL Model

a) Capital costs for ULSD production provided by UOP to the API/NPRA and others published by IFP for similar product objectives were compared with the draft NETL results
b) Capital costs and equipment required for revamping 500 ppmS hydrotreaters to produce ULSD were summarized
c) These equipment requirements were compared with the Bechtel NHT used for the draft NETL model

## Major Issue: State of Technology

ULSD production from the diesel feedstocks common to US refineries requires different technology than has been used for 500 ppm diesel. This technology has been evolving, and reduced cost technologies are being developed.[1,2,3,4,5,6,7] Full reliability basis commercial experience was required for the ENSYS study for the DOE, and this resulted in the utilization of two stage processes. Single stage processes are now being licensed by technology suppliers, which will produce ULSD with a product sulfur at or below 10 ppm.

The technology suppliers have had to re-evaluate their technology multiple times over the last 3 years as diesel sulfur target expectations moved from 150 to 50 to 30 to 10 ppm. Thus, their confidence in a single stage solution has had to be extended, and additions have been made to the technology at several target S levels.

Technology suppliers are now fully confident in their ability to achieve 10 ppm with a single stage process, and this technology is being tested in refining applications. Significant technology advances have been made in the areas of catalysis and reactor internals, but most other changes have only required application of commercially demonstrated engineering solutions (eg recycle gas scrubbing). The combined application of these technologies is being confirmed as noted in the papers referenced above.

With each of these changes has come additional cost, and the 500 ppm DHT which might have cost $1000-$1200/BPSD, is now expected to cost significantly more than that amount. These costs are discussed next:

<u>Modifications to 500 ppm Diesel Hydrotreaters to produce <10 ppm S Diesel</u>

Mustang Engineering presented a Paper at the 2001 NPRA that discusses the technology changes required to accomplish this product sulfur reduction.[8] This paper illustrates the additional equipment required in ULSD hydrotreaters that were not included in 500 ppm DHT units.

Revamps to two “800 psig” DHT units, each with 30 MBPSD are discussed, and the costs of these changes are shown:

| Unit 1 Changes | Cost $Million | Unit 2 Changes | Cost $Million |
|---|---|---|---|
| - Add Hydrogen booster compressor and replace cylinders in existing make-up compressor<br>- Add quench and feed distributors to reactor<br>- Add PSA unit | $10.27<br><br>$340/ BPSD | - Unit 1 changes, excluding PSA, plus<br>- Recycle gas amine scrubbing (already in unit 1) | $7.47<br><br>$250/BPSD |
| - New reactor with 100 % catalyst addition | $8.25<br><br>$275/BPSD | - New reactor with 100% catalyst addition | $8.6<br><br>$290/BPSD |
| Total, $Million On-Site Capital | $15.52<br>$615/BPSD | - | $16.07<br>$540/BPSD |

The differences between these numbers for the two units are created by these being actual current units, with different hardware.

UOP[1] estimated the cost of adding comparable capability at  $15 million for a 30 MBPSD unit. The addition of a second unit to reduce the 500 ppm S diesel to <10 ppm was estimated at $30 mm, making the revamp more economic. In each case these are onsite capital costs only. Off site capital depends on what other facilities are available at the specific refinery and will be strongly dependent on the availability of pipeline hydrogen.

<u>Full ULSD Unit Costs</u>

In 1999 the API and NPRA requested that 5 technology vendors provide estimates for units to produce LSD at 100, 50 and 10 ppm S from a range of feedstocks, based on the 1997 API/NPRA survey of Refiners:

| Feed/Composition as Percentage | SRGO | LCO | Coker | Sulfur ppmw |
|---|---|---|---|---|
| 1. Reference Feed | 69 | 23 | 8 | 9000 |
| 2. Lo S Reference | 69 | 23 | 8 | 5000 |
| 3. 80% SR | 80 | 15 | 5 | 7500 |
| 4. 50% SR | 50 | 40 | 10 | 11,000 |
| 5. Three Equal Fractions | 34 | 33 | 33 | 17,500 |

UOP, IFP, Haldor Topsoe, MAK, and SYNSAT all provided single stage LSD unit costs to the NPRA and API, for these feedstocks, with the varying sulfur objectives. UOP[7] provided the following input for 10 ppm ULSD:

| Feed | Total Capital, MM$ (30 MBPD, 1999$) | Capital $/bpsd* (25 MBPD, 2000$) |
|---|---|---|
| 1. Reference Feed | 53 | 1960 |
| 2. Lo S Reference | 52 | 1920 |
| 3. 80% SR | 53 | 1960 |
| 4. 50% SR | 56 | 2070 |
| 5. Three Equal Fractions | 61 | 2250 |

*Calculated using 0.65 scale factor, and 4% annual cost adjustment

Table 3 below compares the UOP results and draft NETL and ENSYS model results for the Reference feed and the 50% SR feed. The draft NETL capital is about 70% of the UOP capital in both cases, and the ENSYS capital exceeds the UOP by 20% and 50% because of the addition of costly second stages by ENSYS.

IFP estimated the on-site cost of DHT units in their 1997 publication[9] for a 50:50 SR:LCO feed with 18,600 ppmS. Capital cost varies significantly, depending on product objective:

**IFP DHT Capital Cost Estimates**

| Product Objective | Stages/Cat Type | Product S | Product Cetane | Onsite $/BPSD (2000$ at 25 MBPSD scale) |
|---|---|---|---|---|
| Sulfur Reduction | 1/ Base Metal | 500 | 42 | 1200 |
| Product Stability | 1/ Base Metal | <200 | 43 | 1440 |
| Moderate Cetane Improvement | 1/ Base Metal | <50 | 48 | 1800 |
| High Cetane Improvement | 2/ Noble Metal | <5 | 55 | 2480 |

None of these product objectives are focused directly on 10 ppm S. With a 40% offsite cost, the Product Stability unit would come in somewhat below the total capital cost of the 10 ppm UOP unit. The addition of minor capital should allow this high pressure unit to achieve the additional sulfur reduction to 10 ppm. Significant catalyst improvements have also occurred since this paper was published which would result in lower sulfur levels than were estimated.

Adding cetane improvement, as would be required in Europe to reach the higher cetane level (50) in diesel, brings the total capital range up to $2500-$3500/bpsd for this moderate difficulty feedstock. However, some US refineries which refine very heavy crudes and rely on coking for residue upgrading, also require cetane upgrading to meet the 40 cetane US requirement. The “Equal Fractions” feed used for the API/NPRA work reflects this type of feed, but the UOP capital figure did not include the added cost of increasing the cetane number of this feed, only the sulfur reduction. This is a substantially more difficult feed than was used for the IFP estimates, and capital costs will exceed the figures discussed above for two stage units.

## NPC and Charles River Study Capital Costs

The Charles River study reported relying on capital costs generated by the NPC for their study. The NPC report includes a $42 MM capital cost in 1998$ for a 35 MBPD ULSD. Adjusting to 25 MBPD and 2000$ this number reaches $1480/bpsd. Charles River utilized offsite costs between 25 and 40% of onsite, 20% contingency, and a 20% re-run requirement. Combining these factors with the NPC capital as reported, would push the total USGC capital basis above $2600/bpsd. If the NPC capital already included offsites, the Charles River Capital would then be about $2100/bpsd. Reporting on these issues is sketchy.

## Draft NETL Model Design Basis

The draft NETL model capital cost was based on a Naphtha Hydrotreater design for treating the naphtha product from a Coal Liquefaction Plant. Bechtel designed the plant in 1991, and provided the cost estimate.

NETL calculates the capital cost by adding two numbers, the cost of the reactors and the cost of the rest of the equipment. The reactor cost is calculated separately for each feed and product based on the LHSV calculated to be required to achieve the ULSD product sulfur level. The rest of the ULSD process equipment is assumed to cost the same as the Bechtel design equipment, after adjustment for unit feed rate. No validation of this assumption has been provided.

Comparing the Bechtel process description with the ULSD process design provided in the Mustang paper above, yields the following reactor and other process equipment comparisons:

| **Reactor Comparison** | BECHTEL NHT Process | ULSD Process | Comments |
|---|---|---|---|
| Feed and Product | Naphtha from coal liquefaction plant (C5-350 F Product) to O.2 ppmw N | 2/3 SRGO and 1/3 LCO with 18, 000 ppm S (D-86 90% point 640F) to <10 ppmw S | Feeds and product objectives are totally different , but determination of the LHSV required should allow for catalyst volume to be adjusted |
| Reactor Conditions | 950 psig Vapor Phase with 700 psig H2 at reactor outlet, 2.0 LHSV | 760 psig Liquid Phase with 385 psig H2 at reactor outlet, 0.75 LHSV | NETL claims to be able to estimate the cost of reactors for the ULSD application basis a method for which they have not provided validation |
| Reactor Average Temperature | 550 deg F | 635 deg F | Combined reactor T, P , H2pp, and H2S pp differences could result in different reactor metallurgy |
| Hydrogen Recirculation through reactor | Not provided, but lower basis consumption below | 1400 SCF/B | Hydrogen recycle compressor would be undersized in Bechtel design/estimate |
| Hydrogen Consumption in reactor | 125 SCF/B | 355 SCF/B | Hydrogen makeup compressor would be too small in Bechtel design/estimate, but no make up compressor is mentioned in the Bechtel process description |
| Process Comparison (Continued) | BECHTEL NHT Process | ULSD Process | Comments |
| Interstage Hydrogen Quench in reactors | Not required basis low Hydrogen Consumption | Required to limit outlet temperature basis high hydrogen consumption | Adds significantly to reactor volume and cost |
| Feed Distribution and Redistribution | No Special Provision | Specialized Internals designed for Trickle flow reactors with no tolerance for bypassing | Adds significantly to reactor volume and cost |

Given the above differences, validation of NETL’s proposed reactor cost calculation technique is essential.

| **Other Equipment Comparison** | Bechtel NHT Process | ULSD Process |
|---|---|---|
| Reactors | 2 | 2 |
| Heat Exchange: | | |
| Product to Feed | Yes | Yes |
| Reactor Effluent to Feed | Yes | Yes |
| H2 Preheat | No | Yes |
| Stabilizer Feed | No | Yes |
| Product Air Cooler | Yes | Yes |
| Vessels: | | |
| Product Separators | High and Low pressure | Cold and Hot |
| Recycle Gas Scrubber | No | Yes |
| Product Stabilizer | Refluxed Column with Reboiler, Condenser and Accumulator | Steam Stripper with Overhead Condensor, and Accumulator |
| Compressors: | | |
| Makeup | No | Yes |
| Recycle | Yes, small | Yes |
| Hydrogen Purification: | Not required because pure H2 expected to be available | Required to utilize Cat Reformer hydrogen in refineries |

Given these significant differences in the other equipment contained in these two process plants, it would be fortuitous if the non-reactor costs would be equivalent at the same capacity. There is absolutely no reason to believe that the individual pieces of equipment, even when both designs include them, would be the same cost.  To expect their sums to be equal is unrealistic.

<u>Draft NETL Cost Totals</u>

The Bechtel NHT on site cost estimate was $900/BPSD (1999$), with $820/BPSD attributed to non-reactor costs at an 18 mbpsd throughput. Adjusting this to 25 mbpsd throughput, and 2000$, the non-reactor costs are only $760/BPSD. Reactor costs added to these on the same basis start at $75/BPSD, and despite the NETL adjustment, do not bring the cost of the Bechtel NHT into a realistic range for ULSD processing.

<u>ENSYS Model</u>

Acceptance of single-stage ULSD units as commercially viable requires changes to the ENSYS model. My suggestions are:

1. For refiners who have 500 ppm hydrotreaters, there are two categories:
   a) Those with high pressure units comparable to the base design mentioned in the UOP[1] and Mustang[8] papers.
      i) These units can be revamped at costs calculated from these two papers of about $600/BPSD (ISBL, before contingency)
   b) Those who have revamped their low pressure hydrotreaters (400 to 600 psig) to reduce sulfur to 500 ppm
      i) These units can be supplemented by an additional unit to reduce sulfur levels below 10 ppm at a cost of about $1000/BPSD basis a UOP[1] estimate (ISBL, before contingency).
      ii) If the current unit can be used to meet off-road diesel or heating oil sulfur levels, or has neared the end of it's mechanical viability, those units can be replaced at costs of about $2000/BPSD per the UOP[7] estimates.
2) Refiners who are currently treating a high percentage of cracked and coker stocks (60 to 70%), to improve cetane in the US have achieved S levels below 500 ppm as a byproduct of this treatment. The additional cost of reducing sulfur below 10 ppm for these refiners will be higher than estimated by UOP or IFP.
   a) Confidential information is available to ENSYS for estimating these costs.
   b) More of these stocks will need to be upgraded if off-road diesel is subjected to reduced sulfur levels
   c) These stocks can be treated for sulfur reduction for heating oil at costs comparable to the UOP ULSD estimates.

## Conclusions

**1. Draft NETL Economic Factors are reasonable, and consistent with those used by ENSYS**

2. Draft NETL Capital cost estimates do not reflect estimated costs for ULSD applications because of missing/undersized equipment and inadequate accounting for the specialized nature of ULSD reactors.

3. Draft NETL ULSD model does not include all the equipment and capability which technology providers believe is required for ULSD production

4. ENSYS capital cost estimates are higher than current ULSD technology costs since they reflect higher requirements for technology commercialization

5. ENSYS costs should be restructured to reflect revamped and new single stage units. Cetane improvement must be included for US and non-US refiners to enable additional cracked stocks to be included in diesel.

Table 1

Economic Factor Comparison

| Economic Factors | NETL | ENSYS |
|---|---|---|
| Capital Contingency | 20.0% | 15% |
| Unit Size, MBPSD | 50 | 25 |
| Effect on Cost vs 25 M | -22% | Base |
| Capital Charge 10% ROI | 17.2% | 17.20% |
| 15% ROI | | 24.20% |
| Maintenance (on-site) | 4.0% | 4% |
| Maintenance (off-site) | 2.0% | 2% |
| Taxes, Insurance, and Misc | 2.1% | 2.10% |
| Stream Days Operating | 311 | |
| Stream Factor | 0.85 | 0.85 |
| Offsite Capital, %Onsite | 45 | 40 |
| | | |
| Revamp Unit Cost | Calculated | 60% |
| Offsite Capital, %Onsite | 30 | 40 |

**Table 2**
**Cost Estimate Basis Comparison**

| COST ESTIMATE BASIS | |
|---|---|
| **NETL** | **ENSYS** |
| Base Estimate: Bechtel Synfuels Study from '91 or '92 | Base Estimates: UOP, Lyondell, IFP, API/NPRA Roundtable (1998-2000) |
| Feedstock: Synfuel Naphtha | Feedstock: Diesel and Blendstocks |
| Product Objective: Uncertain | Product Objective: ULSD |
| Escalated to Current Costs Using Cost Indices | Escalated to Current Costs Using Cost Indices |
| Methodology: | Methodology: |
| Separate Reactor Costs (Installed?) from other ISBL capital costs | Select base single stage estimate for feedstocks and sulfur levels, with weighting for feed blend components |
| Estimate size of Reactors for two stages using kinetics of HDS reaction, feedstocks, and HDS requirements | Add second stage costs depending on degree of aromatics saturation required, with weighting for feed blend components |
| Estimate cost of reactors for these two stages | Combine first and second stage costs. |
| Scale the cost of the other ISBL costs from the Bechtel estimate to the size required | |
| Add these costs together and adjust for location factor | |
| Revamps: Add the cost of one reactor, and increase installation cost, plus interstage scrubbing | Revamps: 60% 0f new unit cost |
| | |
| | |

**Comparison of Draft NETL and ENSYS Capital Model for EIA**
**Report by Richard Foley, Consultant**

## Table 3
## Model Results Comparison with UOP Estimates

**NETL Model with API/NPRA Cases**

| | Total Feed | Straight Run | SR S | LCO | LCO S | Coker Diesel | CO S |
|---|---|---|---|---|---|---|---|
| Feed | BPSD | BPSD | wt % | BPSD | wt % | BPSD | wt % |
| Reference | 25000 | 17250 | 0.5 | 5750 | 1.65 | 2000 | 2.2 |
| 50% SR | 25000 | 12500 | 0.5 | 10000 | 1.65 | 2500 | 2.2 |

| | On-Site | On-Site | Total | NETL Total | UOP | ENSYS Total |
|---|---|---|---|---|---|---|
| | Capital | Capital | Capital | Capital | Capital | Capital |
| Feed | $Million | $/BPSD | $Million | $/BPSD | $/BPSD | $/BPSD |
| Reference | 23.3 | 930 | 33.79 | 1352 | 1960 | $ 2,380 |
| 50% SR | 24.9 | 970 | 36.12 | 1445 | 2070 | $ 3,110 |

## References

1. B. Bjorkland, et al, (UOP), "The Lower it Goes, the Tougher it Gets (The Practical Implications of Producing Ultra-Low Sulfur Diesel)," paper AM-00-16 presented at the NPRA Annual Meeting, San Antonio, Texas, March 2000.

2. M. Hunter (KBR, Inc), "MAKFining Premium Distillates Technology: The Future of Distillate Upgrading," paper AM-00-18 presented at the NPRA Annual Meeting, San Antonio, Texas, March 2000.

3. E. Bingham, et al, (Haldor Topsøe, Inc.), "Improved Reactor Internals for Syncrude's HGO Hydrotreaters," paper AM-00-19 presented at the NPRA Annual Meeting, San Antonio, Texas, March 2000.

4. J.L. Nocca, et al, (IFP), "Cost-Effective Technologies for 21st Century Gasoline and Diesel Fuel," Hart Fuels Conference, Washington, DC, September 1998.

5. T. Tippett, et al, (Haldor Topsøe), "Ultra Low Sulfur Diesel: Catalyst and Process Options," paper AM-99-06 presented at the NPRA Annual Meeting, San Antonio, Texas, March 1999.

6. R.van der Meij, (Akzo Nobel Catalysts), "Meeting the Coming Sulfur Regulations in Diesel," Symposium on Diesel Desulfurization sponsored by NPRA and API, Houston, Texas, February 2, 1999.

7. T. Heckel, et al, (UOP), "Technology Update," Symposium on Diesel Desulfurization Sponsored by NPRA and API, Houston, Texas, February 3, 1999.

8. R.E. Palmer, et al, (Mustang Engineers and Constructors), "Revamp Your Hydrotreater to Manufacture Ultra Low Sulfur Diesel Fuel", paper AM-01-22 presented at the NPRA Annual Meeting, San Antonio, Texas, March 19, 2001.

9. R.O. Koseoglu, et al, (IFP), "Distillate Hydrotreating Routes: from Deep HDS to Cetane Number Improvement", Upgrading Heavy Ends with IFP, pp 75-82, 1997.