UPGRADER OPTIMIZATION

A NEW LOOK AT THE REGIONAL UPGRADER
WITH SELECTED ADDITIONS

ALBERTA CHAMBER OF RESOURCES
OIL SANDS TASK FORCE
WITH
ALBERTA DEPARTMENT OF ENERGY
AND
ENERGY MINES AND RESOURCES CANADA

RSN 233-942-805

Alberta Energy

96-03611/

ACKNOWLEDGEMENT AND DISCLAIMER

The research project for which this report is submitted was funded (in part) from the Alberta Department of Energy.

This report and its contents, the project in respect of which it is submitted and the conclusion and recommendations arising from it do not necessarily reflect the view of the Government of Alberta, its officers, employees or agents.

The Government of Alberta, its officers, employees or agents and consultants make no warranty, express or implied, representation or otherwise, in respect of this report or its contents.

The Government of Alberta, its officers, employees and agents and Consultants are exempted, excluded and absolved from all liability for damage for injury, howsoever caused, to any person in connection with or arising out of the use by that person for any purpose of this report or its contents.

TABLE OF CONTENTS

TABLE OF CONTENTS

TABLE OF CONTENTS

TABLE OF CONTENTS

TABLE OF CONTENTS

LIST OF TABLES

LIST OF TABLES

LIST OF FIGURES

ACKNOWLEDGEMENTS

The Technical Advisory Board provided appreciable advice and much assistance at the personal level. Bert Lang as the Alberta Chamber of Resources' Project Manager also provided many valuable comments and timely approvals. Manuel Torres of the Alberta Department of Energy and Mr. Bill Dawson of CANMET also provided many inputs as internal approvals. The CIBC's Rob Francis handled the major billing and fiscal management role very capably and the Alberta Chamber of Resources Don Currie provided administrative assistance.

AOSTRA's Roger Bailey and Riaz Padamsey were always on call and of appreciable assistance, and the development of "upgraded" Cold Lake upgrader yield data by the latter was very much appreciated. Purvin and Gertz' Tom Wise was also of much value beyond the provision of value forecasts and Dennis Bobiy provided a number of comments beyond his brief report on alternate construction approaches.

Mobil's Grant Karsner provided many useful comment on natural gas conversion and Mobil processes, including middle distillate dewaxing.

EXECUTIVE SUMMARY

1. INTRODUCTION

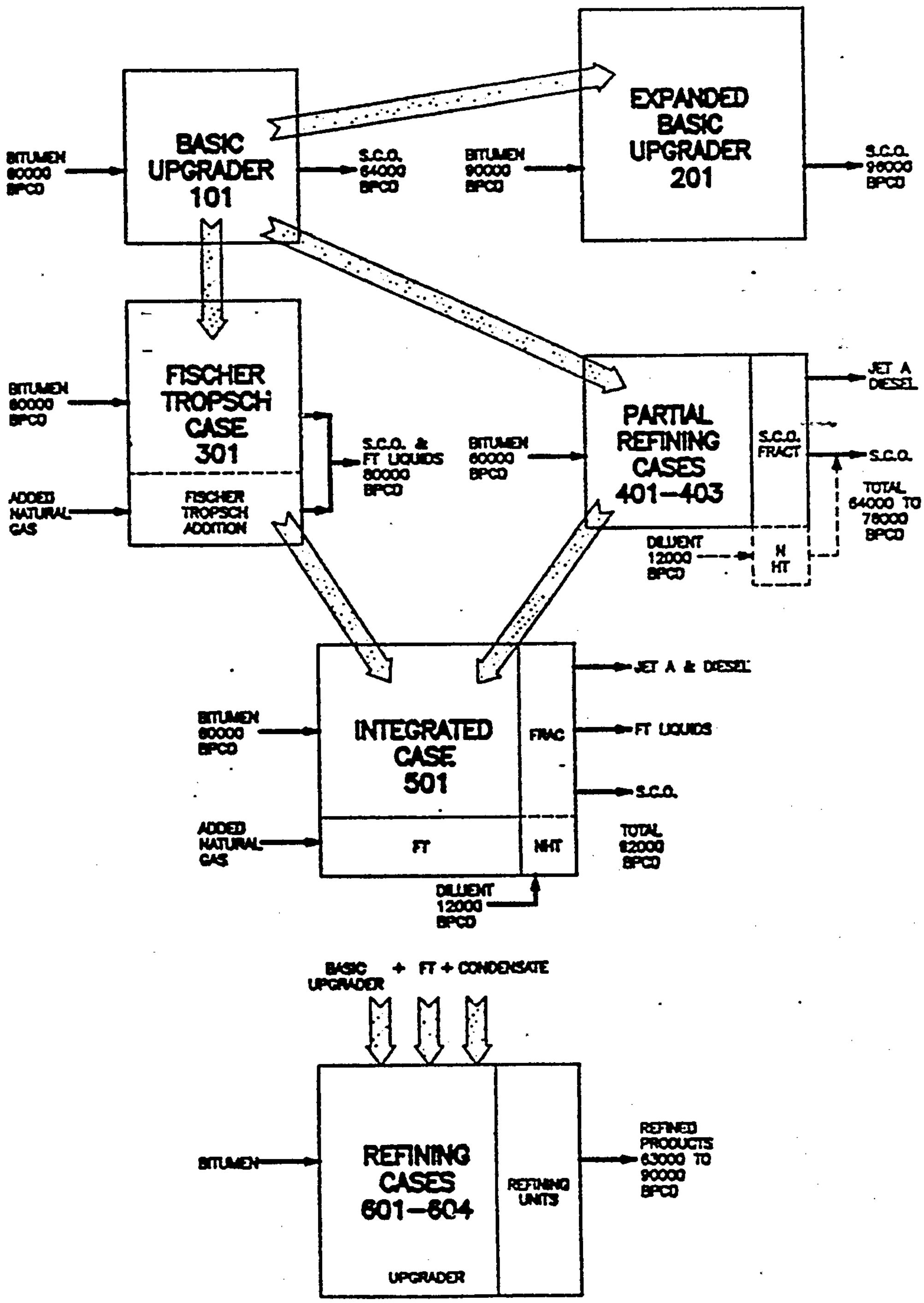
This study investigated a series of alternates to improving upgrader economics through adding onsite processes and in one case by expanding the Basic Upgrader from 60,000 to 90,000 BPCD. The Diverse Interests case upgrader of the 1990 Regional Upgrader Business Plan Study of the Alberta Chamber of Resources' Oil Sands Task Force, was used for the Base Case. That plant is based on a generic high conversion, high hydrogen addition primary upgrading plus fully integrated secondary hydrotreating. The Basic Upgrader would convert Cold Lake and Athabasca bitumen to a premium synthetic crude oil, with all "add-ons" producing readily merchantable products.

This current study has been funded by the Alberta Department of Energy, in part via its Hydrogen Research Program; by the federal Department of Energy, Mines and Resources; and by the Oil Sands Task Force member companies - Amoco, Canadian Occidental, Husky, Imperial, Shell and Suncor.

Product returns, and capital and operating costs were reforecast from the 1990 report or adjusted to first quarter 1993 (1Q93) costs. Cases were compared on a before tax net present value basis, using a 10% discount factor over the 28 year life of the upgrader.

The various cases considered are outlined on the first diagram. The second and third diagrams illustrate the incremental R.O.I.'s compared to the Base Case - neglecting taxes and inflation. It also assumes the prices forecast in Table 1, largely by Purvin and Gertz.

Figure 1



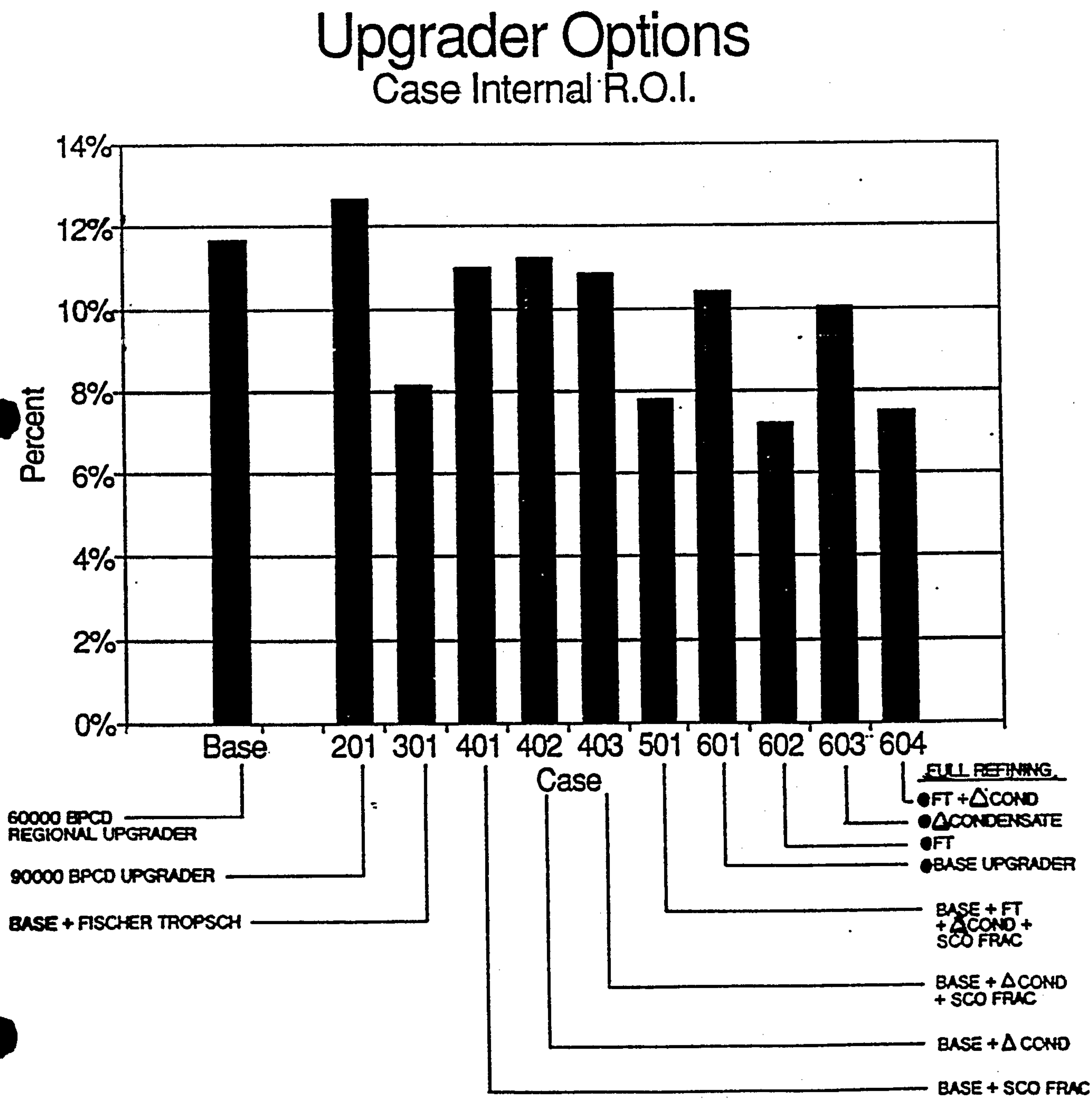
OVERVIEW OF CASES

Table 1
Edmonton Cost/Netbacks
In Constant 1Q93 Canadian Dollars

Stream	1Q93 Value ($/bbl)	2000 Forecast Value ($/bbl)	2010 Forecast Value ($/bbl)	Basis
Dubai FOB Mid East	(17.00 U.S.)	(18.01 U.S.)	(21.28 U.S.)	World Crude Reference P & G
Alberta Light Sweet Crude	24.91	27.65	32.58	Purvin & Gertz
Raw Bitumen	13.91	15.51	19.81	Purvin & Gertz
Diluted Bitumen	17.58	19.56	24.07	2/3 bitumen, 1/3 diluent for inventory use only
Diluent	24.91	27.65	32.58	Purvin & Gertz
Intermediates (in prim/sec upgrading)	18.27	20.06	24.57	Dilbit + 50¢ For inventory use only
Regional Upgrader S.C.O.	26.54	29.49	34.59	Purvin & Gertz
Regional Upgrader Components	26.54	29.49	34.59	For inventory use only
F-T Middle Distillate	35.70	38.07	42.79	See text - Section 3.2
F-T Naphtha	24.91	27.65	32.58	Industry estimate (equal light sweet crude)
Gasoline (Regular)	31.67	34.43	39.24	Purvin & Gertz
Jet A-1	31.96	34.55	39.25	Purvin & Gertz
Diesel (0.05% S) (40 Cetane)	31.51	33.87	38.60	Purvin & Gertz
Field Butanes	15.35	20.11	24.38	Purvin & Gertz
Propane	14.70	18.39	22.21	1990 study (no change)
Electricity - 0.9 plant service factor	3.1 ¢/kWh	3.1 ¢/kWh	3.1 ¢/kWh	Industry estimate (b)
Sulphur	50/tonne	50/tonne	50.00/tonne	1990 study (no change) per tonne
Natural Gas	[1.55]	[2.35]	3.26	Purvin & Gertz per million BTU
Byproduct Hydrogen	[1.56]	[1.80] per 1000 scf	Not calculated	See narrative; note maximum available is less than upgrader needs
Methanol	21.00	22.50	24.21	Industry estimate
Pitch	-38.00/tonne	-38.00/tonne	-38.00/tonne	(c)

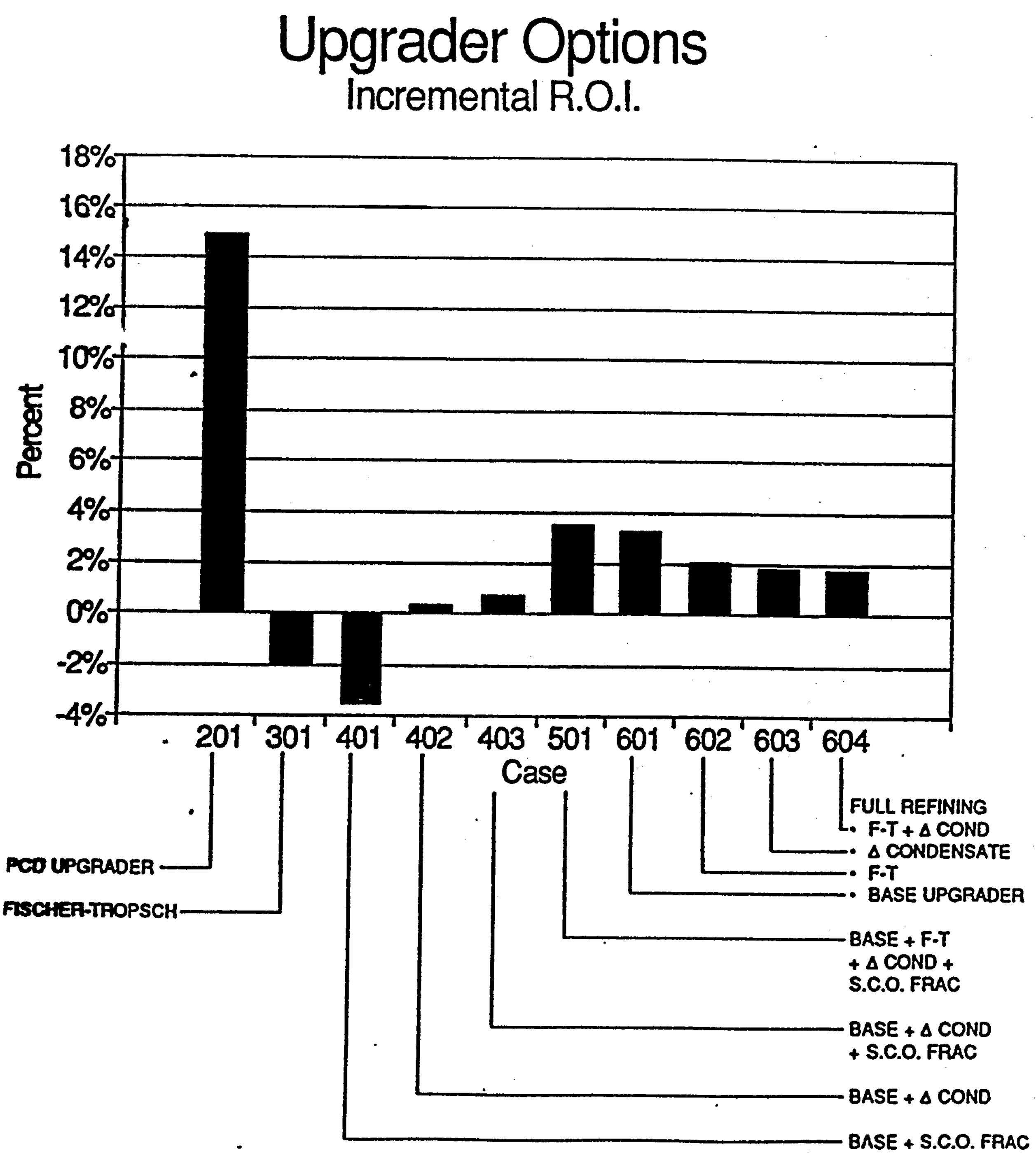
Notes: (a) Value over life of upgrader corrected to 1Q93 at 10% discount factor.
(b) While utility believe prices will decline in 1Q93 terms over the next four years, such is not assumed here.
(c) Disposal cost allocated to operating cost.

Figure 2



Upgrader Options
Case Internal R.O.I.
14%
12%
10%
8%
6%
4%
2%
0%
Percent
Base
201
301
401
402
403
501
601
602
603
604
Case
60000 BPCD REGIONAL UPGRADER
90000 BPCD UPGRADER
BASE + FISCHER TROPSCH
FULL REFINING
FT + Δ COND
Δ CONDENSATE
FT
BASE UPGRADER
BASE + FT + Δ COND + SCO FRAC
BASE + Δ COND + SCO FRAC
BASE + Δ COND
BASE + SCO FRAC

Figure 3


Upgrader Options
Incremental R.O.I.
18%
16%
14%
12%
10%
8%
6%
4%
2%
0%
-2%
-4%
Percent
201
301
401
402
403
501
601
602
603
604
Case
FULL REFINING
• F-T + Δ COND
• Δ CONDENSATE
• F-T
• BASE UPGRADER
PCD UPGRADER
FISCHER-TROPSCH
BASE + F-T
+ Δ COND +
S.C.O. FRAC
BASE + Δ COND
+ S.C.O. FRAC
BASE + Δ COND
BASE + S.C.O. FRAC

This study has only used publicly available data, hence, certain proprietary data may have been bypassed which improve economics, especially in the natural gas conversion cases.

2. EXPANDED UPGRADER

At 90,000 BPCD of feed, the Expanded Upgrader would use 3 conversion, hydrogen production and sulphur recovery trains, with other processes single train. At this size, the diluted bitumen distillation and following vacuum column would be the largest such units in Alberta with significant field vessel fabrication.

The Expanded Upgrader would produce 95,840 BPCD of a 35.7°API S.C.O., above light sweet par crude in 1993, the margin rising slightly with time. The S.C.O. from both 60,000 and 90,000 BPCD cases would find ready markets in northern tier states as well as in Canada (but pipeline contamination might drop the value in Chicago markets).

The additional capital for expansion to 90,000 BPCD is estimated at $685 million - $22,800 per BPCD of feedstock ($21.500 BPCD of S.C.O.)

As expected, the Expanded Upgrader shows lower operating and capital costs, the return on added capital being well above the Base Case facility.

3. NATURAL GAS CONVERSION

Earlier studies had indicated a technical fit for the addition to an upgrader of natural gas conversion via the Fischer-Tropsch synthesis route. While adding about 25 percent to the upgrader's liquid products, hydrogen would also be produced for upgrading.

Addition of such a scheme is estimated to add $1,050 million in capital or $62,700/BPCD of incremental products (16,740 BPCD). The product value averages above S.C.O. due to a premium over diesel expected for the 9,040 BPCD of middle distillate that can be blended with about 15,000 BPCD of the same fraction of current quality S.C.O. to improve the latter's cetane number to a 43 level as need by most light crude refineries.

The F-T naphtha will receive approximately light par crude value for petrochemicals, but is a very poor refinery feedstock. The F-T add-on does not appear particularly attractive economically. An alternate approach of using partial oxidation and purchased oxygen to convert natural gas only for F-T feed - with a parallel conventional natural gas to hydrogen unit for upgrader hydrogen - appears at least equally viable. But in such a situation, F-T is not particularly synergistic with upgrading unless the premium qualities of the F-T middle distillate are essential in the S.C.O.; something not now foreseen.

Synthesis gas production and natural gas conversion both appear areas where improved and/or preferably new technologies are needed.

4. PARTIAL REFINING

The addition of an S.C.O. fractionator was explored with production of 6,000 and 9,000 BPCD of jet fuel and diesel, with the potential of producing the rest of the upgrader's S.C.O. in various types, differentiated by fractional composition.

The direct jet and diesel sales actually improve the marketability of the rest of the S.C.O. The addition of 12,000 BPCD of diluent was also considered to provide an even more acceptable S.C.O. - one that has enough naphtha to be considered for a refinery's basic crude oil.

The economics of naphtha addition to the S.C.O. are not particularly attractive due to the $95 million capital cost being offset by only another $2 million in revenues.

The added offplot piperacks, tankage, etc., all tend to greatly inflate the cost above the bare unit cost in both these sub cases.

The economics of such additions are not apparent unless one assumes that the upgrader can consistently receive a higher return from differentiated S.C.O. products than from a single product (as Suncor are claiming) - approximately $21 million more a year for each added dollar per barrel. With such an increase (or equivalent prevention of such a reduction from perceived value) the S.C.O. fractionator at least appears attractive.

The study concludes that differentiated S.C.O.'s and/or specific products should be further explored by an upgrader proponent but possibly with fractional desegregation in the secondary hydrotreating system.

5. INTEGRATED CASE

This case assumed F-T plus S.C.O. fractionation plus added diluent. As the individual cases leading to this case were not particularly attractive and there is little synergy between F-T - the most expensive add-on - and upgrading, this case is not discussed further here.

6. FULL REFINING CASES

These cases were added towards the end of the study to test the viability of full refining IF markets can be developed for the gasoline. Middle distillate demands are expected to continue to increase in both Canada and the U.S. with markets for the upgrader/refinery's output. However, sufficient gasoline production capability appears to exist in all but accessible markets but possibly western Canada for the foreseeable future. Some refineries will have to adapt to reformulated fuels but this will be at a much lesser expense than a new refinery. But with the addition of MTBE the gasoline products of the scheme developed here will meet probable U.S. national standards and reformulated qualities (but olefins will be above California and New England standards).

A relatively conventional refining scheme based on catalytic cracking, alkylation, catalytic reforming and isomerization is assumed, but with an added TAME unit. The latter will convert high vapour pressure C_5 olefins (smog reactive species) and purchased methanol to a premium octane, low vapour pressure component providing some oxygen to the product.

The refining scheme designed to process all S.C.O. will cost an incremental $660 million in the case with 12,000 BPCD of diluent added. The difference between product sales, assuming a minimum gasoline approach and feedstocks now including small amounts of butanes and methanol, rises by $111 million compared to the Basic Upgrader. After increases in operating costs the margin drops to about $66 million a year in 1Q93 terms. None of the refining cases sparkle economically.

If an added $2/bbl can be attained for the gasoline, say in penetrating U.S. reformulated gasoline markets, the return is only about 11.2% versus 10.6% at the base gasoline price.

7. BUY OR MAKE HYDROGEN

A brief side study revisited this topic from the 1990 study and concluded that there is merit in further consideration of purchase of, say, 70 percent of the upgraders' needs. But there are still a number of supply security risks to be assessed.

8. CONCLUSIONS

This study has not identified any breakthroughs. Of all alternates considered, only two appear to warrant detailed inspection - S.C.O. fractionation (with or without diluent/condensate addition) to improve/guarantee good product prices and full refining, the latter only if gasoline markets can be firmed up.

The Fischer-Tropsch natural gas conversion route suffers from very high capital costs and does not appear appropriate even with a lower cost partial oxidation approach. There are major research and development opportunities in F-T and natural gas conversion generally.

DISCLAIMER

The data, opinions and conclusions advanced in this report are those of the authors and are not necessarily in accord with the views and/or policies of the government of Alberta, Energy, Mines and Resources Canada and/or the Alberta Chamber of Resources.

1.0 INTRODUCTION

1.1 PREAMBLE

The report summarizes a study into a variety of alternate approaches that may increase the financial return of bitumen upgrading. The 1990 Oil Sands Task Force Regional Upgrader Business Plan's "Diverse Interests", 60,000 BPCD, ultra high conversion, high hydrogen addition route has been used as the Base Case throughout this study.

The "optimization" in the report's title is a misnomer to the extent that none of the schemes presented here were fully optimized - indeed only the operator of a specific project can do that - rather this study provides clues and directions as to some alternate routes to be considered by future upgrader proponents.

English units have been used in this report to be consistent with the 1990 Business Plan.

1.2 STUDY ORGANIZATION

This study has been funded by the Alberta Department of Energy, the federal Department of Energy, Mines and Resources and the following oil companies: Amoco, Canadian Occidental, Husky, Imperial, Shell and Suncor. The Alberta Chamber of Resources' Oil Sands Task Force was the study's manager.

The study has been under the general direction of a management committee consisting of Mr. Bert Lang of Suncor as Chairman, Mr. Manuel Torres of the Alberta Department of Energy and Mr. Bill Dawson of the Department of Energy, Mines and Resources. Erdal Yildirim of Canadian Occidental was the driving force behind the earlier work and provided overview of this study. Don Currie of the Alberta Chamber of Resources (ACR) and Robert Francis of the CIBC provided the administrative and financial management functions.

A Technical Advisory Board of the ACR's Oil Sands Task Force provided technical overview and many contributions throughout the study.

This study was coordinated by Stanley Industrial Consultants Ltd.'s (SICL) T.J. McCann with D. Tuli on hydrogen and synthesis gas production and F-T synthesis support facilities, and Kilborn's J. Jansen on capital cost estimating. SICL's P.H.S. Magee provided the refining and operating cost bases with D. Lubarsky on the fiscal models and R. Dingman on capital spreadsheets. Purvin and Gertz's T. Wise provided the vast majority of price forecasts. Energy International, under A. Singleton, provided the F-T synthesis process systems. D. Bobly briefly analyzed alternate construction approaches.

1.3 ECONOMIC BASES

The 1990 Business Plan provided a format for evaluation of alternate cases. Due to the number of alternates being considered, a simple net present value approach and internal return on investment is used, neglecting inflation and taxes but allowing for expected changes over a 28 year upgrader life.

1.4 TECHNOLOGY BASES

This study has used only data publicly available in order that it may be freely distributed. Thus, no proprietary data are included.

2.0 CASES CONSIDERED

2.1 PREAMBLE

The earlier OSTF studies indicated that the addition of natural gas conversion via Fischer-Tropsch synthesis to increase S.C.O., or S.C.O. equivalent, production was technically feasible with co-produced hydrogen balanced to upgrading needs. But was F-T economically viable? - the major question addressed in this study.

Suncor activities, public via late 1992/early 1993 paper, and analysis of the upgrader's S.C.O. fractional composition indicated that the more or less desegregation of the single S.C.O. of the 1990 report might be economic.

Evaluation of the S.C.O. composition also noted that the addition of naphtha would probably aid in marketing the S.C.O., particularly as it would more closely mimic light sweet crude oil in refineries designed for such crudes, allowing use of the modified S.C.O. as a basic rather than an incremental feedstock.

In early March of 1993, Imperial Oil's M. Ghosh presented a concept for splitting diluted bitumen into a heavy vacuum bottoms fuel fraction (for emulsified fuel use) and a diluent/bitumen tops blend, noting interest by 2 refiners, at least. In effect, naphtha is added to the "S.C.O." product. While diluent will probably be in short supply by 2000, at least one-third should be available for addition to S.C.O. when bitumen otherwise moving to market is upgraded.

The question of economics of full refining as opposed to merely producing an S.C.O. for conversion to finished products elsewhere, has been an ongoing question and is the last one addressed.

It must be noted that the term upgrader optimization has been used in the hope that upgrading economics can be improved by adding/revising process steps - what are the most appropriate ones? Only an upgrader proponent can truly answer the questions raised, but this study should provide some directions.

The S.C.O. product specification and expected S.C.O. yields in the Base and Expanded Base Cases are shown in Table 2.1-1.

2.2 BASE CASE

The 60,000 BPCD "Diverse Interests Case from the 1990 Regional Upgrader Business Plan was selected as the Base Case for this study. Minor changes in product handling were made to suit other cases, otherwise the original concepts were untouched.

New hydrogen unit costs were developed from a specific process design in order to be fully consistent with all other cases, but the basic design concept was unchanged.

Sufficient diluted bitumen has been assumed available from both Cold Lake and Athabasca sources, with diluent returned to the producing field.

A product pipeline to a new Edmonton terminal was added with provision for S.C.O. product movement to all three Ft. Saskatchewan/Edmonton area refineries, as well as to refineries on the west coast via the TransMountain System and Ontario and mid west refineries via the InterProvincial systems in batches up to 300,000 barrels.

Table 2.1-1
Light/Synthetic Crude Quality Comparison

Factor		ACR Synthetic Target[1]	ACR Estimated*	Current Quality Synthetic Crude Oil▲	IPL Blend (Alberta Light Par)▲
S.C.O.					
• Gravity, °API		30 (min)	36.7	32.8	38.0
• Sulphur, wt %			<0.01	0.17	0.5
Distillation, LV %					
• C_3 and lighter		3 (max)	0	2-4	1
• C_4's		15-0	2.9	5	1-3
• C_5 - 71°C			—	14	6.5
• C_5 - 177°C (350°F)			17.7	28	27.5
• 71 - 193°C					17
• 177 - 260°C				16	11
• 177 - 343°C		40-45	46.8		
• 204 - 343°C		25-40	41.2		
• 343 - 524°C			36.7		
• 343 - 566°C			N/A/	33	26
• 566°C +			0.0	1	9
Properties					
• C_5 - 71°C	Octane, (R+M)/2		60	60	65
• C_5 - 177°C	Nitrogen, wppm	1	<0.5		
• 71 - 177°C	N+2A, LV %			70	72
	N+2A, wt %		61		
• 177 - 260°C	Aromatics, LV%	22 (max)	<18		
	Smoke Point, mm	20 (max)	21		
• 193 - 288°C	Aromatics, LV %			38	22
	Smoke Point, mm			18	24
	Freeze Point, °F			-67	-31
• 193 - 343°C	Sulphur, wt %			0.04	0.25
	Cetane Number			41	50
	Aromatics, LV %			45	26
	Pour Point, °F			-45	-10
• 177 - 343°C	Cetane Number	43 (min)*	44		
	Sulphur, wppm	500 (max)	<5		
• 343 - 566°C	Sulphur, wt %			0.34	1.0
	Nitrogen, wppm			1400	<1200
	Gravity, °API			18	24
	"K" Factor			11.4	11.9
	Aromatics, LV %			60	40
	Polycyclic Aromatics			33	15
	Carbon/Hydrogen, wt %			7.3	6.9
• 343 - 524°C	Sulphur, ppm		100		
	Nitrogen, wppm	1000 (max)	<100		
	Gravity, °API		25.7		
	Carbon/Hydrogen, wt		7.1		

Notes:

* Revised 1993.

▲ Supplied by Purvin and Gertz, "Synthetic Crude Market Analysis", January, 1990, of 1990 Report.

The S.C.O. is expected to find a ready home in light sweet crude oriented refineries served by InterProvincial, particularly at St. Paul, Superior, Sarnia, Mississauga, and Nanticoke, where pipeline contamination is minimal compared with that to refineries served by the looped system via Chicago.

The prior studies indicated that the S.C.O.'s diesel fraction would have a cetane of 43.5 to 44.5 and a recommendation appeared there to raise the S.C.O. cetane specifications to 43 to allow refiners to meet a 40 cetane product when low grade streams were added. Thus, the S.C.O. specification was adjusted to 43 (from 40).

2.3 EXPANDED BASE CASE

As total product output can reach as high as 92,000 BPCD level in other cases, this case considered an upgrader 150 percent of the Base Case - 90,000 BPCD of bitumen and approximately 96,000 BPCD of product.

No change in process or auxiliary system configuration from the Base Case has been made. The same markets are attainable, given the ongoing decline in Canadian light sweet crude production.

2.4 FISCHER-TROPSCH CASE

In this case the natural gas demand increases by a factor of slightly less than 5 to provide sufficient hydrogen for the Base Case Upgrader and synthesis gas (2 hydrogen plus 1 carbon monoxide) to produce approximately 16,000 BPCD of Fischer-Tropsch naphtha and middle distillates (Jet A and diesel equivalents).

The F-T naphtha will probably be sold to a Sarnia ethylene producer at light crude value and the midule distillates would be used to enhance the qualities of similar fractions from conventional synthetic crude oils providing blends that meet full Jet A (kerosene type) and diesel product specifications. Alternately, the F-T middle distillate could be used at any refinery with smoke and/or cetane problems with heavy crude based U.S. refineries very specific targets.

The F-T Case is based on a fixed bed F-T synthesis design provided, insofar as proprietary constraints would allow, by Energy International whose staff has long experience in F-T synthesis process development. A promoted cobalt catalyst in fixed bed (tubular) reactors was selected as most appropriate for this case. The F-T raw liquid product will be converted into the premium middle distillates - very high smoke point and cetane (but low densities) - in a hydrocracking/dewaxing product finishing section.

The bulk of the overall study has been relative to the entire F-T system as full balances and equipment sizes were needed for cost estimating. Special emphasis was placed on synthesis gas production - steam methane reforming, CO_2 capture and hydrogen recovery (for upgrading) - because of its major impact on capital and operating costs. As noted, EI provided the design of the F-T synthesis system. The F-T product finishing costs have been based on literature and file data - with process balances being preliminary due to lack of public data.

2.5 PARTIAL REFINING CASE

There are two sub cases - one with fractionation only and the other with naphtha added via diluent addition.

Evaluations of light sweet crude consuming refineries long the U.S. northern tier and in Ontario, indicated that at least 10 percent of the original S.C.O. could be sold as middle distillate products leaving a continuing high value S.C.O. In practice, the percentage might well rise as high as 20 percent. In order to separate out such products, a full sized S.C.O.

fractionation system was added to the Basic Upgrader - the full size allowing sale of all S.C.O. as specially blended S.C.O.'s to suit specific refiner needs (and to achieve the maximum average S.C.O. return as Suncor are showing).

As noted previously, additional naphtha would enhance the value and more particularly, potential continuous sales to most light sweet crude based refineries. Thus, a sub case examines adding 12,000 BPCD of diluent to the upgrader's feed. This added naphtha necessitates an added hydrotreater whose design is set for a very sour condensate, should such ever be selected as the incremental feedstock (although not assumed here).

2.6 REFINING CASES

Four possible refining cases were considered based on the Basic Case Upgrader, with no incremental feeds, with F-T added, with diluent added and with both F-T and diluent facilities added.

A relatively conventional refining configuration was selected - catalytic cracking, C_3C_4 alkylation, C_5C_6 isomerization and C_7 to C_9 or C_{10} catalytic reforming. The catalytic cracking unit shows special synergy in the primary and secondary upgrading units, eliminating any heavy fuel oil production.

However, in order to meet low summer time gasoline vapour pressure specifications, a unit to reduce C_5 olefin production to a minimum became necessary. A Tertiary Amyl Methyl Ether route with C_5 olefin isomerization was selected. This results in some oxygen in the product gasolines as well as significantly reduced olefin levels. With addition of MTBE (over the fence at the Edmonton terminal end), the upgrader/refinery's gasolines will probably meet U.S. national reformulated standards (but be slightly above California and New England olefin levels).

The TAME unit introduces methanol as a "new" feedstock. A butane isomerization facility was also added to reduce the cost of n-butane for alkylation (as at the 2 Edmonton refineries).

The refined products can move to Edmonton markets (via existing loading racks); to Calgary (via the APPL pipeline); to Kamloops, Vancouver and Washington refineries (via TransMountain); and to eastern prairie, U.S. northern tier and Ontario refineries (via InterProvincial). Michigan/Ohio markets are also accessible via the Buckeye system from InterProvincial at Marysville, just upstream of Sarnia.

At this time, Chicago area markets appear doubtful due to pipeline contamination questions. While refined product prices were netted back to Edmonton from Chicago, this basis is still applicable to other mid western markets. (Jet fuel movements to Vancouver and Sarnia also suspect relative to contamination, and may necessitate redistillation in existing facilities.)

3.0 FEED, INTERMEDIATE AND PRODUCT PRICING

3.1. PRICING AND COST SUMMARY

The pricing used in this study is shown in Appendix A which contains the Purvin and Gertz report with year by year projections for most in and out streams (in U.S. 1993 dollars) as well as a comparison in Canadian dollars. Their 1990 report contributions should be referred to in the 1990 Business Plan as it discusses S.C.O. valuation in some detail. The Canadian dollar has been valued at $0.80 U.S. throughout this study.

For this study, Purvin and Gertz have provided a natural gas price forecast. The major change in natural gas cost over the forecast period must be noted as it materially impacts on upgrader economics in future years. The projected minor strengthening in refinery margins is also of note, but not enough to offset natural gas.

The refined produced prices have been based on Purvin and Gertz estimates of expected netbacks at Edmonton for sales in Chicago. While Chicago area sales now appear unlikely, the use of Chicago as a basing point is reasonable and appropriate for all but movements to the west coast.

In order to simplify analysis of Full Refining cases in this study, only a generic Canadian quality, regular gasoline has been used. The quality of the gasoline will be close to that expected in late 1990's U.S. reformulated gasoline - only MTBE need be added. The refining case configurations all have the capability to produce an 89 road octane pool to achieve 87 for regular and up to 93 for premium (before added MTBE providing up to 1.5 numbers additional). However, most sales are envisaged to non refining marketers who sell mostly regular grades and seek the lowest price product supply at all times. Gasoline production capacity is not expected to become particularly tight in any region except perhaps western Canada through the forecast period. Thus, while there may be a negative bias in gasoline pricing, it is not expected to be serious.

The diesel pricing is based on a generic 40 cetane number low sulphur blend to meet late 1993 U.S. standards. There is a likelihood that by 2000 cetane level could go to 45 minimum, above the upgrader's current capabilities in all but cases with F-T middle distillate added. The latter's pricing is discussed in Section 3.3 below.

The electricity estimate is based on current utility charges and an assumed no charge in 1993 dollar terms in the future. (Utilities predict a slight fall in constant dollars not shown here.)

3.2 ECONOMIC ANALYSES BASES

The data of Table 1 and Appendix A show rising feedstock and product pricing through 2010 in 1993 dollars. A rising differential between products and feedstock are predicted, more so relative to S.C.O. than to refined products as the premium qualities of the S.C.O. become of more value to refiners. But through the forecast period, natural gas cost is predicted to rise at an even greater rate.

To provide estimates of the value of the various "add-ons", this study has used the pricing forecast provided by Purvin and Gertz and the capital cost estimates detailed in Appendix B (extended to cover the project life). The capital expenditure profile, planning/engineering and construction schedules were assumed unchanged from the 1990 basis. Each case considered by this study, including the economic sensitivity cases discussed in Section 11, were compared by the following measures:

1) Simple ratio of operating margin (revenues minus feedstocks and operating costs) to capital cost based on 1993 prices.

2) Net present value of the operating margin cash flow (before taxes, no inflation or special financing) over the project life, using a 10 percent annual discount rate and forecast pricing.

3) Internal rate of return of the investment based on the operating margin cash flow.

The Appendices contain additional data for a reader desiring to do more detailed calculation than here.

3.3 COMMENT

The impact of taxation, innovative project financing, special discounts, etc., will change the economic attractiveness of the project. This study was primarily concerned with evaluating and selecting the best technical "fits" and "add-ons" to the Basic Upgrader, thus more complex economic analysis was not carried out. Further economic analysis of the selected process add-ons package is strongly recommended.

3.4 F-T MIDDLE DISTILLATE VALUATION

The F-T middle distillate will be used to improve the quality of refinery jet and diesel fuels from smoke and cetane number viewpoints, respectively. In each such use, F-T middle distillate will be reducing aromatics through dilution. The F-T middle distillate is extremely light (low specific gravity), with a volumetric energy content well below that of conventional jet and diesel fuels, thus, it is hard to see it selling directly to customers; even to city bus companies who are very concerned with visible diesel exhausts (which will be greatly reduced with the total lack of aromatics in the F-T material).

To place a value on these middle distillates for all but the Full Refining cases (where they will be used to enhance the diesel pool to over 47 cetane number), blending with low cetane conventional synthetic crude oil middle distillate to produce a refinery acceptable cetane blend, is compared to the addition of an aromatic saturation unit.

Assuming 33 cetane for the conventional S.C.O. middle distillate (CSMD), 60 blending cetane for the F-T middle distillate (F-TMD) and 43 for the refinery blend (RB), indicates that the 9,000 BPCD of F-TMD will blend with 15,300 BPCD of CSMD to produce 24,300 BPCD of RB.

A unit to produce that much RB from CSMD will require about 23,100 BPCD of feed, assuming a net volumetric gain of 5 percent. Current aromatic saturation technology - e.g. the Criterion/CE Lummus SynSat process - would require a deep, higher temperature desulphurization step before a colder precious metal aromatic saturation step. Such a unit will cost approximately $70 million in the Edmonton area (using estimating consistent with other parts of this study). Operating costs per barrel of product will be approximately $1.75 - $1.20 for hydrogen (at $1.00 per thousand scf) and about 11 cents each for fuel and electricity, catalyst, maintenance, and incremental other operating costs. This cost will be offset by approximately $1.60 due to yield gain. Thus, the net cost will be largely capital related. For simplicity, we have assumed a 20 percent return before tax as required by the refiner (or upgrader) - approximately $1.66. Thus, the overall net charge will be roughly $1.81 in 1Q93, placing the CSMD that much below the diesel market price (after crude fractionation).

Converting the $1.66 figure to a premium over diesel market value indicates a value of $35.70 (1Q93). Hydrogen costs will rise by 2000 due to higher natural gas costs, thus, the differential over diesel has been kept constant.

3.5 PURCHASED HYDROGEN

3.5.1 AVAILABILITY

The sensitivity of overall costs to purchase of part of the hydrogen needed in upgrading is discussed in a separate section, but here the cost of purchased hydrogen is developed. In all cases with F-T, except for a sensitivity case involving natural gas partial oxidation in lieu of steam methane reforming, the F-T systems SMR units produce the upgrader's hydrogen directly and thus the F-T route is not amenable to purchased hydrogen unless the F-T system size is reduced.[1]

The 1990 report indicated the availability of hydrogen from the following major sources by 1998:

Table 3.5.1-1
Regional Byproduct Hydrogen Availability

Location	Plant	Source Unit	Pure H_2 Available after Purification (mmscfd)	Notes
E. Edmonton	Alberta Envirofuels	Butane Dehydrogenation	12	MTBE Plant
N.E. Edmonton	Celanese	Methanol	61	
Ft. Saskatchewan	Dow	Chlor Alkali	9	
Ft. Saskatchewan	Dow	Ethylene	66	New 1994
Total			148	

1 In the FT cases, the hydrogen for FT product finishing is taken from purge gas processing, plus the margin provided over upgrader needs.

The Dow chlor alkali figure may be high by the year 2000 as chlorine is phased out of many uses, but two Bruderheim sodium chlorate plants already have about 5 million scfd available, half of which is now vented. It also appears that Alberta Envirofuels or and equivalent new plant will expand MTBE production and, hence, increases hydrogen supply potential. Dow's ethylene capacity could also expand. Cross ties to Petro-Canada's Edmonton and Shell's Scotford refinery hydrogen units would provide minor incremental supply and some on line balancing. Esso Chemical's Redwater fertilizer plant has a shutdown ammonia plant that can be converted to produce approximately 40 million scfd of hydrogen.

A maximum byproduct availability of about 150 million scfd can be reasonably assumed, but that availability will drop to 80 to 85 million with any outage at the largest producer.

3.5.2 BALANCE

The basic 60,000 BPCD regional upgrader requires 168 million scfd including a 6 percent safety margin, which is partly used in the various add-on facility cases. This supply is based on two 50 percent trains, except in the F-T Case where hydrogen is supplied from the front end of the F-T system when excess hydrogen over F-T needs is separated out for upgrader use.

If only 1 train is provided, the situation will be very tight whenever a major byproduct source is out of operation. Possibly storage will be needed to provide some surge capacity, also of benefit to on-site hydrogen production to avoid flaring hydrogen following a sudden drop in demand, and to allow faster upgrader throughput buildings. However, this study has not studied the system nor defined the risks and economics of storage.

As the single hydrogen unit must be kept on-line at, say, 60 percent of capacity to provide surety of supply, even with all sources available it is assumed here that over the year 70 percent of upgrader hydrogen needs will be purchased.

3.5.3 PURCHASED HYDROGEN COSTING

The byproduct hydrogen supply has the following cost sectors:

a) Replacement natural gas,
b) Margin for seller,
c) Facilities to purify and compress at source, and
d) Pipelines to Regional Upgrader.

- Replacement Natural Gas

Based on lower heating values and the above noted \$1.55 per million BTU (\$2.35) in natural gas in 1Q93, the replacement natural gas will cost \$0.47 per 1,000 scfd of hydrogen (\$0.71 in 2000 in 1Q93 dollars).

- Margin for Seller

This will be an item of appreciable negotiation and will be greatly impacted by the variability of offtake. The seller will require appreciably more instrumentation to take purification off gases and replacement natural gas into his fuel gas system. For example, burners now on nearly pure hydrogen must be able to use a range of compositions all the way to natural gas. This can get very expensive.

For this study, a seller's margin of 20 percent of the replacement fuel value has been assumed. This should not escalate with fuel gas price increases (in 1Q93 terms) - $0.09 per 1,000 scfd.

- Facilities for Purification and Compression

The 1990 report estimated these in 1989 dollars at $84 million (assuming an added allowance of $2.6 for chlor alkali hydrogen). This converts to approximately $90 million in 1Q93 costs.

Operating costs were estimated at $14 million (assuming $1.4 added for the chlor alkali stream). These costs will drop somewhat with below 100 percent loads, but the demand component of electrical costs will tend to keep operating costs relatively constant, regardless of average offtake.

Assuming 20 percent R.O.I. before tax return on these facilities indicates annual charges of about $32 million.

- Pipelines

The 1990 report costs related pipelines at $7.7 million. Adding in a variety of short connectors and converting to 1Q93 dollars, gives a total cost of about $10 million. Allowing for normal gas pipeline rate of return plus operating costs, gives an annual total cost of $2 million for the pipeline portion.

- Summary

Cost Sector	Annual (1Q93) Millions of Dollars
- Gas Replacement (168 million demand - 58 produced on-site) x 365 days x .9 F.S. x \$0.47/1000 scf x 1,000 scf	\$18.9
- Seller Margins	3.8
- Plant Site Facilities	32.0
- Pipeline Charges	2.0
Total	56.7 x 10^6

Note that all capital costs for the above system totalled about \$110 million, somewhat above that of the voided hydrogen unit. The unit cost of purchased hydrogen in 1Q93 thus works out to \$1.56 per 1000 scf.

In 2000, the price will rise to about \$1.80 (in 1Q93 dollars) due to rising gas costs.

4.0 BASE CASE

4.1 INTRODUCTION

In the 1990 Regional Upgrader Business Plan, the 30,000 BPCD case was labelled the "Reference Case" and the 60,000 BPCD case was titled "Diverse Interests Case". In this update and extension of that report, the "Base Case" is merely a 1Q93 version of the "Diverse Interests Case" - i.e. 60,000 BPCD of bitumen, with minor changes relative to product S.C.O. transport to match other cases.

AOSTRA developed an upgraded set of yields for a generic ultra high conversion primary plus secondary route, such as Veba's Combi Cracking or CANMET plus integral hydrocracking for a Cold Lake bitumen case. The quality of this data was considered preferable to that presented in the 1990 report for a 50/50 Cold Lake/Athabasca blend.

The battery limits of this Base Case (and all other cases of this study) include the primary upgrading units, the secondary upgrading units, the associated utilities and offsites to support these primary and secondary units, interconnecting facilities, related new pipelines as far as Edmonton and an Edmonton terminal.

The 1990 study cost estimating approaches were reviewed and a slight change was made to reflect the anticipated mode of construction, as well as to update its 4Q89 costs to the first quarter of 1993.

The Synthetic Crude Oil specification has been changed only to adjust the cetane number specification to 43 (from 40), the minimum as suggested by Purvin and Gertz in 1990 to allow refiners to blend in lower cetane stocks when meeting the normal 40 of diesel products. This has no effect on design or operations as 43.5 to 44.5 is anticipated in this Base Case, anyway.

In the following portions of this section, changes from the 1990 bases are underlined.

4.2 DESIGN BASES

The following are the design bases for the Base Case Regional Upgrader:

- Capacity

  The upgrader is designed to process 60,000 BPCD of bitumen.

- Crude Assay

  The upgrader is designed to process either Cold Lake bitumen (Table 4.2-1 repeated from the 1990 report), or Athabasca bitumen (Table 4.2-2 repeated from the 1990 report) or combination (separately or combined) of these two crudes. The "normal" feed to the upgrader consists of 50% Athabasca bitumen and 50% Cold Lake bitumen by volume. The crude is supplied to the upgrader as a blend with diluent in the following concentrations:

  - Athabasca 55%
  - Cold Lake 65%

- Synthetic Crude Quality

  The synthetic crude quality meets the specifications outlined above in Table 2.1-1. <u>(Note the increase in diesel cetane from the 1990 version.)</u>

Table 4.2-1
Crude Assay of Cold Lake Bitumen

FRACTION, F	C5-320	320-400	400-650	650-700	700-975	975+	TOTAL
GRAVITY							
API	57.8	37.1	25.2	19.6	16.1	2.5	10.8
SG	0.7475	0.8393	0.9030	0.9365	0.9587	1.0560	0.9944
MASS							
#/HR	4,580	18,360	124,836	38,512	212,272	471,520	870,080
DIST.	0.53%	2.11%	14.35%	4.43%	24.40%	54.19%	100.00%
VOLUME							
BPCD	420	1,500	9,480	2,820	15,183	30,619	60,022
DIST.	0.70%	2.50%	15.79%	4.70%	25.30%	51.01%	100.00%
COMP.							
CARBON	85.08%	86.16%	85.86%	85.30%	85.24%	82.78%	84.02%
HYDROGEN	13.54%	12.60%	12.50%	11.90%	11.00%	9.60%	10.54%
SULFUR	1.38%	1.24%	1.63%	2.70%	3.29%	5.95%	4.41%
NITROGEN			0.01%	0.03%	0.15%	0.68%	0.40%
OXYGEN				0.07%	0.32%	1.00%	0.62%
METALS, WPPM							
VANADIUM						277	150
NICKEL						101	55
ANILINE PT., F	120	120	130				
CCR, WT%						24.3	13.2
VISCOCITY, CS @ 210F						26400	73

1. ASH FREE BASIS - ASH CONTENT = 0.05%

28-May-90:R.PADAMSEY

Table 4.2.2
Crude Assay of Athabasca Bitumen

FRACTION, F	C5-320	320-400	400-650	650-700	700-975	975+	TOTAL
GRAVITY							
API	44.7	38.2	25.2	20.5	12.8	1.5	8.8
SG	0.8031	0.8338	0.9030	0.9309	0.9806	1.0639	1.0086
MASS							
#/HR	3,512	10,212	91,600	40,704	258,464	477,996	882,488
DIST.	0.40%	1.16%	10.38%	4.61%	29.29%	54.16%	100.00%
VOLUME							
BPCD	300	840	6,956	2,998	18,074	30,808	59,976
DIST.	0.50%	1.40%	11.60%	5.00%	30.14%	51.37%	100.00%
COMP.							
CARBON	84.51%	85.74%	85.90%	85.33%	85.05%	82.42%	83.73%
HYDROGEN	12.87%	12.61%	12.30%	11.80%	10.90%	9.42%	10.31%
SULFUR	2.62%	1.65%	1.74%	2.73%	3.77%	6.39%	4.90%
NITROGEN			0.01%	0.06%	0.16%	0.64%	0.40%
OXYGEN			0.06%	0.08%	0.12%	1.12%	0.65%
METALS, WPPM							
VANADIUM						460	250
NICKEL						138	75
ANILINE PT., F	93	127	117				
CCR, WT%						22.0	13.3
VISCOCITY, CS @ 210F							65.6

1. ASH FREE BASIS - ASH CONTENT = 0.7%

28-May-90:R.PADAMSEY

- Number of Independent Processing Trains

The primary and secondary upgrader design is based on two 50% capacity trains. The number of reactors is based on the maximum size of reactor transportable to site, being limited to 800 tons and a diameter of 14 feet. The hydrogen and sulphur units are also twinned at 50% and 75% capacity per train, respectively.

- Hydrogen Supply

The hydrogen plant design is based on steam methane reforming technology. The hydrogen purity is 99.5% (volume) or better, and the two 50% trains are each designed with 6% excess capacity based on upgrader processing of normal feed. Figure 4.2-1 outlines a basic hydrogen unit, developed specifically for this study, to be fully consistent with the Fischer-Tropsch Case's steam methane reformers.

- Residue Disposal

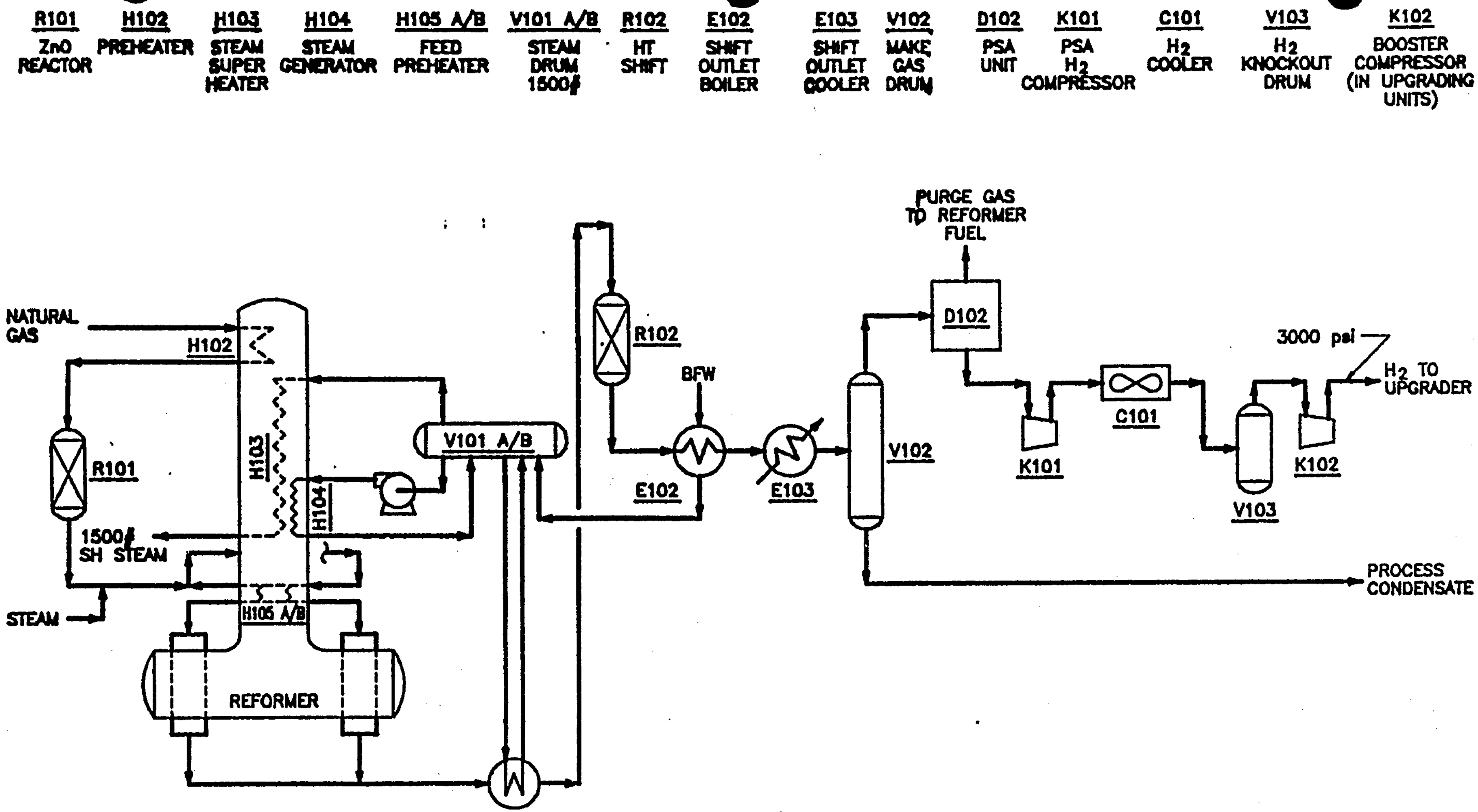
The residue, due to the small quantity produced, is assumed disposed of at a remote landfill site.

- Tankage and Pipeline

Table 4.3-3 below sets out the tankage of this Base Case, and Figure 4.3-1 below provides an outline of the connecting pipelines.

- Sulphur Plant

The sulphur plants are designed to recover 98.7% of sulphur contained in the feed. (The slight change upwards from 1990 is to match ERCB requirements for gas plants.)



BASE CASE
SMR PROCESS FLOWSHEET
FIG 4.2-1

- Steam and Power Plants

  The steam and power plants are designed to meet all the steam and boiler water requirements of the upgrader complex without any export or import. Power generation is minimized to match utilization of the excess high pressure steam.

- Location/Site Conditions

  The upgrader complex is located 20 miles northeast of the Edmonton city limits, and has the following infrastructures and services:

	From Site
- North Saskatchewan River	5 km
- Feed and Product Pipelines	3 and 20 km
- Electrical Transmission Lines	1 km
- Natural Gas Supply Line	1 km
- Rail Line	1 km
- Major Road	1 km
- Phone Line	5 km

  The site is assumed to be fairly level "farm land". For simplicity in defining the site, a site to the east of Shell's Scotford styrene plant has been used as shown in the hydrogen byproduct study of the 1990 report - the changes in pipeline distances relates to this siting and the addition of local feed and product pipelines.

4.3 DESCRIPTION OF OVERALL DESIGN

The processing schemes developed for each of the upgrading technologies considered in 1989/1990 are not repeated here.

The estimated overall yields and product quantities are shown in Table 4.3-1 for Cold Lake bitumen only.

Differences Between Cold Lake and Athabasca Crude

In the 1990 report, Veba presented only 50/50 blend data but the other two licensors presented cases for Athabasca and Cold Lake and 50/50 blends. Both of the latter showed more gas oil and less naphtha with Athabasca feed but with differences less than 3 percent for any given fraction. While CANMET showed a higher overall yield (1.5 volume percent) - with added hydrogen and additive - on Athabasca compared to Cold Lake, H-Oil showed the converse - a 1.5 decrease. The latter process is more impacted by feed qualities than are the non catalytic hydrogen addition routes.

There will be significant hydrocracking in the secondary hydrotreating of the generic high conversion, high hydrogen addition Basic Upgrader in order to convert "excess" gas oil to naphtha and middle distillate. The flexibility of such hydrocracking has not been examined in this study, although such will be essential in project specific studies. The beginning of run, end of run yield differences relative to hydrocracking will be at least as great as the differences noted by CANMET.

The use of only updated Cold Lake yield data in this study appears to introduce a minor bias, especially relative to average naphtha and below average gas oil yields. However, the degree of hydrocracking of the secondary hydrotreating system can be very significantly influenced by design revisions, catalyst selections (as more than one is considered very probable), operating conditions and run lengths between regeneration. This study assumes that there will be sufficient flexibility in primary and secondary upgrading design to provide essentially the same overall yields and yield structure with mixed as well as 100% Cold Lake and Athabasca bitumens.

A 90% service factor is used for all units.

Each of the units and associated offsite and utilities is briefly described below:

- Atmospheric Distillation

One unit to process 66,667 BPSD of bitumen and 54,120 BPSD of diluent is designed for all cases except the Expanded Base Case.

- Vacuum Distillation

One unit to process 57,670 BPSD of 650°F+ feed is designed for all cases except the Expanded Base Case.

- Primary Conversion and Secondary Hydrotreating

(Includes gas recovery, S.C.O. stabilization, gas clean-up and amine regeneration.) These will be 2 trains of a generic design similar to Veba Combi Cracking or CANMET with integral hydrotreating, each train processing 34,230 BPSD of 975°F+ in their primary units and an additional 32,644 of BPSD of minus 975°F material in their integrated hydrotreating unit.

- Hydrogen Production

The 2 hydrogen production steam methane reforming trains - each at 50% - are sized to produce a total of 168 million scfsd in total at 99.5% minimum purity. In this Base Case, this includes a 6% margin for varying feedstock qualities and slight changes in secondary hydrotreating hydrogen requirements.

- Sulphur Plants

Two 75% units, each sized for 352 LT/SD are provided. These units will be of conventional design with cold bed adsorption to maximize recovery.

Table 4.3-1
Cold Lake Base Case - Overall Mass Balance
(Ash Free Basis)

Fraction	Mass Flow (#/Hr)						API	Density	Calc BPCD	Crude WT %	Feed Vol %	Prod Vol %
	C	H	S	N	O	Total						
FEEDS												
Crude	730737	91702	38381	3514	5414	869748	10.9	0.9940	60000	100.00	100.00	
H2		28422				28422				3.27		
TOTAL	730737	120124	38381	3514	5414	898170				103.27		
C5+	730737	91702	38381	3514	5414	869748	10.9	0.9940	60000	100.00	100.00	
PRODUCTS												
H2S		2364	37833			40197				4.62		
NH3		696		3249		3945				0.45		
H2O		658			5263	5921				0.68		
C1-C3	31328	8328				39656				4.56		
C4	12839	2675				15514	115.0	0.5740	1853	1.78	3.09	2.88
C5-350	97475	17959	0	0	0	115434	58.4	0.7451	10623	13.27	17.70	16.49
350-400	33052	5425	1	0	0	38478	48.0	0.7883	3347	4.42	5.58	5.20
400-650	265739	42147	6	3	0	307895	34.3	0.8534	24739	35.40	41.23	38.41
650-975	274264	38666	31	6	3	312970	25.7	0.9001	23842	35.98	39.74	37.02
975+	16040	1207	509	256	148	18160				2.09		
TOTAL	730737	120125	38380	3514	5414	898170			64404	103.27	107.34	100.00
C4-975	683369	106872	38	9	3	790291	36.7	0.8413	64417	90.86	107.36	100.02
C5-975	670530	104197	38	9	3	774777	35.1	0.8493	62552	89.08	104.25	97.12

- Sour Water Stripper

  A single sour water stripping system with a capacity of approximately 222 USGPM is provided.

- Waste Water Treatment

  This unit is sized to treat the various waste water from the Base Case Upgrader.

- Steam and Power Plant

  This plant is designed to make the upgrader complex self sufficient in steam and dearated boiler feedwater. This unit also generates a portion of the upgrader complexes electric power requirements by expansion of 600 psig steam.

  A spare boiler is sized and included for start-up and upsets.

- Tankage and Pipelines

  Figure 4.3-1 illustrates the pipelines planned in the Base Case:

  - Diluent from existing pipeline.
  - Diluent return to existing pipeline.
  - Short S.C.O. line to Shell refinery (and by crossovers to both the Suncor S.C.O. and via the latest to Shell to the AOSPL Syncrude S.C.O. pipeline. But no specific blending facilities are provided.
  - S.C.O. line to Edmonton following existing lines and using an existing R.O.W. in Edmonton's congested east end.
  - Cross ties to Edmonton refineries.

- Return from 2 Edmonton area S.C.O. tanks to InterProvincial to permit batches to full line #2 rate, if desired, and to TransMountain if the upgrader lines 100,000 BPSD capacity is not sufficient. (This bypasses IPPL's smaller generic S.C.O. tankage, avoiding contamination with other S.C.O.'s. (Suncor apparently is now planning the same approach.)
- A cross tie to Imperial's (heavy) crude line running south may also be provided to allow S.C.O. to move to Montana refineries.

Table 4.3-3 outlines the anticipated upgrader related tankage.

- Raw Water Treatment

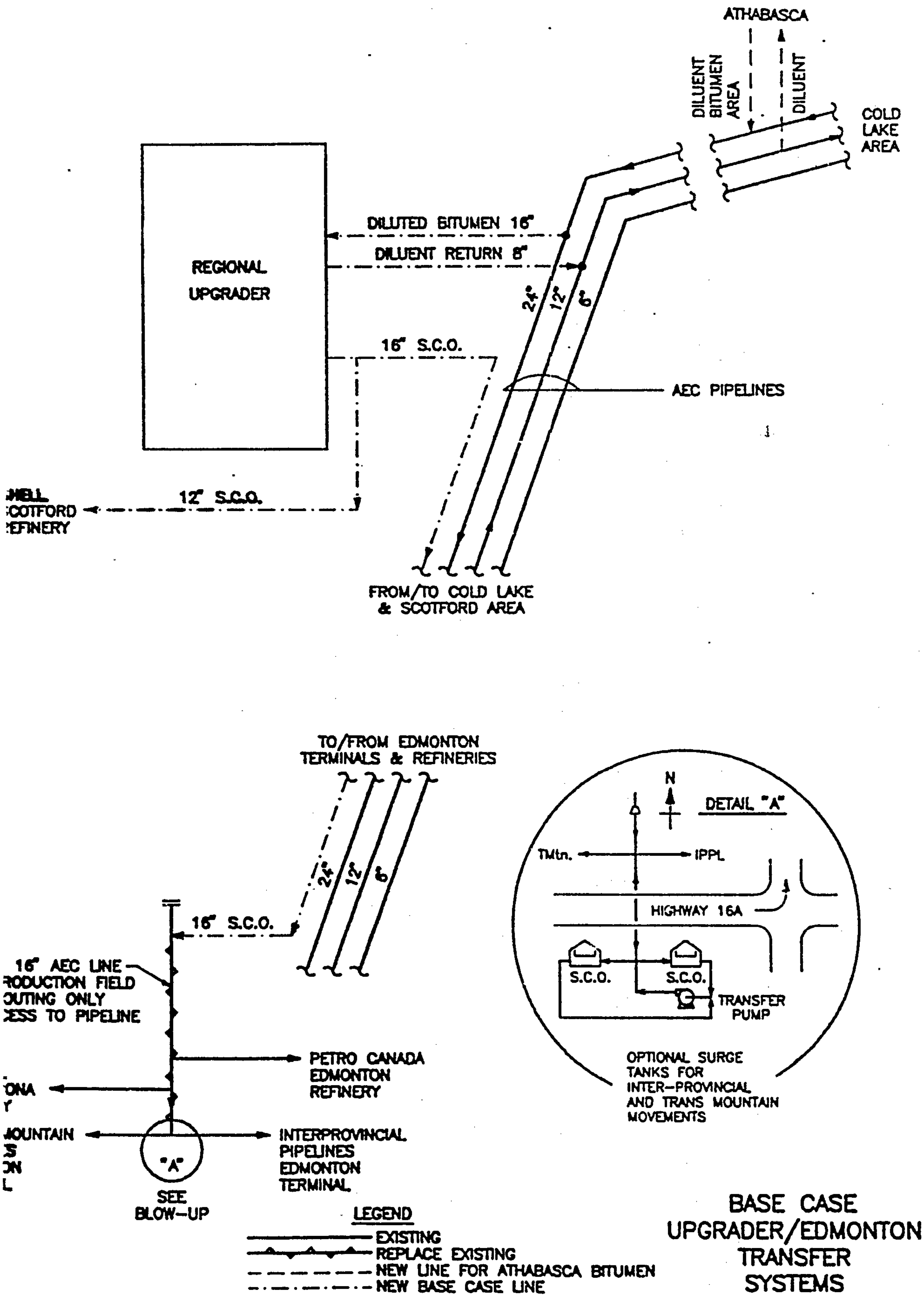
This unit is designed to treat raw water from the North Saskatchewan River for portable water upgrader utility water, cooling tower make-up and boiler feedwater make-up.

- Other Offsites and Utilities

Plant and instrument air, cooling water, fuel gas, flares and flare headers, fire protection, etc., will be based on conventional refinery systems.

- Interconnecting Pipeways

These will be provided as required.



BASE CASE
UPGRADER/EDMONTON
TRANSFER
SYSTEMS
FIG 4.3-1

Table 4.3-3
Basic Upgrader Tankage

Where	Material	Days Storage	Use/Production Rate BPDS	No. of Tanks	Individual Tank Size (gross bbls)	Type/Specials
Site	Diluted Bitumen	8	110,000 (max)	4	220,000	Cone roof (a) c/w mixers
Site	Diluent	5	43,000 (max) (b)	2	120,000	Floating roof
Site	Virgin Atmos Distillates (<700°F)	3	16,000 (max) (b)	1	55,000	Cone roof - N_2 blanket - circ heat/mix system
Site	Vacuum Gas Oil	3	17,000 (max) (b)	1	60,000	Cone roof - N_2 blanket - insulated - circ heat/mix system
Site	Vacuum Bottoms	3	36,000 (max) (c)	1	120,000	Cone roof - insulated - circ heat/mix system
Site	S.C.O.	10	70,000	4	180,000	Floating roof - circ heat/mix system
Edmonton	S.C.O.	(5)	(300,000 bbl batches)	2	170,000	Floating roof
Site	S.C.O. Quality Slops			1	20,000	Cone roof - circ heat/mix system
Site	Light Sour Slop			1	10,000	Floating roof
Site	Heavy Slop			1	30,000	Cone roof c/w heating system - insulated
Site	Wet Slop			1	1,000	Cone roof - insulated/heated/special water draw

Notes:

(a) Provide vapour recovery system to minimize odours.
(b) Sized for 100% Cold Lake bitumen.
(c) Sized for 100% Athabasca bitumen.

- Common and Services Buildings

    - Operating Centre - control room, switchgear, change, etc.
    - Central Maintenance
    - Medical Centre
    - Fire House
    - Central Laboratory
    - Offices
    - Central Warehouse

4.4 PRODUCT YIELDS AND PROPERTIES

Table 4.3-1 above provided the estimated yield structure and elemental balance for the Base Case for Cold Lake crude.

There is significant hydrocracking in the secondary hydrotreating, and yield patterns can be adjusted by changing the conditions in that step during operations and by changing catalysts to suit.

4.5 UTILITY BALANCE

The upgrader complex is designed to be self-sufficient in steam and the treatment and distribution of water obtained from the nearby North Saskatchewan River. Essentially all of the required fuels for the upgrader complex is produced in the upgrader in the form of gas from primary/secondary conversion units. Natural gas is imported for hydrogen plant feed. A portion of the upgrader complex's power requirement is generated in the steam and power plant by expansion of excess high pressure steam. The remainder of the required power is purchased. The net gas and power imports are summarized in Table 4.8-1.

4.6 CAPITAL COST ESTIMATES

Capital cost estimates are expressed on 1Q93 Canadian dollars. The costs of Table B-1 of Appendix B were developed as follows except for the hydrogen units:

a) The general bases for the 1990 "Diverse Interests" case were reviewed.

b) Appropriate factors were then used for each account in the Table to convert from 4Q89 costs to 1Q93 costs, allowing for inflation and minor construction related practice changes to match those expected in the 1994 to 1996 field construction period.

The hydrogen units were estimated by developing a new process design complete with equipment sizes and pricing all major equipment from file data, and with Foster Wheeler assistance for the steam methane reformer per se.

The Direct Field Expenses estimated for each plant sector were consolidated for estimating the general field expenses, engineering and procurement, owners budget, start-up budget, capital spares, and allowances for omissions and contingencies. The 1990 report's estimates for initial catalysts and chemicals was prorated to 1Q93 costs. The total installed plant cost in Appendix B includes all these cost elements.

The Base Case Capital sector Direct Field costs were developed as noted above. The other costs were developed as follows:

- General Field Expense

These are construction costs for supporting the direct field labour, including field supervision, temporary facilities, construction equipment, consumable supplies, tools and services. This expense was estimated at 144% of field labour.

- Engineering and Procurement Costs

This covers all the costs for engineering, procuring and constructing the project as incurred by the EPC contractors. This cost is based on an average requirement of 3,400 manhours per million dollars of direct field cost expenditure and is costed at $52.00 per manhour.

- Bussing and Travel Premiums

This allowance covers the cost of bussing for the direct subcontract labour.

- Initial Catalysts and Chemicals

This cost reflects the initial inventories of catalysts and chemicals. It was originally derived in detail based on individual unit requirements and those estimates were inflated to 1Q93 values.

- Owner's Budget

This is an allowance which covers owner's staff costs during engineering and construction, the cost of obtaining all necessary permits, studies, insurances and other miscellaneous costs not part of the construction estimates. This budget is estimated at 10% of total constructed cost.

- Start-up Budget

  This budget allows for personnel costs and fixed operating costs for the initial start-up. It includes permanent staff plus contractor, licensor and equipment vendor personnel. This budget is estimated at 5% of total constructed cost.

- Capital Spares

  This is an allowance which covers the cost of spare parts in the warehouse and is estimated at 3% of equipment cost. Installed spares are included in the direct field costs.

- Allowance for Omissions

  This is an allowance to cover design and for estimating deficiencies, and is estimated at 10% of all costs discussed above.

- Contingency

  This is an allowance for unforeseen costs which are likely to occur, and is estimated at 10% of all capital cost items discussed above. This allowance should bring the total estimated costs to the order of 30% of actual.

The following items are not included in the capital cost estimate:

- Land costs, leases and right-of-way
- Access roads to the plant's fence
- Railroads
- Escalation

- Financing charges
- Process royalties
- Production related costs

4.7 WORKING CAPITAL ESTIMATES

Working capital is shown in Table 4.7-1 and includes the value of raw materials, intermediate product and synthetic crude oil at the values of Appendix A, assuming all tanks are half full.

4.8 OPERATING COST ESTIMATES

Operating costs were re-estimated from the 1990 report on the following bases, as reported in Table 4.8-1:

- Energy Imports

  The natural gas was costed at $1.55 per million BTU's as in Appendix A (but the variation with time must be noted). The average cost of electricity at a plant service factor of 0.9 was developed by increasing the average demand by 10% to develop the peak demand for demand cost component determination, with the average usage rate used to develop the energy charge cost. Appendix A shows the average costs developed for the 0.9 service factor. Raw river water was assumed free, but electricity for pumping is included.

- Catalysts and Chemicals

  The annual cost is based on individual unit requirements inflated to 1Q93 from the 1990 report.

- Operating Labour

This is the cost for the staff directly concerned with the upgrader operations. Laboratory and technical staff are also included in this category at 25% of direct operating labour. Operating labour was costed at $76,000 per man-year. This amount includes payroll burden and fringe benefits at the rate of 30% of base salary.

- Maintenance Labour

This is equivalent annually to 2% of the total constructed cost, and based on the use of subcontract labour.

- Maintenance Material

This is equivalent annually to 2% of the total constructed cost.

- Miscellaneous Operating Supplies

This item was assumed at $0.55 million (up 10 percent from 1990).

- Administrative and Support Expenses

This is the cost for administrative and support staff costed at $76,000 per man-year. This amount includes payroll burden and fringe benefits at the rate of 30% of base salary.

- Office Costs and Miscellaneous

  This allowance covers other office expenses, travel and contractual services not related to the upgrader per se. This item was assumed at $0.55 million (up 10% from the earlier report).

- Insurances

  The annual cost is estimated at 0.25% of the total installed plant cost.

- Local Taxes

  The annual cost is estimated at 0.5% of the total installed plant cost.

- Interest on Working Capital

  An annual cost of 7% of working capital is assumed.

Table 4.7-1
Base Case Working Capital
in 1,000's of 1Q93 Canadian Dollars

Feed/Product	Average Value $/bbl	Average Inventory bbls	Cost 1,000's
Diluted Bitumen	17.58	440,000	7,735
Diluent	24.91	120,000	2,989
Intermediates and Slop	18.27	147,750	2,699
S.C.O.	26.54	530,000	14,066
Total Working Capital			27,489

Table 4.8-1
Base Case Operating Cost Estimates
in 1,000's of 1Q93 Canadian Dollars

Variable Cost	
• Natural Gas (51,680 x 10^6 BTU/CD)	29,238
• Electricity (33.6 MW average)	8,273
• Catalysts and Chemicals	8,902
• Pitch Disposal	4,022
Sub Total	50,435
Semi Variable Cost	
• Operating Labour	12,920
• Maintenance Labour	22,295
• Maintenance Materials	22,295
• Miscellaneous Operating Supplies	550
• Administration and Support	4,560
• Office Costs and Miscellaneous	550
• Insurances	2,787
• Local Taxes	5,574
• Interest on Working Capital	1,924
Sub Total	73,455
Total Operating Costs	123,890

5.0 EXPANDED BASE CASE

5.1 INTRODUCTION

This section covers the design concepts and capital and operating costs for a Regional Upgrader 50 percent larger than the basic 60,000 BPCD of bitumen feed case. This case was added to provide a comparison between the economics of added products and a larger upgrader. The 50 percent increase was based on a third primary/secondary upgrading train identical to the two of the basic plant.

The battery limit and scope definitions continue as in the Base Case, including upgrading, associated upgrader site utilities and offsites, local pipelines and an Edmonton terminal.

5.2 DESIGN BASES

The design bases remain unchanged from the Base Case, with the exception of the following:

a) Capacity now 90,000 BPCD,
b) 3 primary and secondary upgrading trains of equal size,
c) 3 hydrogen production trains (versus 2 in the Base Case), and
d) 3 sulphur production trains, each 50 percent of capacity, with 99.0 percent recovery efficiency (versus 2 at 98.7% in the Base Case).

5.3 DESCRIPTION OF DESIGN

With the above revisions to the number of trains, prorating of other unit/sector sizes to 1.5 times the capacity, is applicable in all cases, except that the pipeline size rises only to 18 inch.

5.0 EXPANDED BASE CASE

5.1 INTRODUCTION

This section covers the design concepts and capital and operating costs for a Regional Upgrader 50 percent larger than the basic 60,000 BPCD of bitumen feed case. This case was added to provide a comparison between the economics of added products and a larger upgrader. The 50 percent increase was based on a third primary/secondary upgrading train identical to the two of the basic plant.

The battery limit and scope definitions continue as in the Base Case, including upgrading, associated upgrader site utilities and offsites, local pipelines and an Edmonton terminal.

5.2 DESIGN BASES

The design bases remain unchanged from the Base Case, with the exception of the following:

a) Capacity now 90,000 BPCD,
b) 3 primary and secondary upgrading trains of equal size,
c) 3 hydrogen production trains (versus 2 in the Base Case), and
d) 3 sulphur production trains, each 50 percent of capacity, with 99.0 percent recovery efficiency (versus 2 at 98.7% in the Base Case).

5.3 DESCRIPTION OF DESIGN

With the above revisions to the number of trains, prorating of other unit/sector sizes to 1.5 times the capacity, is applicable in all cases, except that the pipeline size rises only to 18 inch.

5.4 PRODUCT YIELDS AND PROPERTIES

Table 4.3-1 in the Basic Upgrader section, and Table 2.1-1, provide data on the qualities of the product. The overall yields (on a Cold Lake only basis) are estimated at 96,626 BPCD on a C_4/975°F basis and 93,828 BPCD on a C_5/975°F basis.

5.5 UTILITY BALANCES

Capital costs for the Expanded Base Case are set out in Appendix B. Fuel gas purchases and electricity demands will each rise by 50 percent over the Base Case.

5.6 CAPITAL COST ESTIMATES

The capital costs (see Appendix B-1) for these expanded cases have been adjusted linearly from the Base Case where trains are added and at appropriate exponents for all other cost sectors. The latter generally follow those used previously to convert from 30,000 to 60,000 BPCD, but it must be recognized that the atmospheric and vacuum units will be the largest by far in Alberta and field erection of the two major columns will be needed.

The added sulphur recovery efficiency is believed achievable without significant changes and, hence, no correction was made to the unit cost. However, use of identical costs for incremental trains does add some fat as there will be savings due to much joint purchasing and design. (Identical not mirror image trains are assumed throughout this study.)

5.7 WORKING CAPITAL ESTIMATES

See Table 4.7-1 in the Basic Upgrader section, with money tied up increased by a factor of 1.5 to $41,234,000.

5.8 OPERATING COST ESTIMATES

These generally prorate up from the Base Case following the factors used there or 1.5 times in the case of utilities, as shown in Appendix B. Miscellaneous operating, administration and support, and office and miscellaneous costs were increased over the Base Case by 20% after a 10% inflation adjustment.

Table 5.8-1
Expanded Case Operating Cost Estimates
in 1,000's of 1Q93 Canadian Dollars

Variable Cost	
• Natural Gas (77,520 million BTU/CD)	43,857
• Electricity (50.4 MW average)	12,358
• Catalysts and Chemicals	13,353
• Pitch Disposal	6,033
Sub Total	**75,601**
Semi Variable Cost	
• Operating Labour	14,820
• Maintenance Labour	31,622
• Maintenance Materials	31,622
• Miscellaneous Operating Supplies	700
• Administration and Support	5,472
• Office Costs and Miscellaneous	700
• Insurances	3,953
• Local Taxes	7,905
• Interest on Working Capital	2,885
Sub Total	**99,680**
Total Operating Costs	**175,281**

6.0 FISCHER-TROPSCH CASE

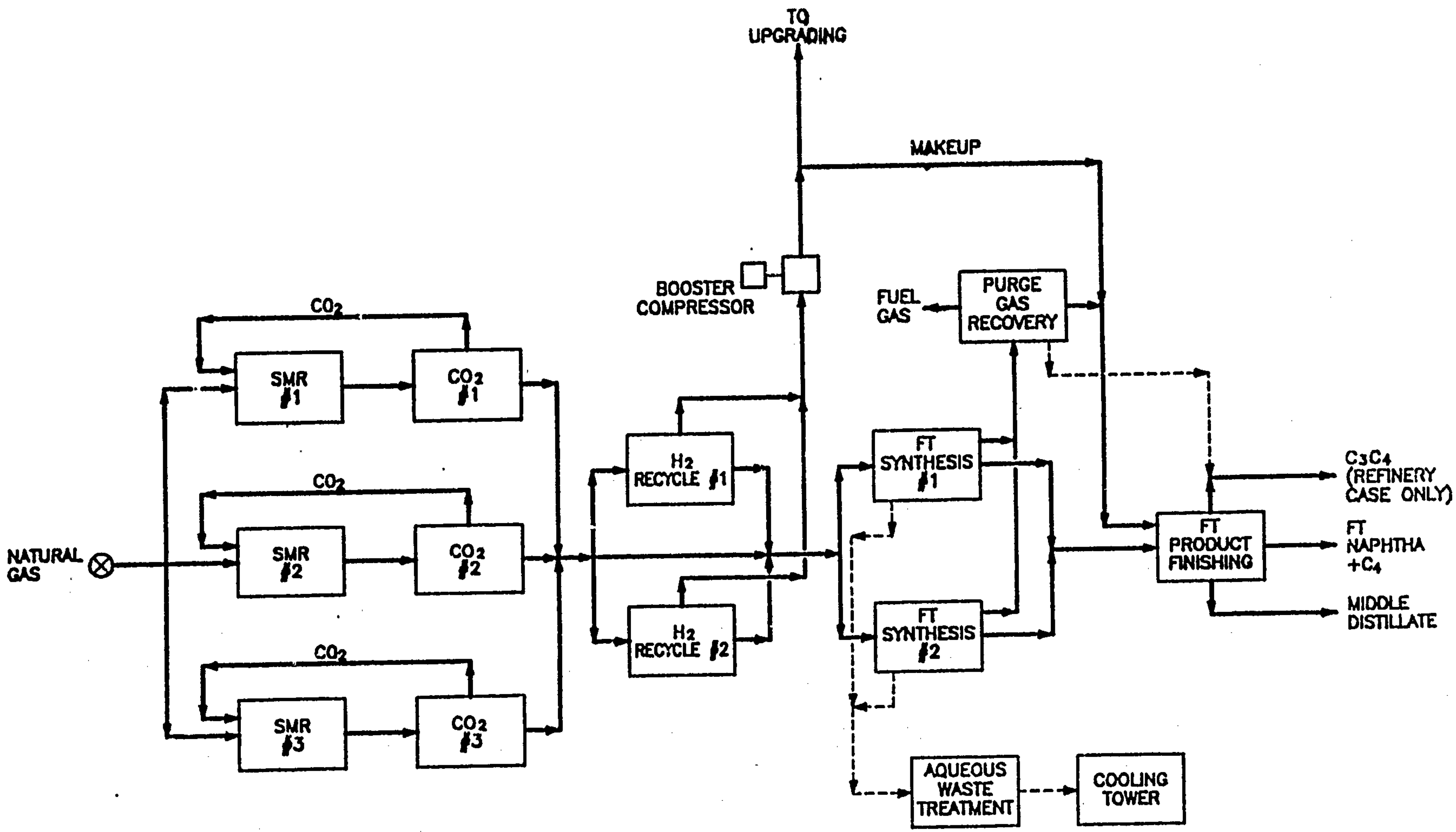
6.1 INTRODUCTION

A natural gas conversion to hydrogen and incremental liquid products system is added to the Base Case in this section. The Fischer-Tropsch process was selected for detailed study due to two current major projects introducing natural gas as a feedstock and prior selection as the most promising route for this study.[2]

Figure 6.1-1 provides an overview of the various process steps. The initial steam methane reformers convert natural gas, steam and recycle CO_2 to an $H_2/CO/CO_2/CH_4$/tr N_2 raw synthesis gas. As the CO_2 does not react in F-T synthesis, it is removed and recycled to produce more CO in the reformers. The 400 psig gas is now separated into a pure (99.5% plus) hydrogen for upgrading needs - 168 million scfd - and feed for F-T synthesis with an H_2/CO ratio of 2/1. The F-T synthesis reactions convert that synthesis gas to a wide boiling range of hydrocarbons and a variety of oxygenates. These liquids are further processed to produce naphtha and middle distillate products, the latter of exceptional properties due to the near total absence of aromatics.

Section 3.3 above discussed the valuation of the middle distillate - Jet A and diesel - boiling range material. The naphtha product will receive approximately light sweet crude oil as a petrochemical feedstock. The butane content of the product will be discounted to field-butane value due to its high concentration, but will join the naphtha as ethylene feedstock.

2 Shell in Malaysia and a South African group have major natural gas based F-T plants in construction. A small Colorado plant started up in 1991 to convert landfill methane and CO_2 to diesel and wax speciality products. But FT has been used for many years in South Africa converting coal derived synthesis gas to petroleum products.



OVERALL
FISCHER TROPSCH SYSTEM
FIG 6.1-1

The estimated liquid product yield is 16,735 BPCD; 1,295 of butanes, 5,765 of C_5-C_9 naphtha, and 9,675 of C_{10} to 343°C end point middle distillates. There is no oxygen left and the streams are sulphur and nitrogen free. The naphtha is a very paraffinic material suited best for ethylene production; the premium middle distillate qualities were noted and valued in Section 3.3 above.

Neglecting butanes, the liquid yields are about 24 percent on the bitumen feed to the upgrader and 23 percent of the C_5 plus S.C.O. from the Basic Upgrader. The F-T liquids could be blended with S.C.O. but are much likely to be sold separately.

This section adds a complete F-T complex on the same site as the Base Case Upgrader, with additional local pipeline and added tankage on the Edmonton end.

The F-T synthesis system process balances were developed by Energy International with the study team handling all other aspects. There are some proprietary portions of the process not developed here in detail.

6.2 DESIGN BASES

The upgrader's design bases continue as in the Base Case, except for pipelines and tankage. Some utility and waste treatment revisions are noted in the next section. The F-T system is based on 520 million scfsd of synthesis gas to F-T synthesis and 168 million scfsd of hydrogen to upgrading at the same pressure as in the Base Case.

Pipelines, tankage, F-T utilities and waste treatment are discussed in the next section.

6.3 STEAM METHANE REFORMING

As shown in Figure 6.3-1, natural gas feedstock is desulphurized using ZnO adsorbent in order to prevent sulphur passing to the reformer catalyst in this step. The sulphur content is reduced to less than 0.5 ppm. The desulphurized gas is then mixed with superheated steam and the carbon dioxide stream from the CO_2 removal unit, and is then further heated to ca 950°F before entering the catalyst filled tubes of the reforming furnace. For this project, a steam to carbon ratio of 2:1 has been set with reformer tube outlet conditions of 880°C and 254 psia. The syn gas produced has a hydrogen to carbon monoxide ratio of 3:1.

The hot flue gas after the reforming section is used to preheat natural gas, superheat steam, heat boiler feedwater and preheat combustion air for the reforming furnace. The hot syn gas leaving the reformer tubes is cooled by steam generation and preheating of boiler feedwater. The cooled syn gas is separated from the condensed water.

Capital costing has been developed from Foster Wheeler cost data for the reformer furnace plus study team process design of the up and downstream ends of the system.

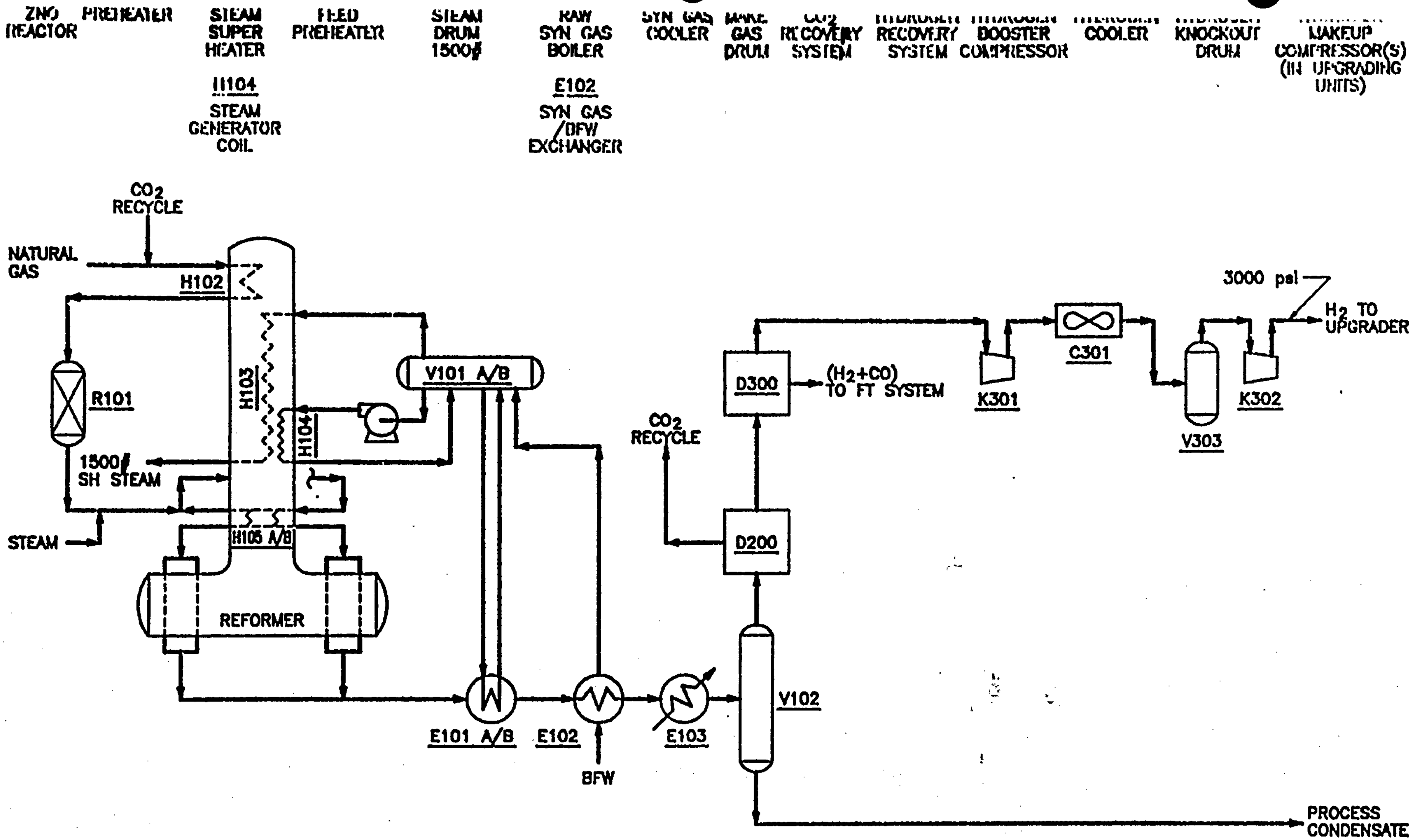
6.4 CO_2 RECOVERY

The carbon dioxide from the syn gas from reforming section is removed in a DGA unit, as shown in Figure 6.4-1. The water saturated syn gas is contacted with a solution in the contactor for CO_2 removal. The CO_2 lean gas leaves the contactor at about 250 psia and 110°F and is fed to the membrane unit, where one-third of the hydrogen in the gas is removed to make syn gas with hydrogen to carbon monoxide ratio of 2:1.[3, 4]

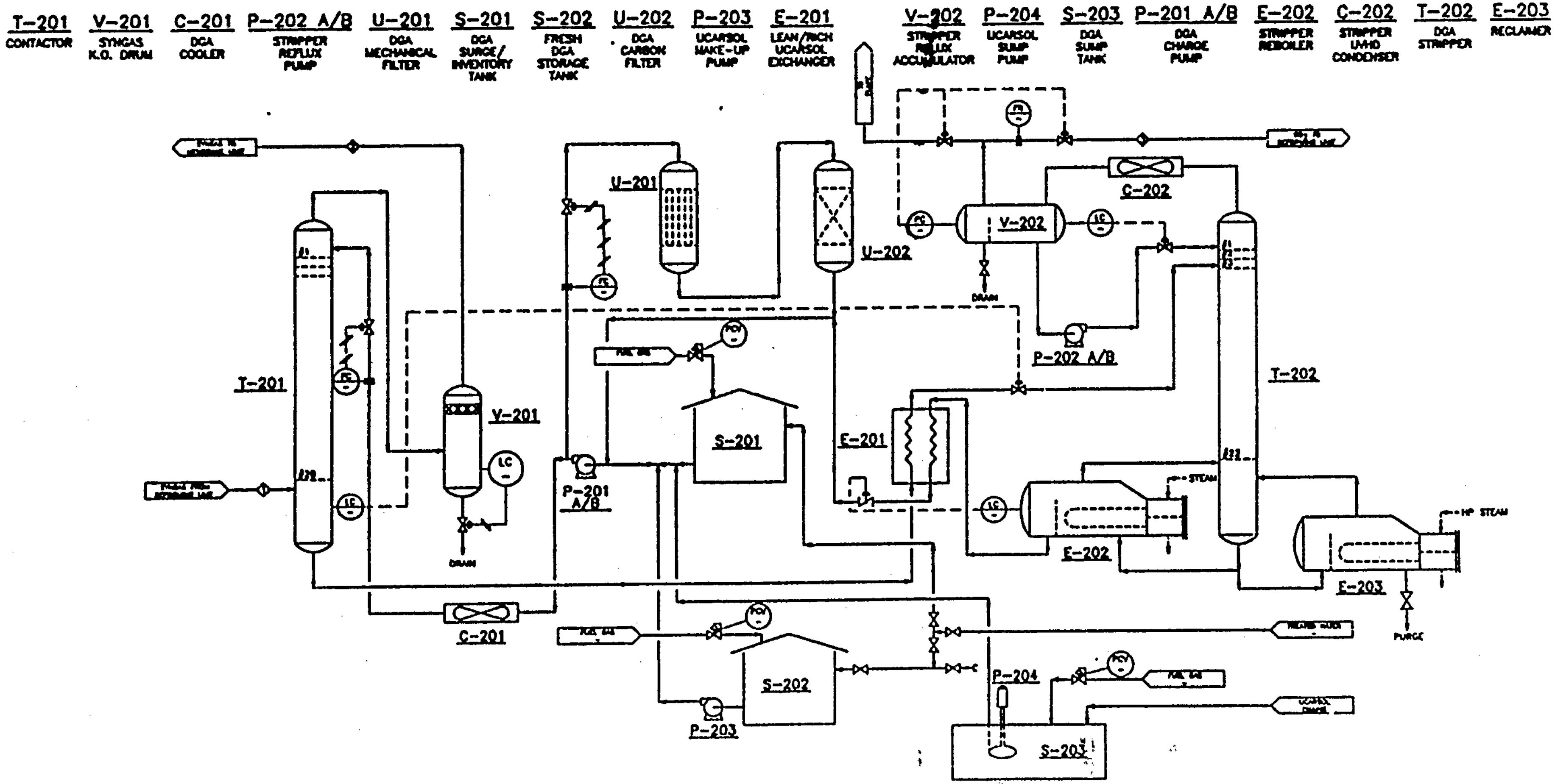
---

3 A portion of the $3H_2/CO$ syn gas may bypass the membrane unit with the latter producing 1/1 or similar blend.

4 it now appears that a P.S.A. system will be needed for the final clean-up.



FT CASE
SMR PROCESS FLOWSHEET
FIG 6.3-1



CO₂ RECOVERY SYSTEM

FIG 8.4-1

All equipment shown was sized and costed from file data in developing the capital cost estimate.

6.5 HYDROGEN SEPARATION

After prolonged consideration of alternates a scheme, proposed by Air Products, was selected with only a portion of the synthesis gas is processed through two parallel Pressure Swing Adsorption (PSA) trains to separate out the hydrogen for upgrading. The synthesis gas from the PSA trains is recompressed to F-T synthesis pressure.

This system provides the upgrader hydrogen at less pressure than from the hydrogen units of the Base Case; hence, a booster compression is shown. The latter also provides added pressure for hydrogen make-up to F-T product finishing.

The economic need for high purity hydrogen in the upgrader ruled out single membrane separation schemes. Although one membrane system vendor came close, there were doubts as to maintaining hydrogen purity over a suitably long period and a follow-up PSA unit appeared inevitable.

The capital cost for this section was developed from an Air Products proposal plus study team assessment of auxiliary facility needs.

6.6 FISCHER-TROPSCH SYNTHESIS

6.6.1 PREAMBLE

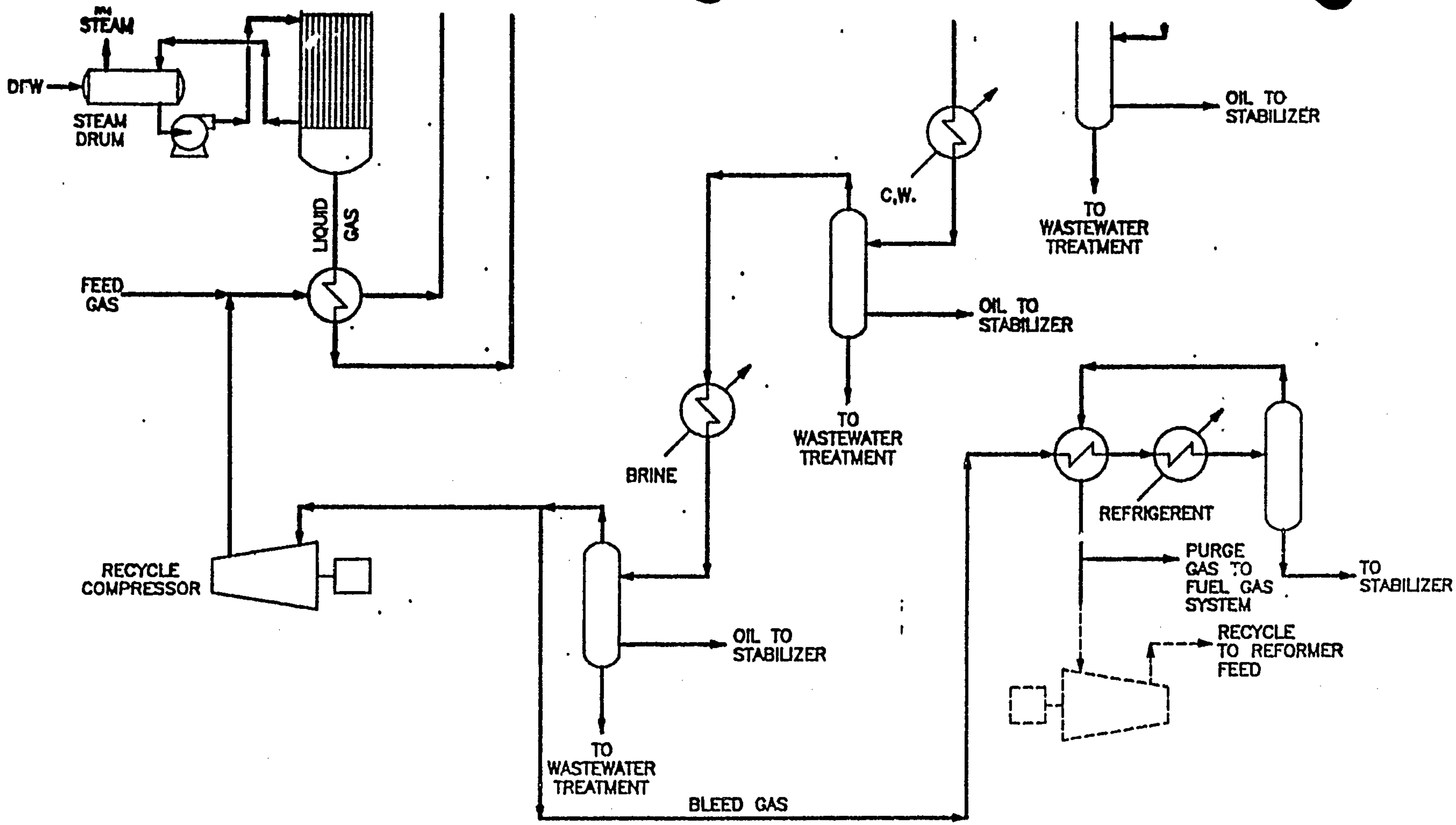
The following descriptions are taken from the Energy International report on F-T synthesis over promoted cobalt catalyst in the tubes of 16 boiling water temperature controlled reactors - 8 in each of two trains. The F-T synthesis system receives its 2 H_2/ 1 CO synthesis gas directly from the upstream hydrogen separation system - i.e. there is not compression of the bulk of the 520 million scfd of make-up synthesis gas feed the F-T synthesis process at 775 psig. But there is very significant recirculation of gas to achieve a high level of conversion.

The EI report, available as a separate document, should be consulted for details. But there are certain proprietary factors, such as catalyst promoter, not covered in the report and only available under a confidentiality agreement.

6.6.2 FISCHER-TROPSCH PROCESS DESCRIPTION

Gas to oil technology is a means of producing premium grade light hydrocarbons in the transportation fuel range from natural gas through the catalytic conversion of carbon monoxide and hydrogen (syn gas) to paraffinic light liquid products. This technology referred to as the Fischer-Tropsch process has been known for nearly 70 years and usually uses an iron or cobalt catalyst to convert the syngas to a wide boiling range mixture of liquids. The process the EI designed in this report is based on reacting a syngas produced by reforming natural gas with an adjusted hydrogen to carbon monoxide molar ratio of 2 to 1 over a cobalt catalyst.

Following a description of the synthesis unit of the F-T process shown in the one flow sheet 6.6.2-1. Syn gas at a pressure of 275 psig and a temperature of 100°F hydrogen concentration adjust system following the CO_2 removal system of the natural gas reforming plant is fed to Location 1 at a hydrogen to carbon monoxide ration of 2.0. While the flow sheet depicts a single item for each function, the plant design is actually based on two identical trains, with each train utilizing eight F-T synthesis reactors.



FISCHER TROPSCH SYNTHESIS

Figure 6.6-1

The fresh syn gas from is blended with recycle synthesis gas from the compressor to provide the total syngas feed to the F-T reactors.

The total syngas feed is preheated by heat exchange with the reactor effluent to a temperature of 422°F at Location 4 from whence it is directed to the inlet of the multi-tubular fixed bed catalytic reactor. Flow distribution to the individual tubes within the reactor is controlled passively by the catalyst loading procedure that assures an equal charge of carefully screened catalyst within each of the multiplicity of tubes within plus or minus 5 percent by weight. The F-T reaction takes place at a steady rates as the syngas flows downward over the catalyst in each tube, producing a range of hydrocarbons with a carbon number distribution governed by the Schulz-Flory probability relationship.

The substantial quantity of heat liberated by the F-T heat of reaction is largely removed by the flow of pressurized liquid water flowing co-currently downward to the shell side of the reactor. The water remains single phase throughout the shell with the water flashing to steam in the steam drum as shown. The temperature of the inlet cooling water is regulated by control of the pressure on the steam drum. A small portion of the F-T heat of reaction is accounted for by the temperature rise of the syngas as it flows downward through the tubes, with the outlet syngas at Location 5 of 442°F being 20° elevated over the inlet temperature.

And, finally, a small amount of additional heat is generated in the reactor and removed by the cooling water as a result of the inevitable condensation of a portion of the heavier hydrocarbons formed, as governed by vapour-liquid equilibrium considerations.

The vapour-liquid stream from the reactors interchanges heat with the reactor feed stream, dropping its temperature to 210°F, and is then directed to oil/water/vapour separator MS 301. The vapour stream from this separator is cooled in two steps to 50°F. The remaining synthesis gas at is split into two parts. The larger fractions are recycled via the recycle gas compressor to blend with the fresh synthesis gas. The smaller fraction is further cooled with refrigerant to a temperature of -35°F to wring out most of the remaining liquid hydrocarbons before retrieving the purge gas stream for recycle to either the reformer feed or use in the plant fuel circuit.

The multiple levels of cooling and liquid separation are necessary to prevent freezing of the several liquid streams that are recovered. These individual hydrocarbon liquid streams are combined to form the feed to the stabilizer column. A single stabilized liquid product is produced that is suitable for intermediate storage prior to hydrocracking/hydrotreating. The co-product water contains most of the oxygenates produced as a byproduct of the F-T reaction. This water requires treatment prior to discharge. The concentrations and nature of the contained materials is such that simple biological oxidation provides an environmentally acceptable solution.

6.6.3 CATALYST SPECIFICATIONS

EI has had commercial experience with a number of the preferred cobalt based F-T catalysts and has just completed a major survey of the technical and patent literature on all aspects of the subject. The catalyst selected for this design study is a generic cobalt based catalyst containing 20% cobalt supported on 1/16 inch extrusions of gamma alumina with small qualities (~1-2% total) of additional constituents that both aid in the reduction of the cobalt and promote hydrocarbon chain growth.

Following are typical physical properties which were used as the basis for the process design provided in this report:

- Surface Area — 130 square meters per gram
- Pore Volume — 0.5 cubic centimeters per gram
- Bulk Density — 45 pounds per cubic foot

- Side Crush Strength — 14 pounds per square inch
- Average Size
  - Diameter — 0.064 inches
  - Length — 0.141 inches

The price of the catalyst with current metal prices is approximately $15.00 per pound, FOB manufacturer.

Vendors who are capable of providing catalysts of this type include:

- EI for kilogram quantities
- Mallinckrodt Specialty Chemicals Co.
- Davidson Division of W.R. Grace Company

EI has clients who prefer to have EI specify, test and provide them with tonnage commercial quantities. EI undertakes the full responsibility for the catalyst supply and selects the company that will actually manufacture the catalyst to EI's specifications.

The spent catalyst is usually returned to the supplier for reprocessing or metals recovery and disposal.

6.6.4 CATALYST LIFE AND REGENERATION

The useful life of the typical supported cobalt F-T catalyst is likely to be in excess of 5 years in the absence of sulphur poisoning; the sulphur gathering activity of the steam methane reforming catalyst from whence the syngas is produced is such that inadvertent exposure to sulphur is quite unlikely. Likewise, physical decrepitation of the catalyst is unlikely given mild flow and temperature conditions of the application, combined with the inherent strength and stability of the gamma alumina support. The current pilot operating experience suggests that 5 years is a conservative life time for the catalyst from a cost estimating standpoint.

There is a slow decline in catalyst activity that accompanies gradual accumulation of heavy F-T wax in the catalyst pores, that results in a need to regenerate the catalyst on a scheduled yearly basis.

The regeneration operations involved three steps: hydrogen stripping, a treatment with hydrogen to hydrocrack wax deposits on the catalyst; a two step oxidation in which the catalytic metals are first oxidized at 100-120°C, and then the carbon is burned off at temperatures up to 310-340°C; and reduction of the cobalt and metal oxides in hydrogen over a programmed temperature range up to 300-320°C.

The hydrogen stripping step of the regeneration procedure, which remove large amounts of accumulated wax from the interstices and pores of the catalyst, is absolutely essential to the smooth conduct of the subsequent oxidation steps.

The hydrogen stripping operation is conducted at a space velocity of approximately 1000 scfh/volume of catalyst, and is increased to 2000 as the hydrogen feed displaces the other gases initially present.

Hydrogen stripping will be complete in 24 hours; the temperature will then be reduced to 100°C. At this time, the system is ready for the initiation of the two stage oxidation.

Oxygen is provided by mixing air with nitrogen to produce an oxygen concentration of approximately 0.4%. This gas is introduced at a space velocity of approximately 1000, and the temperature is maintained at 100°C for 4 hours, completely oxidizing the cobalt and other metals present. Subsequently, the temperature is increased to approximately 300°C and is held there for 48 hours, allowing for the complete burn off of the residual carbon from the hydrogen stripping operation. Upon completion of this step, hydrogen reduction is accomplished by following the procedure for the initial activation of the catalyst. Total time for regeneration is approximately 76 hours.

Given the multiplicity of reactors, there may well be an advantage is providing for the sequential regeneration of 1 or 2 reactors at a time while the plant remains on stream.

6.6.5 CATALYST INSITU ACTIVATION

The F-T catalyst requires activation after loading in the multi-tubular fixed bed-reactor. This would typically be accomplished as a phase of the initial start-up of the plant. By purging the reactor prior to shutdown and maintaining an inert atmosphere (nitrogen or natural gas preferably) during any shutdown period, the catalyst will be maintained in a process ready condition.

The activation procedure is basically a combination of drying and reduction of the cobalt and other metal oxides that are present. The gamma alumina support itself undergoes no chemical change under the conditions used. The conceptual procedure is to purge the reactor with an inert gas followed by purging with hydrogen to be provided from elsewhere within the complex. The purity of the hydrogen is not critical other than requiring a sulphur level no greater than the sulphur level acceptable for methane steam reformer operation, and might be nominally set at 90 percent or higher hydrogen purity.

Hydrogen flow would be initiated at ambient temperature, with temperature programmed to increase to 100°C to cover no less than a 10-hour period of time. Temperature would be held at 100°C for no less than 8 hours. Hydrogen throughout should be maintained at a space velocity of approximately 1000 scfh/ft^3 catalyst.

Next increase the hydrogen temperature at a rate of 10°C per hour or less until the catalyst temperature reach 300°C. Maintain the hydrogen mass flow rate at the same level as before, and at no less than 300°C and not more than 325°C for 16 hours. At the end of 16 hours, reduce temperature to 200°C while maintaining hydrogen flow, and hold at 200°C until ready to introduce synthesis gas preparatory to the start-up of F-T operations.

6.6.6 F-T PURGE GAS PROCESSING

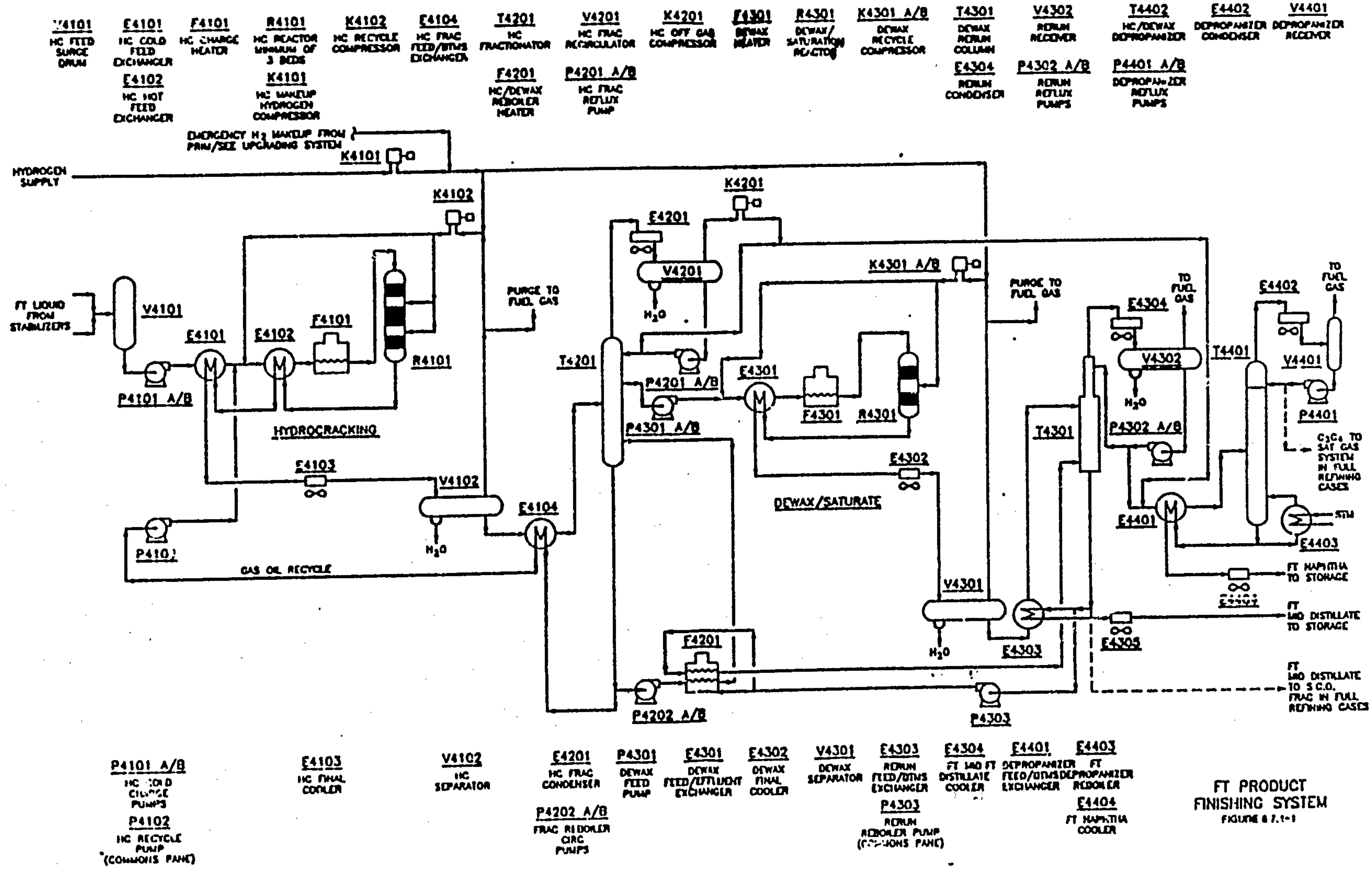
The purge gas contains a substantial volume of hydrogen, much of which will be recovered in a P.S.A. system to provide hydrogen for F-T product finished. (The small residual hydrogen demand will be taken from the "margin" in hydrogen supply to upgrading.) As the bulk of the remainder is methane, it will go to fuel gas.

The nitrogen from the original natural gas feed ends up in this purge gas which precludes total recycle to steam methane reforming without elaborate processing for it removal. In practice, recycle of 50 to 60 percent is foreseen. In the future the C_3 and C_4 content of this stream should be exploited - they are roughly 25% olefins and should be considered for at least alkylation feed in a refinery.

6.7 F-T PRODUCT FINISHING

6.7.1 PREAMBLE

Figure 6.7.1-1 provides a quick overview of the major flows through the single train F-T product finishing system. While the unit titles appear relatively commonplace in practice, the actual services are quite different, especially in the "hydrocracker" area.



FT PRODUCT FINISHING SYSTEM

FIGURE 6.7.1-1

Licensors were reluctant to provide data for the F-T "hydrocracker", but Mobil has provided comments on the dewaxing of middle distillates and the subsequent saturation of olefins produced in dewaxing. Generally the F-T hydrocracker balances are drawing on 1988 UOP hydrocracking of a different Fischer-Tropsch wax. The overall process yields are not considered of the same degree of accuracy as those of other sections of this study.

But the data are considered sufficiently accurate for preliminary cost development to the ±25 percent accuracy level (after 10 percent additions for allowances and another 10 percent for contingencies.) Both hydrocracking and dewaxing/saturation units will operate at pressures below 1000 psig, in a range where conventional hydrotreating costs can be extended to provide "reasonable" capital and operating costs.

6.7.2 HYDROCRACKING

The raw F-T liquids are defined in the Energy International report as follows:

- Rate: 184,696 pounds per day, 16,900 BPCD (@ 0.75 S.G.)
- Composition: C_4's - 4 weight percent, C_5-C_8 - 24, C_9-C_{20} - 49, C_{21}+ - 21 and oxygenates 2
  Saturates - 82 weight percent (linear paraffins)
  Olefins - 16
  Oxygenates - 2

There are essentially no cyclic materials present and only a small portion of isoparaffins. The cloud and pour point of the C_9-C_{20} middle distillate fraction are high - probably in the order of +20°C with the heavier material near +80°C.

Table 6.7-1
Fischer-Tropsch Product Finishing Balance
Preliminary

Component	Weight Percent of Raw Liquids				BPCD	
	Raw Liquid from F-T Synthesis	Change in HC	Change in Dewax/Sat	Net Product (S.G.)	Max N	Max MD
C_4's	4	+1	+0.5	55 (.52)	1,295•	1,295
C_5-C_9	26	+6	+5	37 (.68)	6,623•	5,767
C_{10}-C_{20} (343°C)	47	+14	-6	55 (.75)	8,817▲	9,673
C_{21}+	21	-21	0	0	0	0
Oxygenates	2	-2	0	0	0	0
Total	100	-2*	-0.5*	97.5	16,735	16,735
S.G.	0.75			0.71	—	—
BPSD	16,900				—	—

Notes:

- * Hydrogen addition allowed.
- • In C_4/naphtha blend.
- ▲ In F-T middle distillate product.

The raw F-T liquids are very readily cracked and they must be treated gently. Shell recommends a trickle type reactor with only liquids passing slowly downwards. The reactor must convert oxygenates to water and hydrocarbons, saturate olefins, and crack the over 343°C portion, preferably to 170 to 343°C boiling range material. At the same time, significant isomerization of the linear paraffins to iso-paraffins should be achieved.

The catalyst Shell will use in its Malaysia F-T to middle distillate complex is proprietary with no performance data available. The data used here followed published UOP hydrocracking work on wax from another F-T process and, hence, are not fully comparable.

However, it is clear that a single reactor operating under 1000 psig - probably near 600 psig - will provide the desired heavy ends cracking, deoxidation and bulk olefin saturation. Quench will be needed at several points to control temperatures as exothermic reactions predominate. In practice, to provide such control, liquid recycle may be added to hydrogen quenching.

A relatively conventional fractionation of reactor products is proposed with column bottoms recycling to the reactor. The conversion of heavy ends per pass will be limited in the interest of minimizing reaction temperatures.[5]

6.7.3 MIDDLE DISTILLATE IMPROVEMENT

While significant isomerization is planned in the hydrocracker, the middle distillate fractions of its product will still have waxy components as evidenced by high cloud points, even at acceptable pour points.

---

5 In Full Refining cases, it would appear logical to consider putting the heavy fractions of the raw FT liquid direct to catalytic cracking due to their low volume and processing only the <343°C material in the FT system.

Mobil's middle distillate dewaxing process is planned to extract and crack the waxy linear paraffins. As the resulting product will be olefinic, a saturation step is needed to meet Jet A specifications.

A common reactor is planned with dewaxing at the top and saturation at the bottom. The latter operation requires a lower temperature, hence, an interstage quench is needed. These systems will operate at about 600 psig with significant hydrogen recycle.

6.7.4 PRODUCT FRACTIONATION

A rerun tower will fractionate out the middle distillate product and a C_3C_4 naphtha overhead which will be depropanized in a second tower. (In Full Refining Cases, the latter column will operate as a debutanizer.)

6.7.5 FURTHER WORK

There is more uncertainty in these steps of the F-T system than in any other portion of all cases considered in this study. Aside from pilot tests on F-T synthesis per se, much work remains on optimizing the F-T product finishing - catalyst selections and process configuration/optimizing being paramount concerns.

6.8 F-T AREA WASTE TREATMENT

Water produced in F-T synthesis will be treated for use in cooling tower make-up throughout the complex. Approximately 3,480 pph of oxygenates enter the system in 572 USGPM of waste water from F-T synthesis (plus 20 to 30 USGPM from the hydrocracker section). The treatment consists of a plate type oily water separator, dissolved air floatation, activated sludge, aerated holding ponds, followed by chlorination or other sterilization. A small amount of clean sludge will be withdrawn from the system and land farmed, sent to landfill or to farms as fertilizer.

A small parallel API separator will handle hydrocarbon drains and washdown water from the F-T product finishing area.

Boiler blowdown from the SMR's and F-T reactors will be routed to the upgrader's central boiler blowdown treating system.

6.9 F-T AREA UTILITIES

6.9.1 REFRIGERATION

Two conventional propane refrigeration systems will be provided.

6.9.2 CHILLED WATER

In order to supply the chilled water required for F-T product condensing, two large chiller

6.9.3 STEAM SYSTEMS

Boiler feedwater make-up will be drawn from the upgrader for both the steam methane reformers and the F-T reactor boiler loops.

The steam methane reformers will be designed for internal steam production with a 1500 psig superheated steam byproduct. But the F-T synthesis will produce 1.1 million pph of 150 psig steam. The latter will be used in CO_2 recovery, F-T raw liquid naphtha depropanizing. The surplus of 0.6 million pph over those needs will be superheated in a fired heater and then used, along with the 1500 psig steam surplus to SMR use, to drive the

F-T recycle gas compressors and generate approximately 80 MW of electricity for onsite use and off-site sale. Turbine exhaust steam will be condensed against cooling water.

6.9.4 CONDENSATE AND BOILER FEEDWATER

All possible steam condensate will be recovered with for reuse inside the F-T complex. The unit will obtain its boiler feedwater from an added treatment train in the upgrader's utility plant.

6.9.5 COOLING WATER

A 100,000 USGPM cooling water system will serve the two F-T synthesis systems. Make-up water will be from the F-T waste treatment system and from river water.

6.9.6 INSTRUMENT AIR

The F-T complex will draw on an added 500 scfm air compressor and air drier in the central utility plant.

6.9.7 FUEL GAS

The F-T complex will have its own fuel gas system with natural gas supplementing off gases from the various processes. This system will have continuous BTU content analysis to provide tight control with varying rates of fuel gas make-up streams. Note that there will always be a net make-up to the F-T complex due to very large SMR fuel needs.

6.9.8 FLARES

Each F-T synthesis train will have its own flare of the cylindrical grand flare type for major emergencies. An elevated flare will serve the other portions of the plant as well as the small needs of F-T synthesis.

6.9.9 PIPELINES

Figure 6.9.10-1 outlines the connecting pipelines.

6.9.10 TANKAGE

Table 6.9.10-1 sets out the proposed F-T Case tankage.

6.10 PRODUCT YIELDS AND PROPERTIES

Table 6.7-1 set out the estimated product rates and discussed product qualities.

6.11 UTILITY BALANCE

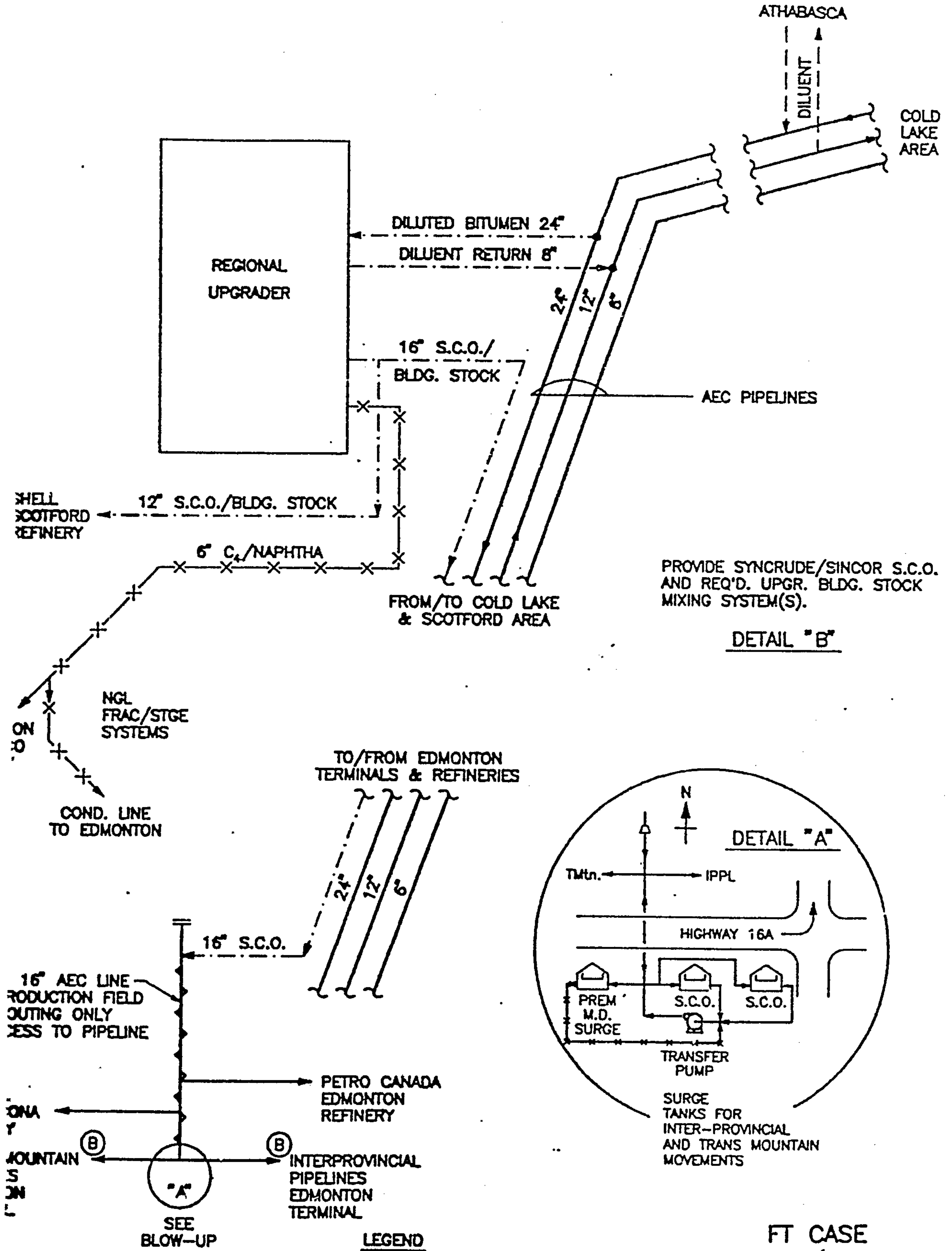
The natural gas and electricity requirements are set out in the operating cost table of Section 6.14. It is to be noted an electrical surplus is indicated with export to the provincial grid of an average of 18.8 MW.

Table 6.9.10-1
F-T Case Tankage

Where	Material	Days Storage	Use/Production Rate BPSD	No. of Tanks	Individual Tank Size (gross bbls)	Type/Specials
Site	Diluent Bitumen	8	110,000 (max)	4	220,000	Cone roof (a) c/w mixers
Site	Diluent Bitumen	5	43,000 (max) (b)	2	120,000	See Basic Upgrader Tankage
Site	Virgin Atmos Distillates (<700°F)	3	16,000 (max) (b)	1	55,000	See Basic Upgrader Tankage
Site	Vacuum Gas Oil	3	17,000 (max) (b)	1	60,000	See Basic Upgrader Tankage
Site	Vacuum Bottoms	3	36,000 (max) (c)	1	120,000	See Basic Upgrader Tankage
Site	S.C.O.	10	70,000	4	180,000	See Basic Upgrader Tankage
Site	F-T C_3C_4 Naphtha (b)	(6)	(300,000 bbl batches)	1	5,000	Bullet (b)
Site	Prem Mid Dist Blending Stock	10	11,000	2	60,000	See Basic Upgrader Tankage
Site	S.C.O. Quality Slops			1	20,000	See Basic Upgrader Tankage
Site	Light Sour Slop			1	10,000	See Basic Upgrader Tankage
Site	Heavy Slop			1	20,000	See Basic Upgrader Tankage
Site	Wet Slop			1	1,000	See Basic Upgrader Tankage
Edmonton	S.C.O.	(5)	(300,000 bbl batches)	2	170,000	See Basic Upgrader Tankage
Edmonton	Prem Mid Dist Blending Stock	(12)	(150,000 bbl batches)	1	170,000	Cone roof

Notes:

(a) See Basic Case Upgrader Tankage.
(b) Provide 5 days naphtha storage if debutanizer built - 25,000 bbl floating roof tank.



LEGEND

- EXISTING
- REPLACE EXISTING
- NEW LINE FOR ATHABASCA BITUMEN
- NEW BASE CASE LINE
- NEW FT CASE LINE

FT CASE UPGRADER/EDMONTON TRANSFER SYSTEM

FIG 6.9.10-1

6.12 CAPITAL COST ESTIMATE

Table 6.12-1 sets out the estimated capital costs by process units for this case, with Table B-1 of Appendix B setting out the overall costs. All equipment was sized and estimates on an equipment factored basis, except for the product finishing area where capacity factoring was done using Turbo file data for similar pressure hydroprocessing units hydrotreating systems, corrected as appropriate for recycle gas rates, etc.

6.13 WORKING CAPITAL COST ESTIMATE

Table 6.13-1 has been developed on the same bases as used in the Base Case for material inventories.

6.14 OPERATING COST ESTIMATES

Table B-2 of Appendix B sets out operating costs developed as in the Base Case.

Table 6.12-1
Fischer-Tropsch Process Unit Direct Field Cost
in 1,000 of 1Q93 Canadian Dollars

Unit	Direct Field Costs
SM Reforming	209,711
CO_2 Recovery	33,053
H_2 Recovery	36,250
H_2 Recovery Purge Gas	4,775
F-T Synthesis	143,805
F-T Hydrocracker	36,000
F-T Dewaxing	19,000
F-T Depropanizer	2,500
Total	485,094

Table 6.13-1
F-T Case Working Capital
In 1,000's of 1Q93 Canadian Dollars

Material	Average Value	Inventory	Value
Diluted Bitumen	17.58	440,000	7,792
Diluent	24.91	120,000	2,989
Intermediates/Slop	18.27	143,000	2,617
S.C.O.	26.54	530,000	14,066
FT Dist Product	35.70	145,000	5,177
Total		1,378,000	32,584

Table 6.14-1
F-T Case Operating Cost Estimates
in 1,000's of 1Q93 Canadian Dollars

Variable Cost		
• Natural Gas (10^9 BTU/CD)	(241.1)	136,374
• Electricity (MW)	(-18.8)	-4,673
• Catalysts and Chemicals		16,630
• Pitch Disposal		4,022
Sub Total		152,353
Semi Variable Cost		
• Operating Labour		15,580
• Maintenance Labour		36,433
• Maintenance Materials		36,433
• Miscellaneous Operating Supplies		550
• Administration and Support		4,560
• Office Costs and Miscellaneous		550
• Insurances		4,554
• Local Taxes		9,108
• Interest on Working Capital		2,281
Sub Total		110,049
Total Operating Costs		262,402

7.0 PARTIAL REFINING CASES

7.1 INTRODUCTION

This section covers the development of concepts, and capital and operating costs for the following two sub cases:

a) Fractionation of S.C.O. added to Basic Upgrader to produce more or less Jet A and diesel products and differentiated S.C.O.'s.

b) Addition of 12,000 BPCD of diluent related hydrotreater capacity.

In practice, the hydrotreater could well operate without the S.C.O. with its product being blended directly to S.C.O. The hydrotreater will be designed for a heavy sour condensate feed should such prove more viable than use of diluent return. (Refiners will definitely prefer the heavier naphtha from such a condensate, although they often have trouble processing it themselves.)

The battery limit definitions continue as in all other cases, including the Basic Upgrader, partial refining units, associated upgrader site utilities and offsites, local pipelines and an Edmonton terminal.

The Partial Refining Cases are primarily related to increasing upgrader return via differentiating the product S.C.O. into a variety of special S.C.O. blends and into Jet A (kerosene type aircraft turbine fuel) and diesel for direct sale. Adding naphtha to the S.C.O. increases its marketability, especially as a refiner's basic crude oil, as opposed to an incremental crude to be bought to fill short term needs.

In the Partial Refining cases, the S.C.O. fractionation capability has been set at 100 percent of S.C.O. production to provide maximum flexibility. However, studies indicate that it is probably safe to assume regular sale of only about 20 percent of the middle distillate as distinct products without impacting on S.C.O. netbacks.

7.2 DESIGN BASES

The Basic Upgrader design bases were described in Section 4 above and do not change for these sub cases, except for tankage (and related pipelines). Cold Lake bitumen has been used as the basis for yields considered in this section but cost and other factors do not change significantly with use of mixed or 100 percent Athabasca bitumen upgrader feed. The distribution of overall S.C.O. fractions with and without withdrawal of specification Jet A and diesel products at rates considered, will vary slightly with the feed but no impact on average unit of value of S.C.O. is anticipated. Variations in S.C.O. fractional yields due to varying feedstocks, will be ~~handled by~~ slight variations in middle distillate and differentiated S.C.O. sales.

The partial refining facilities will be single train with an availability above that of the Basic Upgrader's assumed 90 percent. The partial refining facilities will be air cooled and only incremental fuel, electricity and a small amount of instrument air will be needed. The estimates for utility consumption are based on fractionator operations 25 percent of the time.

Tankage bases are set out in Table 7.3-1 below.

7.3 DESCRIPTION OF OVERALL DESIGN

The new units are as follows:

- S.C.O. Fractionation

The process concept is outlined in Figure 7.3-1. Due to the high volume of butane in the whole S.C.O., a debutanizer has been added to allow reduction of the butane content if/when desired.

- Naphtha Hydrotreater

  This is a simple hydrotreater to process up to 12,000 BPCD of the diluent normally returned to the field. The stabilizer shown in Figure 7.3-2 could function as a debutanizer, if desired, with a C_4 rich sidecut being taken to the S.C.O. fractionator's debutanizer.

  If it is desired to process heavy sour condensates, these must first be distilled to transfer heavier fractions to the secondary hydrotreating units in the upgrading complex.

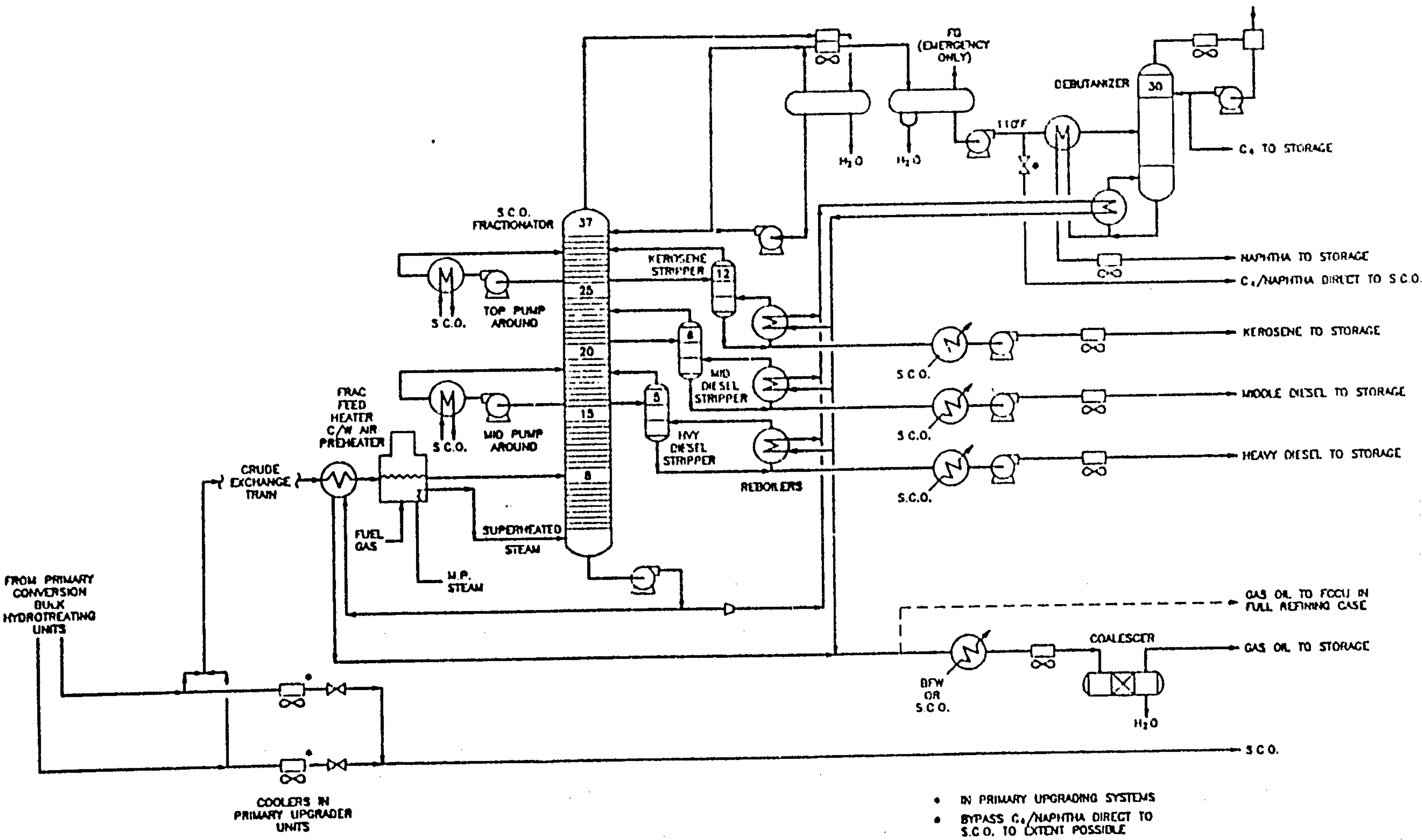
Of the process and utility facilities of the Basic Upgrader, only the following change:

- Hydrogen Production

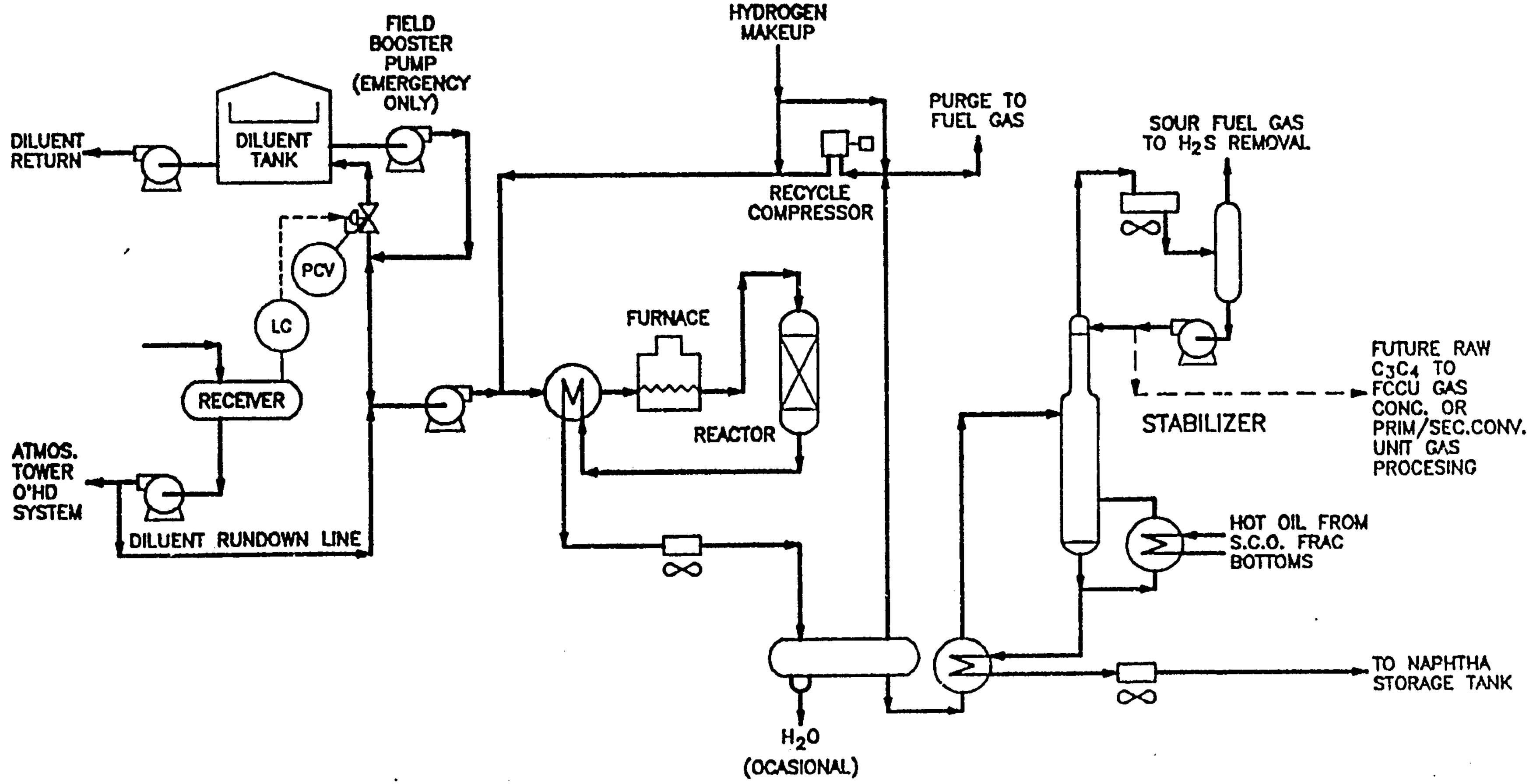
  - In the condensate sub case, up to 1 million scfd of the "million surplus" hydrogen available from the 2 trains will be needed, hence, no new capacity is planned.

- Sulphur Production

  - Up to about 8 tpd of added sulphur will be produced in the condensate addition sub case, but as this is only 1.5 percent of total production, no change in sulphur plant capacity of capital or operating costs is proposed.



PARTIAL REFINING CASE
S.C.O. FRACTIONATION
FIG 7.3-1



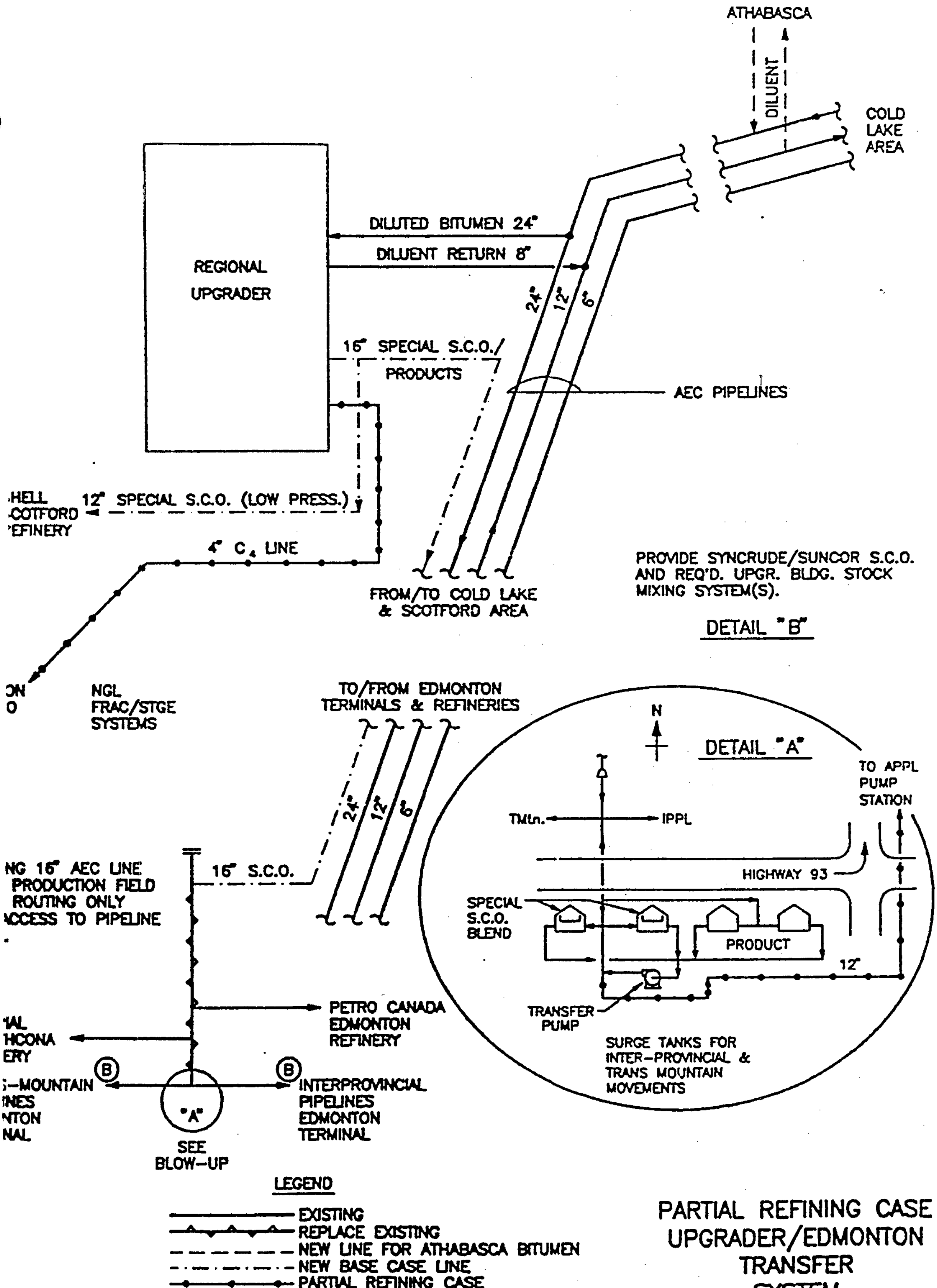
OPTIONAL
NAPHTHA HYDROTREATER IN
PARTIAL REFINING CASE
FIG 7.3–2

Table 7.3-1
Partial Refining Case Tankage (b)

Where	Material	Days Storage	Use/Production Rate BPSD	No. of Tanks	Individual Tank Size (gross bbls)	Type/Specials
Site	Diluted Bitumen	8	110,000 (max)	4	220,000	Cone roof (a) c/w mixers
Site	Diluent Return	5	43,000 (max) (b)	2	120,000	Floating roof
Site	Virgin Atmos Distillates (<700°F)	3	16,000 (max) (b)	1	55,000	Cone roof - N_2 blanket - circ heat/mix system
Site	Vacuum Gas Oil	3	17,000 (max) (b)	1	60,000	Cone roof - N_2 blanket - insulated - circ heat
Site	Vacuum Bottoms	3	36,000 (max) (c)	1	120,000	Cone roof - insulated - circ heat
Site	Light Slop		Variable	1	20,000	Floating roof
Site	Heavy Slop		Variable	1	40,000	Cone roof - insulated - circ heat system
Site	Wet Slop		Highly variable	1	1,000	Cone roof - insulated/head special water draw
Site	Butane	3	2,000 max	1	6,000	Sphere
Site	Naphtha	10	9,000 (b)	2 (b)	50,000 (b)	Floating roof (b)
Site	Kerosene	10 @ max	15,000 max	2	75,000	Cone roof
Site	Medium Diesel	10	6,000	1	75,000	Cone roof
Site	Heavy Diesel/Diesel	7 @ max (15)	32,000 max (15,000 net)	3	75,000	Cone roof - circ heat system
Site	Gas Oil	10	25,000	2	140,000	Cone roof - insulated - circ heat system
Site	S.C.O./S.C.O. Slops	2	63,200 max	2	75,000	Floating roof - mixers
Edmonton	Jet A		(150,000 bbl batches)	1	170,000	Floating roof - mixers
Edmonton	Diesel		(150,000 bbl batches)	1	170,000	Cone roof - mixers
Edmonton	S.C.O./Swing		(Up to 300,000 bbl batches)	2	170,000	Floating roof

Notes:

(a) Vent gas to be treated for odour control.
(b) In the sub case with naphtha hydrotreater added, production rate can go as high as 20,000 BPCD and number of tanks goes to 3 each at 70,000 bbl capacity.



PARTIAL REFINING CASE
UPGRADER/EDMONTON
TRANSFER
SYSTEM
FIG 7.3-3

- Sour Water Stripper

  - There may be an intermittent flow of 10 to 20 USGPM of sour water from the hydrotreater condenser water wash, less than 10 percent of the design feed rate. It is assumed that no change is needed to handle this stream.

- Waste Water Treating

  - The fractionator overhead system will condense 20 USGPM of water when operating. This water will be sweet and very pure - with clay or other adsorbent it can easily be used for boiler feedwater. Thus, no change in waste water treatment facilities is likely.

- Steam Plant

  - The fractionator will need up to 10,000 pounds per hour of medium pressure steam for stripping. No added boiler capacity is foreseen.

- Tankage and Pipelines

  - See Table 7.3-1 and Figure 7.3-3 above. The tankage and blending systems allows for blending and shipping of an infinite range of differentiated S.C.O.'s and middle distillate products, with up to 150,000 barrels per batch down the Interprovincial system.

Differentiated S.C.O.'s will use the base components from the S.C.O. fractionator and the debutanizer's butane through an in-line blending system.

- Buildings

  A minor increase in laboratory capacity will be needed, complete with cetane engine for diesel testing.

7.4 PRODUCT QUALITIES

Standard S.C.O. was previously described in Table 4.3-1. Where diluent (or extraneous condensate naphtha) is added, the sulphur and nitrogen specifications of the basic S.C.O. naphtha fraction will apply - see Table 2.1-1. CGSB specifications will apply for Jet A and diesels with customers to set pour point needs for the latter. Note that static dissipator additive will be used in such products although specified only in Canada.

7.5 UTILITY BALANCES

The partial refining facilities utility estimates appear in Table 7.8-1. The estimates assume fractionation only 25% of the time, but naphtha desulphurization all the time where relevant.

7.6 CAPITAL COST ESTIMATES

The S.C.O. fractionator has been estimated from Turbo actual cost data for a 30,000 BPSD atmospheric column system, very close in other than size, to that of the Partial Refining Case. A size exponent of 0.6 was used. An air preheater system was added to the feed heater to improve fuel efficiency to the 90% level. The same time related changes were used as in the various previous cases.

Turbo's naphtha hydrotreater costs provided a similar guide to costs here. The pressure and space velocities were identical to those used here and different sizes were equated with a 0.65 exponent.

Initial catalyst charge for the hydrotreater sub case was estimated from Turbo data.

Minor allowances have been made in Basic Upgrader accounts relative to the waste water and utility aspects of partial refining.

7.7 WORKING CAPITAL ESTIMATES

Table 7.7-1 presents estimated working capital for the partial refining alternates.

7.8 OPERATING COST ESTIMATES

Table B-2 of Appendix B presents the revised operating costs using the bases previously discussed above under Basic Upgrader for the 3 options.

Table 7.7-1
Partial Refining Cases Working Capital
in 1,000's of 1Q93 Canadian Dollars

Material	Average Value	Sub Case					
		S.C.O. Frac		NHT		S.C.O. Fract + NHT	
		bbls	Value	bbls	Value	bbls	Value
Diluted Bitumen	17.58	440,000	7,735	440,000	7,735	440,000	7,735
Diluent	24.91	120,000	2,989	120,000	2,989	120,000	2,989
Intermediates/Slop	18.27	147,750	2,699	147,750	2,699	147,750	2,699
S.C.O.	26.54	225,000	5,972	530 000	14,066	225,000	5,972
S.C.O. Interm	26.54	365,000	9,687	50,000	1,327	415,000	11,014
Products	31.61 avg	170,000	5,374	-------	-------	170,000	5,374
Total			34,456		28,816		35,783

Table 7.8-1
Partial Refining Sub Cases Operating Cost Estimates
in 1,000's of 1Q93 Canadian Dollars

Onstream Factor	Sub Cases					
	S.C.O. Fract Only 25%		NHT Only 100%		S.C.O. + Fract + NHT 25/100%	
Variable Cost						
• Natural Gas (10^6 BTU/CD)	(53.2)	30,098	(52.4)	29,645	(53.9)	30,494
• Electricity (MW)	(34.6)	8,516	(35.0)	8,613	(35.6)	8,759
• Catalysts and Chemicals		8,902		9,302		9,302
• Pitch Disposal		4,022		4,022		4,022
Sub Total		51,538		51,582		52,577
Semi Variable Cost						
• Operating Labour		13,680		13,680		13,832
• Maintenance Labour		24,133		23,470		25,069
• Maintenance Materials		24,133		23,470		25,069
• Miscellaneous Operating Supplies		550		550		550
• Administration and Support		4,560		4,560		4,560
• Office Costs and Miscellaneous		550		550		550
• Insurances		3,017		2,934		3,134
• Local Taxes		6,033		5,868		6,267
• Interest on Working Capital		2,412		2,017		2,505
Sub Total		79,067		77,099		81,536
Total Operating Costs		130,605		128,681		134,113

8.0 INTEGRATED CASE

8.1 INTRODUCTION

This case considers both F-T and S.C.O. fractionation with added condensate being added to the Basic Upgrader. This case represents the maximum upgrader output per barrel of bitumen considered in this study. As will be seen, the F-T and partial refining options do not mesh significantly unless a 45 plus cetane is required in the market place, other than as upgrader hydrogen needs are provided from the F-T system.

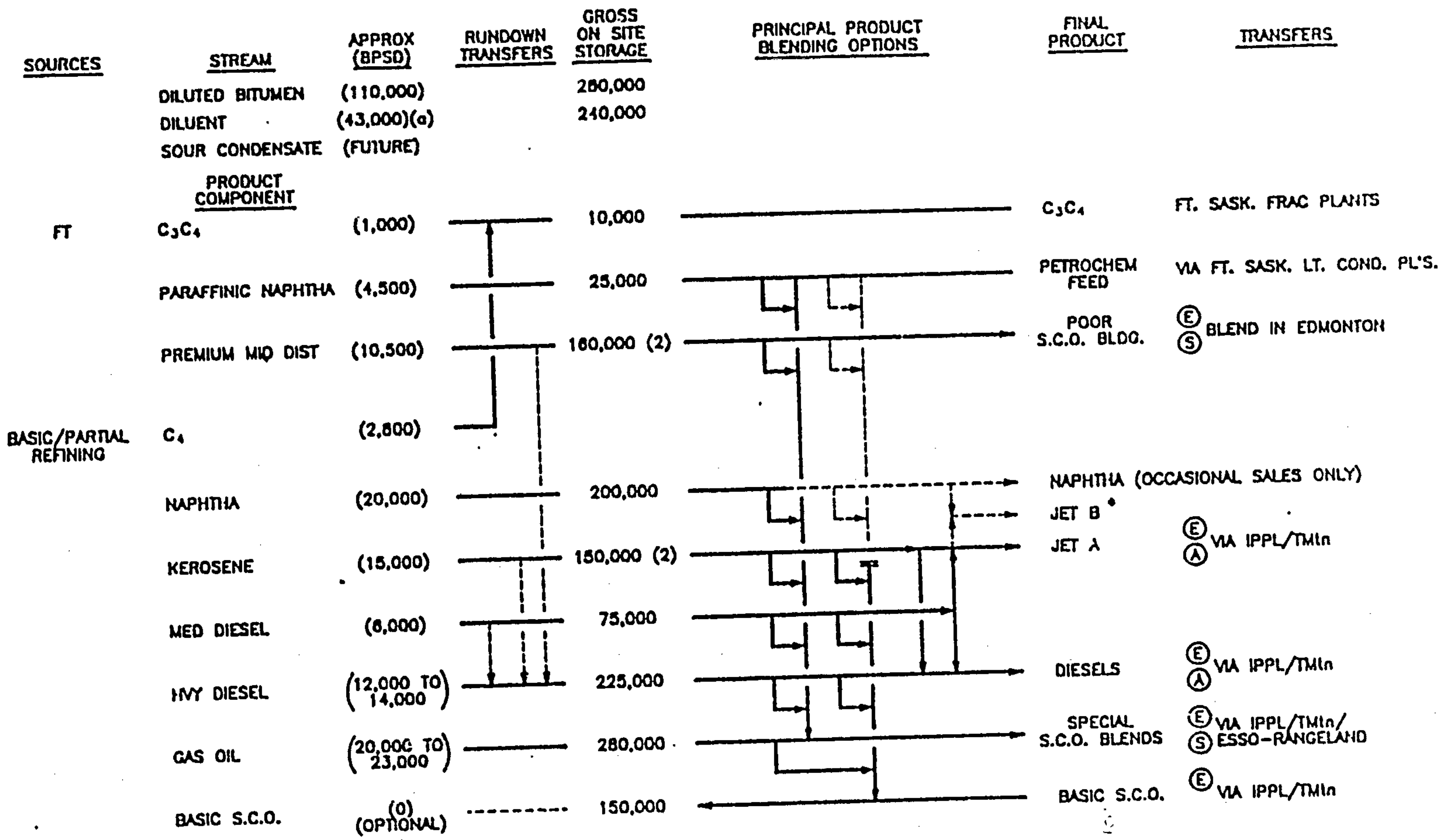
The Integrated Case, however, provides the maximum flexibility in differentiated S.C.O.'s to suit many refinery interests. Indeed, differentiated S.C.O. marketing will be a very major factor in the success of an integrated upgrader. Figure 8.1-1 provides a quick overview of the many differentiated S.C.O. options.

The business complexity of this case rises significantly as very precise product blending to meet extremely sophisticated marketing will be essential to success.

The battery limit and scope definitions continue as in the Base Case, including upgrading, associated upgrader site utilities and offsite, local pipelines and an Edmonton terminal.

8.2 DESIGN BASES

The design bases are essentially the sum of those for the F-T and partial refining with condensate bases, except for pipeline and storage as defined in Figure 8.3-1 and Table 8.3-1 below.



LEGEND

Ⓔ BATCH IN PIPELINE TO EDMONTON WITH 8 EDMONTON END TANK EACH OF 170,000 bbls. TO PROVIDE PIPELINE BATCHES TO 300,000 bbls.

Ⓢ CAN BATCH TO SHELL IN LOCAL PIPELINE

Ⓐ ADD ADDITIVES

——— PRIMARY FLOWS
------ OPTIONAL FLOWS
* NO TANK IN CURRENT SCHEME

INTEGRATED CASE PRODUCT BLENDING
FIG 8.1-1

8.3 DESCRIPTION OF DESIGN

All of the components have been reviewed above in the basic case, F-T Case and partial refining with condensate addition sub case. The proposed tankage is spelled out in Table 8.3-1 and the pipeline connections appear on Figure 8.3-1.

8.4 PRODUCT YIELDS AND PROPERTIES

Appendix C summarizes the yields for this case. Property data appears in the basic case, F-T Case and Partial Refining with condensate sub case discussions.

8.5 UTILITY BALANCES

Natural gas and electricity demands are shown in Table 8.8-1.

8.6 CAPITAL COSTS

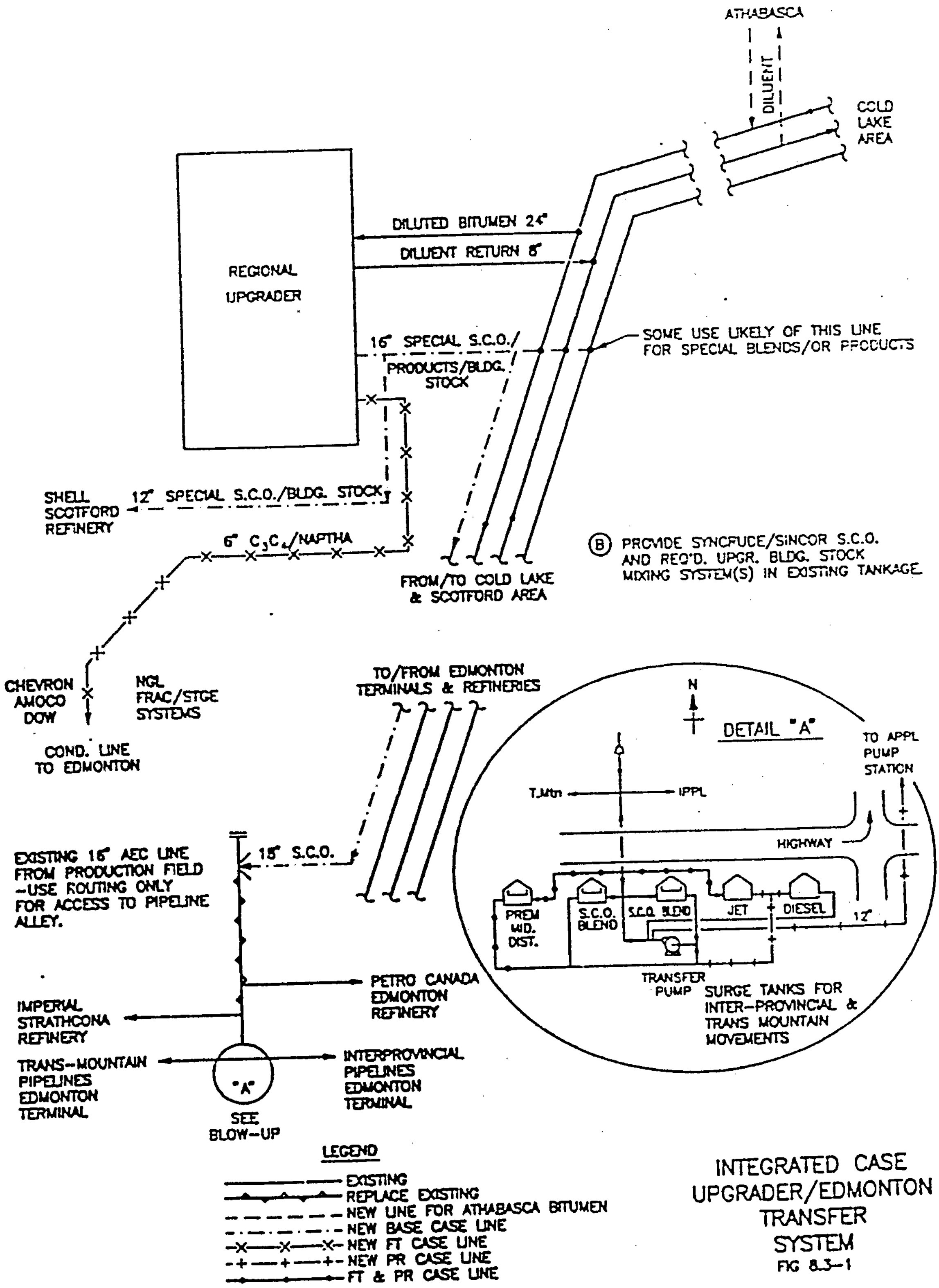
Capital cost development was discussed above relative to F-T and Partial Refining Cases and cost estimates appear in Appendix B.

8.7 WORKING CAPITAL

Table 8.7-1 presents the estimated value of average inventories.

8.8 OPERATING COSTS

Generally these costs were derived from Basic Upgrader, F-T and Partial Refining Cases.



INTEGRATED CASE
UPGRADER/EDMONTON
TRANSFER
SYSTEM
FIG 8.3-1

Table 8.3-1
Integrated Case Tankage

Where	Material	Days Storage	Use/Production Rate BPSD	No. of Tanks	Individual Tank Size Barrels	Type/ Specials
Site	Diluted bitumen	8	110,000 max	4	220,000	Cone roof c/w mixers (a)
Site	Diluent	5	43,000 max	2	120,000	Floating roof
Site	Virgin Atmos Dist	3	16,000 max	1	55,000	Cone roof - N_2 blanket - circ heat/mix system
Site	Vacuum Gas Oil	3	17,000 max	1	60,000	Cone roof - N_2 blanket - insulated - circ heat/mix system
Site	Vacuum Bottoms	3	36,000 max	1	120,000	Cone roof - insulated - circulating heat
Site	Light Sour Slop			1	10,000	Floating roof
Site	Heavy Slop			1	20,000	Cone roof - insulated - circ heating system
Site	Wet Slop			1	1,000	Cone roof - insul/heat special water draw
Site	C_4	3	3,500	1	10 000	Sphere
Site	Naphtha (Paraffinic)	5	4,000	1	25,000	Floating roof
	Prem Mid Dist	10	14,000	2	80,000	Cone roof
Site	Naphtha (Conventional)	10	20,000 (b)	1	70,000	Floating roof
Site	Kerosene	10 @ max	15,000 max	2	75,000	Cone roof
Site	Med Diesel			1	75 000	Cone roof
Site	Hvy Diesel/Diesel	7 @ max (15)	32,000 (15,000)	3	75,000	Cone roof - circulating heat system (b)
Site	Gas Oil	10	25,000	2	280,000	Cone roof - insulated - circ heating system
Site	S.C.O./S.C.O. Slops	2	72,000 max	2	75,000	Floating roof - mixers
Edmonton	Prem. Mid. Dist.	(10)	(150,000 bbl batches)	1	170,000	Cone roof
Edmonton	Jet A	(12)	(150,000 bbl batches)	1	170,000	Cone roof - mixers
Edmonton	Diesel	(5+)	(150,000 bbl batches)	1	170 000	Cone roof - mixers
Edmonton	Special S.C.O./S.C.O	(3+) @ max	(300,000 bbl batches)	2	170 000	Floating roof

Notes: (a) Provide vapour recovery system to minimize odours.

Table 8.7-1
Integrated Case Working Capital
in 1,000's of 1Q93 Canadian Dollars

Material	Average Value	Inventory	Value
Diluted Bitumen	17.58	440,000	7,735
Diluent	24.91	120,000	2,989
Intermediates/Slop	18.27	133,000	2,403
S.C.O.	26.54	245,000	6,502
S.C.O. Intermediates	26.54	645,000	17,118
Products	31.61 avg	255,000	8,061
Total		1,838,000	44,835

Table 8.8-1
Integrated Case Operating Cost Estimates
in 1,000's of 1Q93 Canadian Dollars

Variable Cost		
• Natural Gas (10^9 BTU/CD)	(241)	136,374
• Electricity (MW)	(-16.8)	-4,188
• Catalysts and Chemicals		17,030
• Pitch Disposal		4,022
Sub Total		153,238
Semi Variable Cost		
• Operating Labour		16,492
• Maintenance Labour		39,182
• Maintenance Materials		39,182
• Miscellaneous Operating Supplies		550
• Administration and Support		4,560
• Office Costs and Miscellaneous		550
• Insurances		4,898
• Local Taxes		9,795
• Interest on Working Capital		3,138
Sub Total		118,347
Total Operating Costs		271,585

9.0 FULL REFINING CASES

9.1 INTRODUCTION

This section describes the development of concepts, and capital and operating costs for the Full Refining sub cases:

a) Added to Base Upgrader
b) As a) with F-T synthesis added
c) As a) with 12,000 BPCD of diluent added
d) Base Upgrader plus 12,000 BPCD of diluent and F-T synthesis

The basic refining scheme relies on S.C.O. fractionation, catalytic cracking, C_3C_4 isomerization, TAME and light cycle oil hydrotreating processes to produce gasolines to Canadian market qualities (with some export potential for U.S. reformulated gasolines) and Jet A and diesel for Canadian and U.S. markets. A capability is provided to add MTBE to gasoline in Edmonton to provide (more) oxygen to the extent needed for reformulated gasolines for U.S. northern tier markets as far as Ohio if desired. The capability for differentiated S.C.O. sales also exists, if/as refined product sales do not match upgrading/refining capacity.

The process yields and unit costs have been developed from licensor discussions and literature, and from a detailed analysis of the Canadian Turbo Calgary refinery's initial and revamp costs and utility balances. The latter refinery's configuration includes many units essentially duplicated here (other than as to size) resulting in a high degree of confidence in capital and operating costs.

The battery limit definitions of all Full Refining Cases follows the same outline as in all other cases, including the Basic Upgrader refining units, associated upgraders site utilities and off-sites, local pipelines and an Edmonton terminal.

Maximum and minimum gasoline production levels were set for each of the 4 sub cases. In the case of minimum gasoline cases, summer gasoline vapour pressure and RVP turned out to be well over the 7.5 psi maximum in early trials, but this was lowered to 7 or less through addition of the TAME unit to convert high pressure C_5 olefins to a high octane oxygenate of much lower vapour pressure. (The minimum gasoline case with F-T and condensate processing added still has a 7.5 range RVP.) The addition of a TAME unit also provides one source of oxygenates for use in U.S. reformulated gasoline grades - it also significantly reduces olefins, probably allowing the finished gasoline to meet U.S. north and midwest reformulated specifications (with the addition of some MTBE to bring the total oxygen to the correct level) - when the late 1990 specifications are set. The refining units have not been fully optimized as that is a very major task well beyond this study. However, all yields are consistent or below expected practice in 2000 and, hence, conservative.

9.2 DESIGN BASES

The following are the design bases for each plant section:

- Capacity
  - The complex is designed to process 60,000 BPCD of bitumen.
  - In sub cases b) and d), F-T synthesis is added to produce the hydrogen needed in the complex and added naphtha and distillates.
  - In sub cases c) and d), 12,000 BPCD of diluent has been added to the feed.

- Other Feeds

  - Field butanes are acquired via a local pipeline.
  - Methanol for TAME via truck (from Celanese Edmonton).
  - MTBE if/as needed over fence at Edmonton terminal from Alberta Envirofuels.

- Crude Assay

  - The Full Refining Cases have considered only Cold Lake bitumen. However, refinery flexibility will result in essentially the same range of product slates being possible for Athabasca as for Cold Lake. The synergy of catalytic cracking and residual hydroprocessing is pronounced and results in significant flexibility in the refining yield structure.

- Synthetic Crude Oil Qualities

  - While not considered other than as a study byproduct to keep bitumen processing at a constant rates (as at NewGrade), a complete range of high quality differentiated S.C.O.'s can be produced. (See preceding Base Case table).

- Gasoline Quality

  - Pool octane of 89 road (R+M/2) and 7.5 psi maximum summer/9 annual average RVP, benzene below 0.8 volume percent, plus regular CGSB specifications are the minimum specifications. As noted above, the addition of less than 10% MTBE would bring the gasoline product to expected U.S. future national reformulated gasoline standards. Provision is made for that at Edmonton.

- MMT can be added to increase octane by about 1 octane number if desired for Canadian gasolines.
- An in-line blending system allows production of any octane grade from 87 to 93 in batches up to 150,000 barrels for regular and 75,000 for other grades. The MTBE would be batch added in Edmonton. Note that adding MTBE to export gasoline at Edmonton will save much in MTBE transportation costs, as no special handling systems would be needed. (This appears the only approach to MTBE transport to the U.S. mid west, for example, displacing a rail alternate.)

• Jet A Specification

- CGSB specifications with 21 minimum smoke point and aromatics at or below 20 percent.

• Diesel Specification

- CGSB specifications (now with 40 minimum cetane number) and pour point differentiated in in-line blending of 4 components to suit end user needs. Actual cetane number will average about 43 for the configuration shown. A consistent 45 cetane equivalent would need an added aromatic saturation unit <u>not now in scope</u> (except for the cases with F-T where the pool cetane will be approximately 47).

• Propane Quality

- Current merchant propane specifications will prevail.

- Number of Trains
  - The refining units added to the Basic Upgrader will all be single train.
- Hvdrogen Supply
  - Full on-site supply is planned using the 2 trains of the Basic Upgrader (84 million scfsd at 99.5+% purity - capacity incudes some margin over total net availability).
  - The refining units will use hydrogen from catalytic reforming preferentially with final balancing with the surplus in the upgrader's hydrogen system.
- Residue Disposal
  - See Basic Upgrader - no change but slightly greater quantity (20 tons per day range) due to processing of catalytic cracking residuals).
  - Spent catalytic cracking catalyst will go to cement or landfill (low metals content will permit).
  - All other catalysts will be returned to the vendor or his assignee when spent.
- Utility Philosophy
  - Purchase natural gas and electricity to extent needed to balance demand.
- Location
  - See Basic Upgrader.

- Service Factor

  - 90% used for all units as consistent with refinery experience. Note that differentiated S.C.O.'s will be produced whenever refined product demands are not equal to available upgrader capacity.

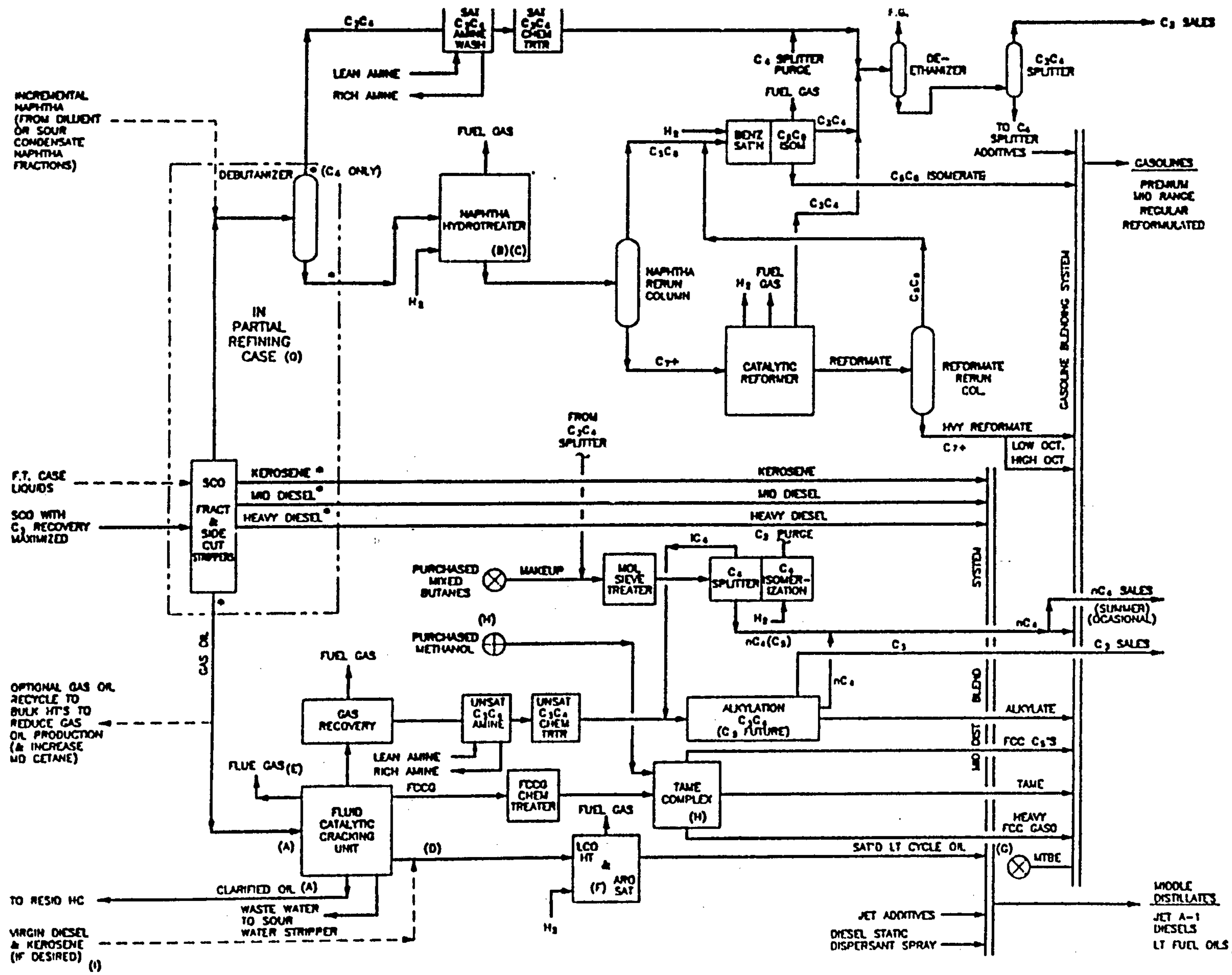
9.3 DESCRIPTION OF OVERALL DESIGN

The processing scheme in the full refining units is shown in Figure 9.3-1 with high and low gasoline product yields shown in Table 9.3-1, assuming 100 percent S.C.O. conversion to refined products. The same table also shows units throughputs with the larger size of high and low gasoline options used for costing.

The Basic Upgrader process is unchanged from the preceding discussions, except as follows:

a) Primary Upgrader

- Accepts 1,000 to 2,000 BPD of FCCU bottoms in addition to vacuum bottoms from bitumen fraction. This recycle will probably be routed to one train with some conventional feed transferred to the other train to balance operations. It is not expected to significantly impact on the co-processing of vacuum bottoms.
- Virtually all of this recycle will be recovered in S.C.O. with significant cracking and hydrogenation.
- No increase in capital or operating costs are anticipated in this study; a possible minor over simplification.



NOTES

A. 90+% RETURNS IN MIDDLE DISTILLATE & FCCU FEED. ASSUMED HERE AS LATTER.

B. 850 PSIG DESIGN (FOR CONDENSATES).

C. EXPAND NAPHTHA HYDROTREATER & UNITS DO STREAM TO HANDLE ADDED NAPHTHA WHERE APPLICABLE. CONDENSATE &/OR FT NAPHTHAS

D. FCCU LT FUEL OIL CAN GO DIRECT TO LFO, BUT VERY LOW DEMAND REQUIRES COMMON DIESEL/LFO PRODUCT IN WESTERN CANADA.

E. SLIGHT INCREASE MAY BE NEEDED IN BASIC CLAUS UNIT EFFICIENCY DUE TO SMALL AMOUN OF SO_2 IN FLUE GAS. BUT SOX TRANSFER CATALYST ADDITIVE IN FCCU SHOULD CONTROL

F. NEED FOR SEVERITY OF SATURATION MINIMAL FT LIQUIDS ADDED, BUT SATURATION VOLUME INCREASE MORE THAN OFFSETS ADDED OP CO 1000 PSIG DESIGN.

G. ADD IN EDMONTON DIRECT FROM ALBERTA ENVIROFUELS FOR REFORMULATED GASOLINE (

H. SEE FIGURE 9.3-3

LEGEND

• POTENTIAL S.C.O. COMPONENT STREAM

⊗ PURCHASED-RECEIVED BY PIPELINE

— BASIC FLOW

– – – OPTIONAL (BUT NOTE POSSIBLE EQUIPMENT SIZE CHANGES)

⊕ PURCHASED-RECEIVED BY TRUCK

FULL REFINING PROCESS SCHEME

FIG 9.3-1

Table 9.3-1
Refining Case Analysis Summary
Preliminary Unoptimized/C_3C_4 Alkylation
BPCD Annual Averages

Rates	Maximum Gasoline (f)				Minimum Gasoline (f)			
	Base	+ FT	+ Cond	Total	Base	+ FT	+ Cond	Total
Bitumen Feed	60,000	60,000	60,000	60,000	60,000	60,000	60,000	60,000
Diluent Feed	0	0	12,000	12,000	0	0	12,000	12,000
Purchased Butanes (a) (e)	2,850	1,550	2,400	1,050	1,200	0	600	0
Purchased Methanol	700	700	700	700	500	500	500	500
Propane Sales (e)	750	850	900	1,000	600	700	700	800
Butane Sales (a)	0	0	0	0	0	200	0	800
Gasoline (89 oct min) (9# average RVP)	32,428	37,736	42,077	47,603	23,373	27,907	31,952	36,831
Middle Distillates (40 plus cetane) (21 plus smoke on Jet A-1)	32,656	41,473	33,616	42,433	39,015	48,688	40,935	50,607
G/D	.99	.91	1.25	1.13	0.60	0.57	0.78	0.73
Principal Unit Throughputs, BPCD (b) (c)								
• C_5C_6 Isomerization	4,700	7,000	9,900	12,100	4,500	6,700	9,600	11,800
• Catalytic Reforming	7,100	12,400	13,800	19,100	5,300	9,800	11,000	15,500
• FCCU (Catalytic Cracking)	24,400	24,400	24,400	24,400	24,400	24,400	24,400	24,400
• Alkylation (C_3C_4)	6,900	6,900	6,900	6,900	4,600	4,600	4,600	4,600
• TAME	2,300	2,300	2,300	2,300	1,900	1,900	1,900	1,900
• L.C.O. Hydrotreater	4,600	4,600	4,600	4,600	9,100	9,100	9,100	9,100
• C_4 Isomerization (Conversion) (d)	3,700	3,700	3,700	3,700	2,500	2,500	2,500	2,500

Table 9.3-1
Refining Cases Analysis Summary
Preliminary Unoptimized/C_3C_4 Alkylation
BPCD Annual Averages
(continued)

Notes:

(a) Varies through year.
(b) These rates should be corrected to stream day bases at a service factor of 0.9.
(c) In all cases, the maximum gasoline sub case will set unit sizes of all but the L.C.O. hydrotreater, whose hydraulic capacity will be set by the minimum gasoline case (reactor capacity need is probably unchanged).
(d) The associated de-IC_4 column will have a capacity of roughly 2.5 times the conversion capacity.
(e) C_3's entering with purchase butanes neglected in this analysis.
(f) In the maximum gasoline case, the FCCU operates at high severity and the naphtha/distillate cut points are at about 350°F. In the minimum gasoline case, the latter reduces to 300°F.
(g) The gasoline pool will meet 1995 U.S. reformulated standards except for oxygenates, but 10% MTBE can be added in Edmonton tankage if desired. However, olefins are well above 1996 California standards in all the sub cases noted - future addition of C_5 to alkylation will decrease olefins in gasolines but not to California levels. A 7.5 psi RVP is attainable in summer months (but is tight!), especially in the fuel sub cases.

b) Secondary Upgrader

- Recycle product hydrogenation will add slightly to hydrogen needs.
- Added flexibility in degree of gas oil hydrocracking and of middle distillate aromatic saturation would be desirable but these must be addressed in further pilot and study work.

c) Hydrogen Production

- Up to 80 percent of the 6 percent capacity margin may be needed in the refining units, depending upon the severity of gasoline reforming.
- Minor additional funds are allowed for compression of reformer byproduct hydrogen to 1,000 psig for LCO hydrotreater use, to 700 psig for naphtha hydrotreater, and to 400 psig for isomerization and C_3C_4 butadiene saturation.

d) Sulphur Plant

- Added production of up to 9 tpd of sulphur is anticipated to have no material impact on the 2 sulphur trains, as they have 50% spare capacity with both on-line.
- Very minor - 1 tpd - increase in sulphur plant recovery efficiency is needed, due to recovery from FCCU off gases. Overall recovery must be increased slightly to offset minor amount of SO_2 in FCCU flue gas.

e) Sour Water Stripper

- There will be an added load due to the FCCU overhead waters containing traces of H_2S and phenols. Allowances have been made for this added load in capital and operating costs. (A second SWS would allow reuse of FCCU water, but such is not assumed.)

f) Waste Water

- The F-T sub case has its own waste water treatment system, but the FCCU introduces some added waste streams from sour water stripping and 3 small dilute caustic wash systems.

g) Steam and Power Plant

- An additional boiler is added to provide enough 600 psig superheated steam for the FCCU air blower and the reformer recycle compressor steam turbine. (In the F-T sub cases, superheated 150 psig would be used for these services, partly in lieu of electricity generation, and no new boiler added.)

h) Tankage

- Tankage is planned for each sub case as set out in Table 9.3-2.
- Product blending will be via in-line computer controlled systems using on-line RVP and octane controls for gasoline.
- Figure 9.3-2 sets out the planned pipeline transfer systems - the only change between sub cases being a slightly larger pipeline in the combined F-T plus condensate sub case.

i) Raw Water Treatment

- Refinery units will be totally air cooled, hence, no changes will be needed relative to cooling water. A minor amount of added boiler feed water make-up will be needed along with expanded condensate systems. The FCCU's steam turbine exhaust steam will be condensed with air (as at Turbo).

j) Other Off-Sites

- Additions to suit refinery areas and processes in instrumenting fuel system, flare, fire protection, etc., will be needed.

k) Interconnecting Pipeways

- For each sub case, allowances have been made for refining related piperacks.

l) Common Buildings

- Extensions to the buildings of the Basic Upgrader have been allowed to handle added technical and administrative staff, added laboratory facilities and added "first aid" type maintenance.

Table 9.3-2
Full Refining Case Tankage
Sizes in 1,000's of Barrels

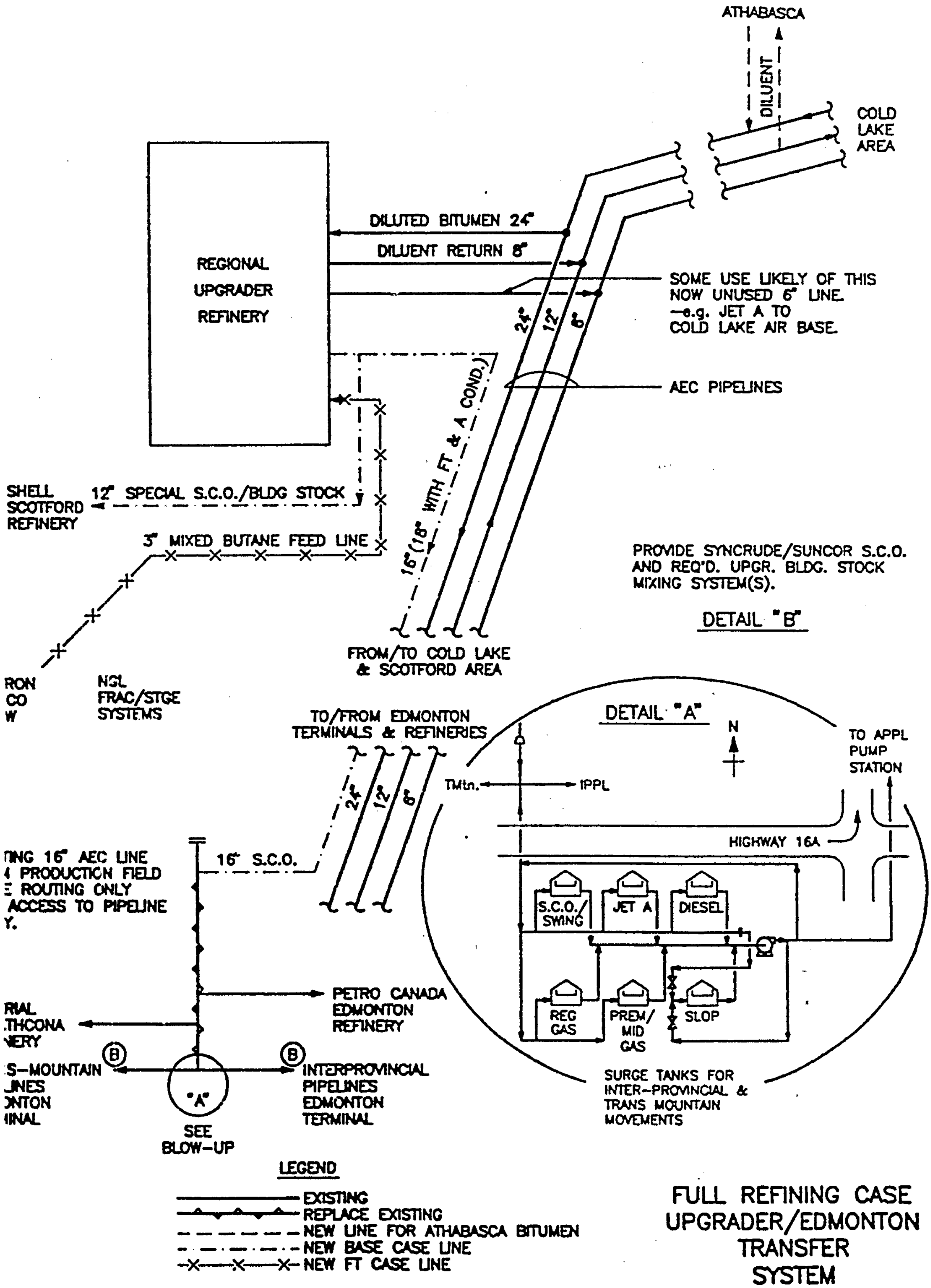
Where	Service	Sub Cases				Type
		Basic	+ Condensate	+ F-T	Total	
Site	Diluted Bitumen	(4) @ 220	4 @ 220	4 @ 220	4 @ 220	CR CW mixers (a)
Site	Diluent/Raw Naphtha	(2) @ 120	(2) @ 120	(2) @ 120	(2) @ 120	FR
Site	Virgin Atmos Dist	(1) @ 55	(1) @ 55	(1) @ 55	(1) @ 55	CR, Blkt, Ht, Mix
Site	Vacuum Gas Oil	(1) @ 60	(1) @ 60	(1) @ 60	(1) @ 60	CR, Circ Heat
Site	Vacuum Bottoms	1 @ 120	1 @ 120	1 @ 120	1 @ 120	CR, Ins, Circ Ht
Site	Light Slop	2 @ 10	2 @ 15	2 @ 15	2 @ 15	FR
Site	Heavy Slop	1 @ 40	1 @ 40	1 @ 40	1 @ 40	CR, Ins, CH+
Site	Wet Slop	1 @ 2	1 @ 2	1 @ 2	1 @ 2	CR, Ins, CH+
Site	Propane	3 @ 5	3 @ 5	3 @ 5	3 @ 5	Bullets
Site	n-Butane	1 @ 10	1 @ 10	1 @ 10	1 @ 10	Sphere
Site	iso-Butane	1 @ 10	1 @ 10	1 @ 10	1 @ 10	Sphere
Site	Naphtha (HT)	1 @ 30	1 @ 30	1 @ 30	1 @ 30	FR
Site	Kerosene	2 @ 75	2 @ 75	2 @ 90	2 @ 90	CR
Site	Medium Diesel	1 @ 75	1 @ 75	1 @ 100	1 @ 100	CR
Site	Heavy Diesel	3 @ 75	3 @ 75	3 @ 80	3 @ 80	CR
Site	Gas Oil	1 @ 140	1 @ 140	1 @ 140	1 @ 140	CR
Site	S.C.O./S.C.O. Slops	2 @ 75	2 @ 75	2 @ 75	2 @ 75	FR c/w mixers
Site	Light Cycle Oil	2 @ 50	2 @ 50	2 @ 50	2 @ 50	CR
Site	C_5C_6 Isomerate	1 @ 40	1 @ 100	1 @ 80	2 @ 70	FR
Site	Low Octane Ref	1 @ 25	1 @ 60	1 @ 60	1 @ 100	FR
Site	High Octane Ref	1 @ 15	1 @ 30	1 @ 30	1 @ 40	FR
Site	Alkylate	1 @ 80	1 @ 80	1 @ 80	1 @ 80	FR
Site	FCC Gasoline	2 @ 70	2 @ 70	2 @ 70	2 @ 70	FR
Site	Gasoline Blending	4 @ 80	4 @ 80	4 @ 80	4 @ 80	FR

Table 9.3-2
Full Refining Case Tankage
(continued)

Where	Service	Sub Cases				Type
		Basic	+ Condensate	+ F-T	Total	
Site	Methanol	1 @ 5	1 @ 5	1 @ 5	1 @ 5	FR
Site	TAME (c)	1 @ 20	1 @ 20	1 @ 20	1 @ 20	FR
Edmonton	Jet A	1 @ 170	1 @ 170	1 @ 170	1 @ 170	CR
Edmonton	Diesel	1 @ 170	1 @ 170	1 @ 170	1 @ 170	CR
Edmonton	Gasoline (d)	1 @ 170	1 @ 170	1 @ 170	1 @ 170	FR mixers
Edmonton	Prem/Mid Range (d)	1 @ 80	1 @ 80	1 @ 80	1 @ 80	FR mixers
Edmonton	S.C.O./Swing	1 @ 170	1 @ 170	1 @ 170	1 @ 170	FR
Edmonton	Slop/Interface	1 @ 10	1 @ 10	1 @ 10	1 @ 10	FR

Notes:

(a) Vent gas processing for odour control.
(b) Provision for future MTBE addition for Alberta Envirofuels.
(c) Provide for truck receipt.
(d) Provide for MTBE addition.



FULL REFINING CASE
UPGRADER/EDMONTON
TRANSFER
SYSTEM
FIG 9.3-2

Following Figure 9.3-2, the following units will be needed for converting S.C.O. to gasolines and diesels:

a) S.C.O. Fractionator

- This is identical to that of the Partial Refining Case, except that in sub cases with F-T synthesis, deethanized dewax/saturator product will move directly to this fractionator to separate F-T kerosene and other middle distillates along with S.C.O. materials.

  The end point of the overhead naphtha will be controlled in this tower, setting the volume going to catalytic reforming.

b) Naphtha Debutanizer

- In the refining cases as much propane as possible should be recovered upstream in the Basic Upgrader. This will maximize C_4 recovery as well as provide a marginally attractive addition to the propane produced in the refining units in any case. The debutanizer is planned to minimize C_3C_4's in naphtha hydrotreater feed.

c) Naphtha Hydrotreater

- In the basic sub case this unit processes only deeply hydrotreated feedstock, but such a unit is still considered essential to prevent excursions of sulphur and nitrogen reaching downstream catalysts.

- When diluent (or extraneous condensate) is added[6], this unit's duty increases appreciably as it must reduce sulphur to below 0.5 ppm and nitrogen even lower. In order to allow for variations in diluent (and/or condensate) composition, a 650 psig unit is planned well above light crude refinery standards (except at Turbo where sour condensates were contemplated). A space velocity of approximately 2 is anticipated with a nickel molybdenum catalyst. Even with such a design, this unit will have a nickel catalyst guard bed on the product full range naphtha going to splitting to provide even more assurance of downstream catalyst protection.

d) Naphtha Splitter

- This will be a conventional distillation system providing an accurate separation between C_5C_6 feed to the isomerization unit and the catalytic reforming unit. It requires a fired reboiler.

e) Catalytic Reforming

- The balances developed in this study indicate a low severity required compared to most North American refinery reformer operations, due to the octane contributions of other gasoline components. However, some diluent (and condensates) have poor reforming octane producing characteristics - e.g. as measured by naphthenes plus 2 times aromatics - and F-T naphtha is as bad as can be found. Note that the low severity operation will significantly reduce gasoline aromatic content.

---

6 The assumed availability of the AEC 6" line from Edmonton permits access to some heavy sour condensates if desired. Use of such condensates will require an extra small distillation unit to control the end point of material going to the naphtha hydrotreater.

Thus, the basic sub case has a four reactor semi regenerative design - rather old fashioned but low cost - and the other sub cases add a cycling capability to the last two reactors is proposed. The off-line reactor is then regenerated as in more conventional cyclic reforming units (as Imperial's Strathcona unit).

The hybrid system was used successfully at Shell's Boniface refinery 20 years ago and is an ideal fit for the low average severity needed here, permitting high severity short runs as required. This approach does not optimize hydrogen production and has a slight yield disadvantage compared to continuous catalytic reforming, but the penalties are considered less important here than lower capital and operating costs.

f) Reformate Splitting

- The reformer's product will contain benzene from the feed (minor) and reformer reactions. In order to control benzene in the overall gasoline pool, the reformate is separated into a C_5C_6 stream (of relatively low octane) for further processing in C_5C_6 isomerization and a high octane aromatics rich stream.

g) C_5C_6 Isomerization

- The naphtha and reformate splitter overhead streams, after drying along with a small amount of hydrogen, go first to a benzene saturation step and then to fixed bed precious metal isomerization reaction. (A very acidic reaction environment is used, maintained with chloride injection. Caustic scrubbing of the off gas is used to capture all HCl produced.) The product is stabilized before going to the gasoline pool.

h) Catalytic Cracking

- The very low sulphur and nitrogen contents and the highly hydrogenated nature of the S.C.O. gas oil creates an ideal Fluid Catalytic Cracking Unit (FCCU) feedstock. Due to low coke expected on the circulating catalyst, a fired feed heater is planned to ensure sufficient heat for reaction. A very short residence reaction system is planned for the maximum gasoline case, but lower temperatures at some residence time may have to be provided for times when middle distillates are to be maximized. However, in this study the minimum gasoline case is largely defined by reducing the end point of the FCCU gasoline. This reduces the aromatics going to gasoline - and those backed out are largely higher aromatics, undesirable in reformulated gasoline in any case.

  The fractionator of the FCCU produces a heavy bottoms fraction with some catalyst fines to be recycled to the primary upgrading units, light cycle oil to hydrotreating, and light ends to the gas concentration system.

i) FCCU Gas Concentration

- This standard absorption/desorption system will capture over 93 percent of propylene and virtually all heavier materials from the FCCU gases. Lighter components will pass through an amine wash for removal of traces of H_2S before going to fuel gas. An olefinic C_3C_4 stream is forwarded to alkylation and C_5 plus components go to the TAME complex.

j) Alkylation

- The olefinic C_3C_4's are lightly hydrotreated to saturate traces of butadiene in the top of the gas concentrations debutanizer before going to an HF alkylation unit. This unit will be of a new ultra low HF hold-up design to minimize potential for significant HF releases under any circumstances. HF is selected due to the much lower catalyst make-up rate and no need for expensive, environmentally sensitive regeneration compared to an H_2SO_4 approach. (The 2 Edmonton refineries each have large older design HF C_3C_4 alkylation units.)

  Propane and n-butane are fractionated out of the alternate product for sale and use in gasoline blending, respectively. Make-up iso-butane is supplied from the C_4 isomerization unit.

k) C_4 Isomerization

- This study concluded it would be difficult to continually purchase enough iso-butane for alkylation. (The existing MTBE plant, for example, indicated little or no surplus i-butane and there are no regional field butane splitters.) Hence, to supplement mixed C_4's from reforming and saturated C_3C_4 treating, field butanes will be purchased for feed to a C_4 splitter, integrated with an n-C_4 isomerization unit. This approach, as practised in the 2 Edmonton refineries, minimizes C_4 purchases and maximizes profits. The isomerization unit product has about 55 percent i-butane and recycles through the de-iso-butanizer. A purge of C_3 rich material will be routed to the saturate C_3C_4 processing system to dispose of C_3 entering with the field butanes. C_5's in the field butanes will be purged out the bottom of the deisobutanizer along with n-butane needed for gasoline blending.

l) TAME Complex

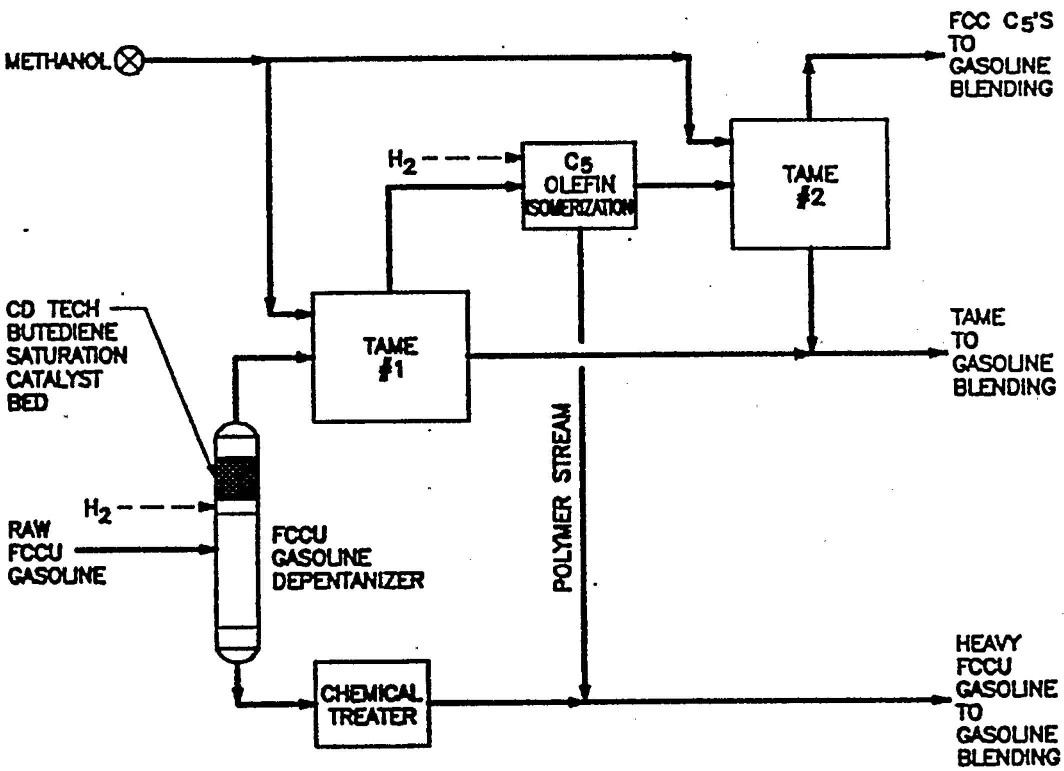
- The two TAME systems will react methanol with certain C_5 iso-olefins to produce Tertiary Amyl Methyl Ether, with an intermediate isomerization step to convert other C_5 olefins to reactive species. CD Tech technology is assumed here for TAME along with Lyondells new C_5 isomerization process. A prior depentanizer is required on the C_5 plus FCC gasoline stream.

  Figure 9.3-3 outlines this system which became essential to control summer time RVP to reasonable levels. (In F-T Cases, a small amount of C_5C_6 isomerization feed may still have to be rejected to hold a 7.5 psi RVP level in the final gasoline.) This route is less expensive and safer than adding C_5 alkylation capacity - it also increases gasoline volume the least while adding oxygen to the final gasolines - to about the 0.8 weight percent level - one-third or so of U.S. reformulated gasolines.

  The system shown also eliminates from final gasoline, 85 percent or so of C_5 olefins which are the most reactive in producing smog - another environmental plus as well as perhaps a necessity in future reformulated gasolines.

m) Heavy FCC Gasoline Treating

- In order to remove traces of mercaptans, a small fixed bed treater will oxidize these to disulphides using a small amount of air, using UOP or Merichem technology.



TAME COMPLEX
FIG 9.3-3

n) Light Cycle Oil Hydrotreater

- This unit is not needed to meet the new diesel sulphur 0.05% sulphur specification, but rather to saturate traces of olefins and significant quantities of aromatics.

The LCO hydrotreater is envisaged as a low space velocity 900 psig unit using initially a single aromatics saturation, non precious metal catalyst capable of coping with 100 ppm of sulphur in the feed. However, the unit will be designed to permit future conversion to a two stage system desulphurization to the 10 ppm level, followed by precious metal catalyst aromatic saturation with intermediate H_2S withdrawal as in the Criterion/CE Lummus SynSat process. Note that the liquid yield was estimated here at 105 percent of feed - saturation costs are more than paid for in yield gain.

The LCO product is expected to have a cetane number in the order of 35 to 40 in the initial configuration, sufficiently high to achieve over 40 cetane in all diesel blends. This is sufficient with F-T to achieve 45 but added middle distillate aromatic saturation is needed in all other cases if a 45 cetane specification becomes fact.

m) Saturate Gas Processing

When diluent or extraneous condensates are to be processed, the C_3C_4 from the de-butanizer will contain sulphur compounds requiring a small chemical treating system consisting of amine wash followed by a chemical treater of the Merox type. The treated C_3C_4's will be combined with C_3 purge from the C_4 splitter, a small C_4 and lighter liquid stream from the C_5C_6 isomerization unit and raw C_3C_4 from the reformer's stabilizer before deethanizing (to fuel gas) and depropanizing with propane joining propane from the alkylation unit for sale.

The depropanizer bottoms go to the C_4 splitter to maximize internal iC_4 recovery and provide $n\text{-}C_4$ for isomerization.

9.4 PRODUCT YIELDS

The expected annual average yields in the various full refining sub cases are shown in Table 9.3-1. Specifications for products were noted above. Note that no S.C.O. product is shown - only gasoline and middle distillates.

9.5 UTILITY BALANCES

Table 9.8-1 summarizes the average natural gas and electricity demands for the various full refining sub cases.

9.6 CAPITAL COST ESTIMATES

The capital cost estimating approaches for this study were discussed in Section 4 above. The following notes refer specifically to the refining related units and tankage system. Capital cost estimates are shown in Appendix B-1.

a) Revision in Basic Upgrader

Allowances were added relative to minor changes in sour water stripping, waste water treating, steam plant, other sites and common buildings to cover revisions noted in Section 3 above. The piperacks account was increased based on preliminary estimates.

Table 9.4-1
Gasoline and Middle Distillate Pool Composition Estimates
BPCD Unless Noted

	Maximum Gasoline				Minimum Gasoline			
	Base	+ F-T	+ Cond	Total	Base	+ F-T	+ Cond	Total
Gasoline Pool Composition (Annual Average) (a)								
• Reformate	4,825	7,760	9,368	12,534	3,385	5,721	7,040	9,659
• Isomerate	4,642	6,863	9,674	11,903	4,450	6,587	9,440	11,627
• FCC Gasoline (b)	12,218	12,218	12,218	12,218	8,126	8,126	8,126	8,126
• C_3C_4 Alkylate	6,903	6,903	6,903	6,903	4,560	4,560	4,560	4,560
• TAME	2,333	2,333	2,333	2,333	1,909	1,909	1,909	1,909
• n-Butane	1,506	1,659	1,580	1,712	943	1003	876	950
Total	32,428	37,736	42,077	47,603	23,373	27,906	31,952	36,831
(RVP before n-butane)	6.1	6.3	6.7	6.8	6.4	6.8	7.3	7.5
(Reformer Severity - RON)	87.0	89.6	91.6	92.6	87.0	91.6	93.4	94.0
Middle Distillate Composition (c)								
• Straight Run S.C.O.	28,098	28,098	28,098	28,098	29,898	29,898	29,898	29,898
• F-T Middle Distillate	0	8,817	0	8,817	0	9,673	0	9,673
• Condensate Middle Distillate	0	0	960	960	0	0	1,920	1,920
• HT Light Crude Oil	4,558	4,558	4,558	4,558	9,117	9,117	9,117	9,117
Total	32,656	(d) 41,473	33,616	(d) 42,433	39,015	(d) 48,688	40,935	(d) 50,607
Gasoline/Diesel	.993	.910	1.252	1.122	.599	.573	.781	.728

Notes:

(a) 89 road octane, 9 psi RVP annual average, CGSB specifications, +0.8 wt % oxygen, benzene <0.8 wt %, no MMT.
(b) Includes some polymer from the C_5 olefin isomerization process.
(c) Kerosene fraction suitable for Jet A, whole boiling range >40 cetane number. Overall pour point in -30°C range except -25°C in F-T Cases.
(d) Pool cetane >45.

Tankage was estimated for each sub case using Kilborn model (consistent with all other cases).

b) Refining Process Units

The following unit costs were scales off actual Turbo costs, with minor corrections for differences in scope: S.C.O. fractionator, naphtha hydrotreating and splitting, FCCU and gas concentration, catalytic reforming, C_5 isomerization and light cycle oil hydrotreating (actually processing a blend of LCO and virgin distillates at Turbo). The unit capital costs for other units were developed from CD Tech data (TAME complex) and file data (splitters, C_4 isomerization, C_3C_4 alkylation) checked with recent literature reference.

9.7 WORKING CAPITAL

Table 9.7-1 summarizes working capital estimates for all cases, based on 50% of tankage being full.

9.8 OPERATING COST ESTIMATES

Table 9.8-1 summarizes operating cost estimates based on the bases noted previously in the Basic Upgrader section, with the following changes in the Full Refining Cases:

a) Gasoline, Jet A and diesel additives were estimated. Note that MMT - gasoline manganese additive - has not been used due to possible gasoline export (not accepted in the U.S.) and lack of need for incremental octane.

b) Various overhead accounts were adjusted to cover added operational overheads, administrative support, and marketing changes as follows:

	60,000 BPCD Other Cases	Refining Cases
• Miscellaneous Operating Expenses	0.55 million	0.8 million
• Administration and Support	4.56 million	6.9 million
• Office and Miscellaneous	0.55 million	1.0 million

Table 9.7-1
Full Refining Case Working Capital
in 1,000's 1Q93 Canadian Dollars

Material	Average Value	Base		+ Cond		+ F-T		Total	
		Inventory	Value	Inventory	Value	Inventory	Value	Inventory	Value
Diluted Bitumen	17.58	440,000	7,735	440,000	7,735	440,000	7,735	440,000	7,735
Diluent/Raw Naphtha	24.91	120,000	2,989	120,000	2,989	120,000	2,989	120,000	2,989
Intermediates/Slop	18.27	148,500	2,713	153,500	2,804	153,500	2,804	153,500	2,804
Butane	15.35	10,000	154	10,000	154	10,000	154	10,000	154
Propane	14.70	2,500	37	2,500	37	2,500	37	2,500	37
S.C.O. Intermediates	26.54	160,000	4,246	160,000	4,246	160,000	4,246	160,000	4,246
Distillate Blendstocks	31.50	275,000	8,663	275,000	8,663	310,000	9,765	310,000	9,765
Gasoline Blendstocks	31.60	320,000	10,112	375,000	11,850	365,000	11,534	420,000	13,272
Methanol	21.00	2,500	52.5	2,500	52.5	2,500	52.5	2,500	52.5
Products	31.61 avg	295,000	9,325	295,000	9,325	295,000	9,325	295,000	9,325
S.C.O.	26.54	90,000	2,389	90,000	2,389	90,000	2,389	90,000	2,389
Totals		1,863,500	48,415.5	1,923,500	50,244.5	1,948,500	51,030.5	2,003,500	52,768.5

Table 9.8-1
Full Refining Sub Cases Operating Cost Estimates
in 1,000's of 1Q93 Canadian Dollars

	Base		Base + F-T		Base + Cond		Total	
Variable Cost								
• Natural Gas (10^9 BTU/CD)	(57.8)	32,700	(241.1)	136,374	(58.5)	33,096	(241.1)	136,374
• Electricity (MW)	(51.6)	12,649	(-0.8)	-247	(52.6)	12,892	(0.2)	65
• Catalysts and Chemicals		13,500		21,130		14,600		22,230
• Pitch Disposal		4,022		4,022		4,022		4,022
Sub Total		62,871		161,279		64,610		162,691
Semi Variable Cost								
• Operating Cost		16,568		18,620		16,720		18,620
• Maintenance Labour		30,761		44,689		31,715		45,410
• Maintenance Materials		30,761		44,689		31,715		45,410
• Miscellaneous Operating Supplies		800		800		800		800
• Administration and Support		6,900		6,900		6,900		6,900
• Office Costs and Miscellaneous		1,000		1,000		1,000		1,000
• Insurances		3,845		5,586		3,964		5,676
• Local Taxes		7,600		11,172		7,929		11,352
• Interest on Working Capital		3,389		3,572		3,517		3,694
Sub Total		101,714		137,028		104,260		138,862
Total Operating Costs		164,585		298,307		168,870		301,553

10.0 SPECIAL CASES

10.1 PREAMBLE

This study has briefly examined 3 other cases to determine whether there is a fit for a) purchased hydrogen, b) partial oxidation of natural gas for F-T synthesis, and c) a methanol based route in lieu of F-T to synergistic incremental products via natural gas conversion.

10.2 BUY OR MAKE HYDROGEN

The cost of acquiring byproduct hydrogen at 70% of Base Case requirements was developed in section 2.2 above. The $1.56 per 1000 scf of hydrogen valuation included the following components:

• Purchase of Raw Byproduct Hydrogen (gas replacement + seller's margin)	0.56
• Pipeline Charges	0.06
• Other Facilities Operating Costs	0.39
Sub Total	1.01
• Other Facilities Capital	0.55

The "other facilities" capital totals approximately $90 million in 1Q93 dollars. Analyzing Base Case data provides the comparison as noted in Table 10.2-1.

Thus, a very simple R.O.I. of 15% before tax is indicated. The differential capital and operating costs could well vary significantly from the above. Hence, these return estimates must be considered as very approximate. Increasing natural gas costs will only very marginally change the above total operating cost differential in favour of the "buy" option, due to fuel needs in steam methane reforming.

There is not enough byproduct hydrogen available to provide all the hydrogen for the upgrader, even with a storage buffer. Full output is needed from one of the large suppliers when the other is off-line, hence, the supply system must have 100% capability of supply from each. There is a very significant cost risk to the upgrader if both major suppliers are out at the same time. This fact would probably sway the upgrader operator to have 100 percent on-site capacity.

While this study has looked in some detail at on-site hydrogen production facilities, it has not examined the 1990 recovery facility concepts to see if new technologies and/or more cost competitiveness would result in lower costs and/or higher yields at the same cost.

10.3 PARTIAL OXIDATION FOR SYNTHESIS GAS

In Malaysia, Shell's natural gas based F-T facility will be using partial oxidation to produce the 2 H_2/ 1 CO synthesis gas needed for F-T synthesis. They will also have a small steam methane hydrogen unit for F-T product finishing. The larger South African natural gas based F-T project appears to be using the same route. But such a route would dissaggregate the F-T system from the upgrader.

Both Texaco and Shell provide natural gas partial oxidation technologies, as simpler versions of their well proven coal and oil technologies for hydrogen and/or synthesis gas. Unfortunately Texaco require a fee for provision of preliminary data, and Shell did not respond. Published data for a small Texaco case has been used to develop approximate capital costs.

Partial oxidation results in appreciably less CO_2 and CH_4 in the raw synthesis gas, allowing elimination of the CO_2 recovery system in the F-T complex as defined in Section 6 above, as seen in Table 10.3-1. Also, the purge gas volume is significantly less. Partial oxidation also permits higher pressure F-T synthesis operation, reducing recycle compression and piping and catalyst/reactor volumes.

At 650 tpd, an on-site oxygen plant could be considered, but an over the fence supply appears more realistic in the region of the upgrader, given large oxygen demands nearby for ethylene oxide production. There is already a large air separation unit within 10 kilometers, and another train and a pipeline there would be less expensive than a new unit in the upgrader. At the central air separation plant, there will be liquid oxygen surge storage as well as added surety of supply as there are several separation trains. But concerns regarding strikes and other upgrader uncontrolled oxygen supply outages must be addressed in future studies.

The partial oxidation cost is very preliminary and requires checking. However, from the very preliminary data of Table 10.3-2, it can be seen that a partial oxidation route appears better economically than the steam methane reforming for the upgrader plus F-T synthesis, and probably better. The CO_2 and hydrogen recovery steps are a major contributor to the SMR route's costs.

10.4 METHANOL VERSUS FISCHER-TROPSCH

In New Zealand, gasoline is produced from natural gas via methanol as an intermediate. While Mobil offers a route to approximately 60/40 gasolines/middle distillates via methanol, they consider F-T a more viable alternate when maximum middle distillates are desired as here. While there may appear to be synergy between a methanol route and the Full Refining Cases TAME unit's methanol needs, the latter uses only about 3 percent of the methanol needed to match the F-T synthesis considered above. Even a world scale MTBE plant would need only about 13 percent.

Table 10.2-1
Buy or Make Hydrogen Cost Bases
in Millions of 1Q93 Canadian Dollars

	Make	Buy 70%	Differential
Natural Gas Purchase	29.2	8.9	-20.3
Hydrogen Purchase	----	22.7	+22.7
Hydrogen Pipeline	----	3.0	+3.0
Hydrogen Recycle Operation	----	14.0	+14.0
Hydrogen Production Operation	3.0 approx.	2.0	-1.0
Total Operating Costs	32.2	50.6	+18.4
Related Capital Costs (Total)	246	123	-123

Table 10.3-1
Dry Raw Syn Gas Compositions

Component	Feed to PSA in F-T Case After CO_2 Removal	Typical Partial Oxidation Gas (b)
H_2	69.43	47.9
CO	23.29	47.9
CO_2 (a)	0.25	3.5
CH_4	6.74	<0.3 est
N_2	0.29	0.4

Notes:

(a) Note that CO_2 recycled to reformer in this case.
(b) Assumes 99.5% purity oxygen used.
(c) Ratio can be adjusted to increase H_2/CO ratio by addition of steam. 0.2 mols of steam to mol of feed gas assumed in this ratio. CO_2 addition can be used to decrease ratio.

Table 10.3-2
Synthesis Gas
Partial Oxidation Versus Steam Methane Reforming
in 1,000's of 1Q93 Canadian Dollars

	This Study Bases SMR	Partial Oxidation	Differential
Hydrogen Production (SMR/PSA)	----	116	+116
Steam Methane Reformers c/w Catalyst	220	----	-220
CO_2 Recovery	33	----	-33
H_2 Separation	36	----	-36
F-T Syn Loop Changes	Base	Base -15	-15
Partial Oxidation for Syn Gas Only	----	140 (c)	+140
Oxygen Supply (over fence)	----	+10	+10
Total Direct Field Expense Only	289	256	-23
Operating Costs			
• Natural Gas	Base	Base -20	-2.0
• Electricity	Base	Base -1.5	-1.5
• Oxygen @ $36/tonne delv'd (b)	----	7.7	+7.7
• Catalysts and Chemicals	Base	Base -1.0	-1.0
• Maintenance	Base	Base -3.0	-3.0
• Other Operating Costs	Base	Base -3.0	-3.0
Total	Base	Base -2.8	-2.8

Notes:

(a) 10% added for F-T product finishing needs (to Base Case costs), but this is offset by savings in upgrader hydrogen compression costs.
(b) Praxair preliminary estimate.
(c) This is very preliminary and requires checking.

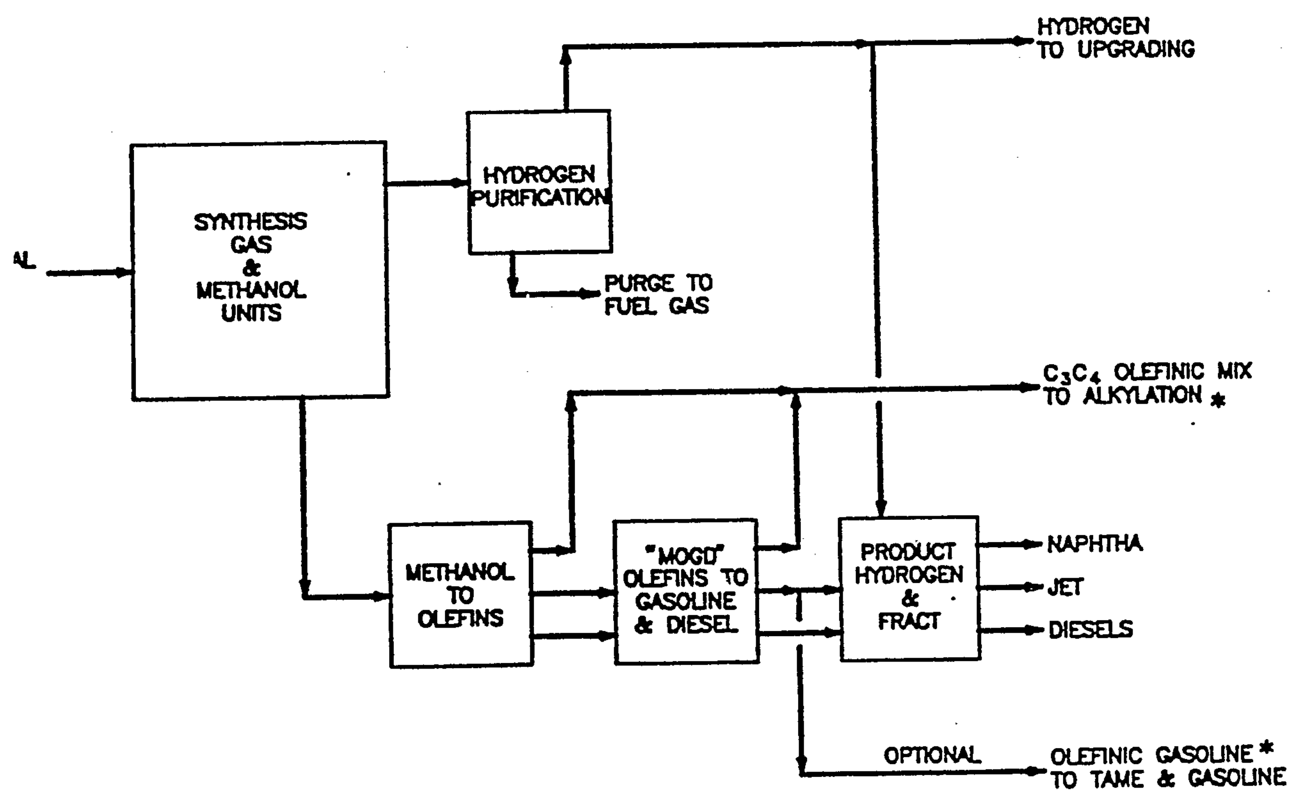
A preliminary analysis indicates that the hydrocarbon liquids production rates will be nearly identical for both F-T and methanol routes. But the methanol route will produce more gasoline and less middle distillates. The latter will be lower in cetane and smoke point than that for the F-T route, but still of good marketable quality. The low temperature properties of the middle distillates are better than those from F-T due to highly branched structures. The gasoline will be very olefinic and aromatic - not much different from FCCU gasoline.

The economics of the methanol route improve significantly if only light olefins and gasoline components are desired. Such a methanol route fits well with the gasoline component producing alkylation and TAME units discussed in the Full Refining Cases. But gasoline is in surplus supply at this time.

This study draws no conclusions regarding the relative economics of F-T and a methanol route.

Inherently a methanol plant based only on natural gas has a hydrogen surplus equal to that of the F-T Case discussed previously. Most such modern methanol complexes use such hydrogen as fuel and are essentially in fuel balance with natural gas only going in as reformer feed. However, the feed to the steam methane reformers of the methanol unit is often spiked with CO_2 to convert a portion of the surplus hydrogen as is planned at Novacor's Medicine Hat plant. In such a case, added natural gas is purchased to make up the fuel shortage.

To match the upgrader's hydrogen demands, approximately 4,500 tonnes a day of methanol capacity will be needed. This would be added via 2 methanol trains, each the size of that at Celanese in Edmonton. The study team developed a cost estimate for such a system for the following trial comparison of an overall methanol system as shown in Figure 10.4-1 with the previous Fischer-Tropsch Case.



* IN FULL REFINING CASES

METHANOL TO NAPHTHA
AND MIDDLE DISTILLATE
FIG 10.4-1

The comparison shows little capital difference and operating costs are nearly identical. The product value for the methanol route will be below that of the F-T route (except in refining cases) due to less premium middle distillate and that of lower value.

Table 10.4-1
F-T Versus Methanol to Incremental Liquids
Preliminary Capital Costs in 1Q93 Millions of Canadian Dollars
Incremental to Base Upgrader

	Fischer-Tropsch	Methanol	Note
F-T System (Total from Table B-1)	1,049	—	
Methanol Units	—	750	4,500 tpd
Hydrogen Recovery	—	60	For upgrader H_2
Methanol Conversion	—	150	MTO + MOGD
Related Offsite and Utility Systems	—	80	
Upgrader H_2 Compression Credit	—	-30	H_2 is at 1000 psig
Total	1,049	1,010	
Likely Range	800 to 1,200	800 to 1,200	

11.0 ECONOMIC COMPARISONS

11.1 PREAMBLE

This study is about how to improve upgrader economics, starting from the Basic Upgrader developed for the 1990 Regional Upgrader Business Plan. Thus, the emphasis here is on differentials from the Base Case. The economic analysis is based on the net operating cash flow (before tax, excluding inflation and with no allowance for grants or special financing) over the project life, assuming constant annual production volumes and feedstock/product pricing as per the Purvin and Gertz forecast.

The various cases are compared by:

a) Ratio of gross margin to total capital cost.
b) Net present value (10% discount) of net operating cash flow.
c) Internal rate of return based on net operating cash flow.

The Expanded Base Case was deliberately added to provide a comparison of upgrading per se with the add-on alternates at a constant bitumen rate.

G.S.T. has <u>not</u> been considered in this study as significant export volumes are expected in all cases, resulting in a "zero balance" G.S.T. situation in all cases. <u>If</u> only domestic sales were envisaged, there will be a slight reduction in revenues minus operating cost margins due to profit and non G.S.T. related costs - e.g. internal labour, nor do the costs include any corporate overheads or marketing costs.

Table 11.1-1
Case Annual Revenue and Cost Summary
(In Millions of 1Q93 Canadian Dollars)

Case	No.	Sub Case	Revenues	Non Capital Costs				Gross Margin	Capital Cost	Margin/ Capital (%)	NPV* @ 10%	Internal Rate of Return* (%)
				Feedstock	Variable	Semi Variable	Total					
Base	101	---	626.6	304.6	50.4	73.5	428.5	198.1	1680.7	11.79	251.4	11.67
Expanded Base	201	---	940.0	456.9	75.6	99.7	632.2	307.8	2365.8	13.01	571.6	12.65
F-T	301	---	812.4	304.6	152.4	110.0	567.0	245.4	2730.3	8.99	-406.8	8.14
Partial Refining	401	Fract Only	637.9	304.6	51.5	79.1	435.2	202.6	1810.0	11.20	163.8	11.03
Partial Refining	402	Cond Only	743.0	413.7	52.6	77.5	542.8	200.2	1775.8	11.27	195.8	11.24
Partial Refining	403	Fract + Cond	754.2	413.7	52.6	81.5	547.8	206.4	1880.9	10.97	138.8	10.84
Integrated	501	Fract + Cond + F-T	939.9	413.7	153.2	118.3	685.3	254.6	2928.7	8.69	-507.2	7.82
Full Refining	601	Basic Refy	731.5	314.9	62.9	101.7	479.5	251.9	2275.2	11.07	-8.2	10.46
Full Refining	602	Refy + F-T	897.1	308.3	161.3	137.0	606.6	290.5	3311.2	8.77	-714.5	7.20
Full Refining	603	Refy + Cond	853.6	420.7	64.6	104.3	589.6	263.9	2342.6	11.27	4.4	10.02
Full Refining	604	Refy + Cond + F-T	1026.4	417.4	162.7	138.9	719.0	307.5	3363.5	9.14	-651.7	7.50

Notes:

* Based on net operating revenue before tax over project life.

Table 11.1-2
Annual Differentials Versus Base Case* Summary
(in Millions of 1Q93 Canadian Dollars)

Case	No.	Sub Case	Revenues	Non Capital Costs				Gross Margin	Capital Cost	Margin/ Capital (%)	NPV** @ 10%	Internal Rate of Return** (%)
				Feedstock	Variable	Semi Variable	Total					
Base Case*	101	----	626.6	304.6	50.4	73.5	428.5	198.1	1680.7	11.79	251.4	11.67
Expanded Base	201	----	313.3	152.3	25.2	26.2	203.7	109.6	685.2	16.0	320.2	14.91
F-T	301	----	185.7	0	101.9	36.6	138.5	47.2	1049.6	4.5	-658.1	-2.0
Partial Refining	401	Fract Only	11.2	0	1.1	5.6	6.7	4.5	129.3	3.5	-87.6	-3.52
Partial Refining	402	Cond Only	116.4	109.1	1.1	4.0	114.3	2.1	95.1	2.2	-55.6	0.2
Partial Refining	403	Fract + Cond	127.6	109.1	2.1	8.1	119.3	8.2	200.2	4.1	-112.6	0.7
Integrated	501	Fract + Cond + F-T	313.3	109.1	102.8	44.9	256.8	56.5	1248.0	4.5	-758.6	3.5
Full Refining	601	Basic Refy	104.8	10.3	12.4	28.3	51.0	53.8	594.6	9.1	-259.6	3.3
Full Refining	602	Refy + F-T	270.4	3.7	110.8	63.5	178.1	92.4	1630.6	5.6	-965.9	2.1
Full Refining	603	Refy + Cond	226.9	116.1	14.2	30.9	161.1	65.8	661.9	9.9	-247.0	1.9
Full Refining	604	Refy + Cond + F-T	399.8	112.8	112.3	65.4	290.5	109.3	1682.8	6.5	-903.1	1.8

Notes:

* Base Case actual values shown for reference.
** Based on net operating revenue before tax over project life.

Table 11.1-1 presents a summary of the revenues, non capital charges, and capital costs for the principal alternates, with Table 11.1-2 presenting differentials off the Base Case being of more interest and importance. Appendix B presents the breakdowns of capital and operating costs from which these data were developed.

11.2 EXPANDED BASE CASE DISCUSSION

As expected, the Expanded Base Case presents a more favourable financial picture than does the Base Case per se. But of more importance here is its 12.7% R.O.I. (and $572 million NPV at 10%). The incremental costs between the Expanded Case and the Base Case shows 14.9% R.O.I. (an improvement of $320 million with NPV at 10%).

However, the length of construction has been assumed to be the same as that of the Base Case which is probably incorrect and an alternate case was tested with construction taking a year longer, dropping NPV to $466 million at 10% and R.O.I. to 12.2%.

11.3 FISCHER-TROPSCH CASE

For the addition of Fischer-Tropsch conversion to the Base Case, the basic R.O.I. is only 8.14% and NPV of $-407 million (at 10% discount) as shown in Table 11.1-1. Table 11.3-1 presents a variety of sensitivities on these incremental costs to test the impacts of capital differences and unit revenues.

The estimated premium value for the F-T middle distillate may not materialize, hence, the sensitivity from the reference sub case where the premium calculated in Section 3.4 above has been used.

Table 11.3-1
Fischer-Tropsch Sensitivities
Including Base Case
(in Millions of 1Q93 Canadian Dollars)

Sub Case	Operating Margin	Capital	OM/Capital (%)	NPV	Internal Rate of Return (%)
F-T Reference	245.4	2730.3	8.99	-380.3	8.27
F-T Low Product Price (a)	230.6	2730.3	8.45	-477.5	7.81
F-T Capital -20% (b)	262.7	2184.2	12.03	173.4	10.93
Ft Operating (c) -20%	297.0	2730.3	10.88	43.0	10.19
F-T Capital & Operating (b) (c) - 20%	304.9	2184.2	13.96	574.0	12.93

Notes: (a) Middle distillate fraction at diesel price.
(b) Capital costs reduced by 20% for the whole complex including Upgrader portion.
(c) Operating costs (including natural gas feedstock to the F-T process) are reduced by 20%. Bitumen feedstock cost to the upgrader is not reduced by 20%.

Table 11.4-1
Partial Refining Sub Case Sensitivities
Including Base Case
(in Millions of 1Q93 Canadian Dollars)

Sub Case	Operating Margin	Capital	OM/Capital (%)	NPV @ 10%	Internal Rate of Return (%)
Partial Refining - S.C.O. Fract Only Reference (a)	202.6	1810.0	11.20	163.8	11.03
Partial Refining - With Condensate Added	200.2	1775.8	11.27	195.8	11.24
PR/S.C.O. Fract - S.C.O. + $1/bbl (a)	238.8	1810.0	12.36	329.0	12.04
PR/S.C.O. + Condensate - S.C.O. + $1.00/bbl	227.9	1775.8	12.83	404.2	12.52
PR/S.C.O. Fract - High Middle Distillate Sales (b)	208.3	1810.0	11.51	221.9	11.39
PR/S.C.O. Fract + Condensate Addition + $1/bbl for S.C.O.	231.9	1880.9	12.33	332.8	11.99

Notes: (a) 6,000 BPCD of middle distillate direct to sales.
(b) 9,000 BPCD of middle distillate direct to sales.

F-T capital and operating (excluding natural gas) costs were developed on less reliable bases than other costs in this study, hence, the capital and operating cost sensitivities.

The return on the addition of Fischer-Tropsch synthesis is not economically attractive, despite the premium some of the liquid products will receive.

It appears very likely that the use of partial oxidation can reduce costs with today's economics. As noted above, an F-T synthesis facility with partial oxidation can be located anywhere - it has not particular synergy with the upgrader.

Reduced synthesis gas costs and reduced F-T synthesis costs must be targeted in future work. The F-T product finishing section is also an area offering some cost saving potential.

11.4 PARTIAL REFINING CASES

In the Partial Refining Cases, a full S.C.O. capability fractionator is costed but assumed to operate only 20% of the time to produce 6,000 BPCD of Jet A and diesel (in a 1/2 ratio) direct to market. The fractionator also allows production of all S.C.O. as differentiated S.C.O.'s with fractional compositions to meet customer refinery needs. Addition of 12,000 BPCD of diluent via a new hydrotreater is also considered to provide an average S.C.O. nearer the fractional composition of conventional light crude oils.

The capital costs of the S.C.O. fractionator and the naphtha are considered reliable. But this case is based on achieving a higher return on product sales and the sensitivities will be primarily market related.

Additional naphtha in the total S.C.O. will obtain the premium of Regional Upgrader S.C.O. over light sweet crudes and, thus, the condensate addition itself is investigated in one sensitivity. The added naphtha should also take the S.C.O. into the "basic crude" category at many refineries rather than one to be considered to fill out the overall crude slate.

The underlying sub studies leading to this report indicated that at least 20% of the S.C.O.'s middle distillates could be produced as product, but perhaps as much as 30% is possible without impacting S.C.O. value.

The addition of the S.C.O. fractionator and the resulting ability to differentiate S.C.O.'s for each customer should improve netbacks, if Suncor experience is any criteria. However, this ability may only offset an otherwise below predicted value for the S.C.O. Hence, the test at S.C.O. plus $1/bbl should be considered only related to the reference case and not as a stand-alone case.

The data of the table indicate "fair" returns on the partial refining alternates. However, the study team believes that such operations should be integral with any new upgrader in order to achieve full market value for the products.[7]

11.5 INTEGRATED CASE

As the F-T Case has poor results, the addition of S.C.O. fractionation and naphtha hydrotreating does not add enough to make an integrated scheme viable at this stage of F-T system development.

---

[7] The 1990 study considered an H-Oil option with 3 hydrotreaters feeding into an S.C.O. blend. Obviously such an approach would provide much of the benefits of the S.C.O. fractionation capability without capital addition.

11.6 FULL REFINING CASE

As the F-T addition appears to have little merit, no further analyses were done beyond the summary in Table 11.2-1 for F-T related cases.

The addition of diluent (naphtha) to the S.C.O. has a major affect on refining economics but via increased gasolines sales and not desired middle distillate sales. However, the addition of diluent does impact on the refinery's gasoline to diesel ratio, with a lower limit of about 0.78 versus 0.60 for a case without such addition.

The addition of diluent at the 12,000 BPCD level also has an impact on the minimum practical summer RVP - 7.4 versus 6.5 for the basic refinery. And the composition assumed here for diluent may have been too heavy - i.e. understated the C_5C_6 content. If so, the RVP would be even lighter and some C_5C_6 would have to be shipped to ethylene production - in the summer, at least in F-T Cases.

The addition the TAME complex reduces concern about gasoline olefins in all cases and adds a touch of oxygen - a public relations plus. The overall gasoline is expected to meet U.S. mid continent reformulated gasoline specifications with the addition of MTBE to provide the added oxygen. Such MTBE will be at low cost and less on a delivered in gasoline basis via the IPPL system than alternate mid west/Ontario pure MTBE alternates.

However, the reference economic evaluation puts the value of gasoline at a Canadian regular quality level. The bulk of purchases will likely be independent marketers with a predominance of regular gasoline sales. There appears to be little increase in gasoline demands (before oxygenate additions) in the U.S. and possibly Ontario (with the announced reductions in gasoline exports). However, western Canada could well be short of gasoline production capacity by 2000 but a shortfall of 30,000 BPCD "feels" unlikely. Thus, valuation of the gasoline at a regular price appears justified.

However, a sensitivity is run for the base and condensate plus cases showing the impact of an average gasoline netback $2/bbl over Canadian quality regular.

As gasoline and middle distillate prices are quite comparable, only the low gasoline production cases were considered.

As middle distillate markets continue to develop, no major challenge in selling 25% as Jet A and 75% as diesel is foreseen, and thus, the middle distillate prices were not tested.

Table 11.6-1
Full Refining Case Sensitivities
Low Gasoline Production

Sub Case	Operating Margin	Capital	OM/Capital (%)	NPV @ 10%	Internal Rate of Return (%)
Base Refining Reference	251.9	2275.2	11.07	-8.2	10.46
Condensate Added Reference	263.9	2342.6	11.27	4.4	10.02
Base + $2/bbl for Gasoline	269.0	2275.2	11.82	130.3	11.20
Condensate + $2/bbl for Gasoline	287.3	2342.6	12.26	184.0	10.92

12.0 RESEARCH AND DEVELOPMENT FACTORS

This study has not considered the Basic Upgrader other than as a foundation on which to build further facilities, hopefully to improve economics. As sound economic bases were paramount to this study, research and development factors are evident only in cases with poor economics.

The poor F-T economics point to much potential for:

a) Reduced cost synthesis gas production,
b) Reduced cost synthesis, and
c) New product finishing routes.

These steps are very interrelated, a point all researchers must keep in mind.

Reduced cost synthesis gas systems will usually be applicable to hydrogen production as well as for synthesis gas for F-T, methanol and similar processes. The recent literature references to "breakthroughs" in going from methane to higher hydrocarbons are primarily relative to processes some time away, at best. Partial oxidation for hydrogen is generally considered as not economically competitive with SMR/PSA today, due largely to oxygen needs.

The F-T synthesis system is expensive and better, more active catalysts are needed to reduce gas circulation rates as well as to reduce catalyst costs. Air Products continues to move its slurry bed process originally planned for methanol towards F-T type products and is now at an intermediate DME - dimethylether - stage. (DME is an intermediate in Mobil's technology used in New England to convert methanol to gasoline.) Air Products are working with a low H_2/CO ratio synthesis gas but is appears probable the process can be adapted in time of a high H_2/CO gas.

F-T synthesis has seen very appreciable research worldwide for many years. But as Shell and South Africans are showing, major niches are finally being found. There remains very major interest in converting remote gas reserves to middle distillates. This study may have erred in not adapting a catalyst system producing a heavier product that would results ultimately in more middle distillate.

The Lyondell C_5 olefin isomerization process points to a strong possibility that isomerization can be integrated directly into F-T synthesis to produce branched isomers, rather than waxy straight chain materials. Such a switch would greatly open up the remote markets noted above.

Shell have developed a major isomerization component in their proprietary F-T was hydrocracking catalyst, but it is not clear if dewaxing can be avoided in cold climates. It would appear preferable to have as much isomerization as possible done before the hydrocracking step.

In the Basic Upgrader there appears to be significant room to adjust/revise secondary hydrocracking/hydroprocessing operations to achieve more middle distillate aromatic saturation and more control of gas oil hydrocracking.

13.0 ENVIRONMENTAL FACTORS

13.1 PREAMBLE

Generally the entire complex is not seen as a major environmental concern as all standard air (SO_2, NO_x), water and land criteria will be met using available technologies.

The use of hydrofluoric acid as the alkylation catalyst may be controversial but is considered by the study team as a better alternate than many trucks a day of sulphuric acid and spent sulphuric acid from/to a regeneration site. However, this must be pursued with Alberta's disaster control agency.

13.2 AIR

The use of low NO_x burners will keep NO_x emissions to a level consistent with provincial national objectives for the region of the upgrader. Such burners are available even for high temperature furnaces such as the steam methane reformers. No need for flue gas NO_x reduction has been seen in this study.

N_2O emissions are becoming of concern due to nitrous oxide's greenhouse gas contribution, but are not anticipated to be above 1 ppm in any emissions from any version of the upgrader complex except possibly from the FCCU regenerator stack and sulphur recovery incinerators, especially where any ammonia has entered the system as from sour water strippers, which does require further study.

Volatile organics will be controlled at source, especially at potential fugitive emission points. Special procedures will be followed to monitor all potential sources during operations.

SO_2 emissions will be almost entirely from the sulphur recovery units. However, in the refining cases there will be traces of SO_2 in the FCCU flue gas. SO_x transfer catalyst additives are available to reduce the quantity in the FCCU flue gas if essential.

Carbon dioxide emissions come from the hydrogen units, boilers and smaller sources in the case of the upgrader, and from process heaters, the FCCU regenerator and steam methane reformers in the add-on units.

Catalytic reforming and C_4 and C_5C_6 isomerization processes use small amounts of chlorides to be scrubbed out of gases going to fuel in at least the last two cases. Due to the anticipated low catalytic reforming severity HCl emissions there should be within provincial guidelines, but this will need confirmation. Extremely tight HF control will be used to ensure compliance with Alberta standards.

13.3 WATER

The F-T system description notes a large additional cooling tower but the refinery description notes air condensing of steam turbine exhaust. The latter is being practised more than in the past and is foreseen in detailed design for large turbines. All other cooling duties in the additional facilities have been assumed as air cooled. There are constraints in withdrawal of water from the North Saskatchewan River and a maximum air cooling will be a must. But that leaves boiler water make-up largely for hydrogen and synthesis gas production and for hydrotreater wash waters.

In the F-T Case, all water used for synthesis gas going to F-T synthesis is recovered as a waste stream - 580 USGPM or so, during detailed design it is expected that such waste water will be found to be treatable for recycle to SMR or POX units.

Sour water stripper bottoms can be reused for hydrotreater condenser wash water with treatment for boiler feedwater to be considered in later design.

The refining operation has several small caustic waste streams that may require evaporation.

13.4 LAND

The off-site disposal of upgrader pitch, as proposed by OSLO, was apparently acceptable to provincial agencies. FCCU spent catalyst at 700 to 1,000 tpy will also go to landfills, assuming that it metal content is as low as now anticipated due to lack of metals in S.C.O.

All of the other spent catalysts will be sent to off-site reclamation/disposal facilities.

13.5 NEIGHBOURS

Sounds and smells can all be readily controlled during the design phase.

13.6 APPROVALS

No major challenges are seen from an environmental viewpoint - water withdrawal appears the major challenge.

14.0 CONCLUSIONS AND RECOMMENDATIONS

14.1 CONCLUSIONS

a) F-T is not a viable add-on to an Alberta upgrader at this time.

b) Differentiated S.C.O.'s, preferably with some added naphtha available and middle distillate products offer means of improving upgrader economics at the $1 to $2/ bbl of overall product level.

c) If significant added gasoline producing capacity is needed in Western Canada and/or for U.S. northern tier markets, adding full refining to a 60,000 BPD Alberta upgrader is a viable scheme. But such a need must be identified.

d) Significant research and development opportunities continue to present themselves in the following areas:

   i) Synthesis gas and hydrogen generation, and
   ii) F-T and similar synthesis of middle distillate products from natural gas.

14.2 RECOMMENDATIONS

The search for options to improve Alberta heavy crude/bitumen utilization must continue, following potential markets; new technology development - in upgrading, refining, petrochemicals and other industries; and in business/process configuration analysis.

APPENDIX A

PRICE FORECASTS

PURVIN & GERTZ, INC.
CONSULTING ENGINEERS
1720 SUNLIFE PLAZA
144-4TH AVENUE S.W.
CALGARY, ALBERTA T2P 3N4

THOMAS H. WISE
SENIOR PRINCIPAL

TELEPHONE 403/266-7086
FACSIMILE 403/264-2556

February 24, 1993

Mr. Bert Lang
Chairman, Oil Sands Task Force
Alberta Chamber of Resources
Suite 1410, 10235 - 101 Street
Edmonton, Alberta
T5J 3G1

Dear Bert:

Re: ACR File No. 63-105-01-01

We were requested by T. J. McCann to revise our October 23, 1992 price forecast for 1993 constant dollars and also to provide a forecast for gasoline, propane, iso-butane and normal butane (field butanes are available). For reformulated gasoline, add approximately $1.50 (1993 US)/B to regular gasoline for Phase 1 (1995-99) and $4.50/B for Phase 2 (after 2000). The revised and supplemental forecast is attached in Table 2.

If you have any questions, please call.

Yours very truly,

PURVIN & GERTZ, INC.

Tom

Thomas H. Wise, P. Eng.

c.c. T. J. McCann, SICL

Encl.
THW/ab
C-1838

TABLE 2

ALBERTA CHAMBER OF RESOURCES
REAL PRICE FORECAST(1) FOR PETROLEUM AT EDMONTON

(Forecast in 1993 U.S. Dollars per Barrel, Unless Noted)

	Historical				Forecast											
	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004
World Crude Price																
Dubai FOB Mideast	15.86	20.47	18.50	17.17	17.00	17.48	18.01	18.01	18.01	18.01	18.01	18.01	18.01	18.01	18.01	18.01
Prices at Edmonton																
Mixed Sweet Blend	18.40	23.50	20.19	19.09	19.93	20.94	21.75	21.83	21.87	21.94	22.02	22.12	22.19	22.26	22.34	22.37
Synthetic Crude - ACR Quality					21.23	22.38	23.21	23.28	23.33	23.41	23.49	23.59	23.66	23.74	23.83	23.86
Bitumen	11.36	13.42	7.36	n/a	11.13	11.01	12.11	12.12	12.17	12.23	12.32	12.41	12.60	12.58	12.66	12.71
Gasoline, Regular Unleaded					26.34	26.61	27.24	27.31	27.34	27.42	27.48	27.54	27.64	27.53	27.53	27.53
Jet Fuel A					25.57	26.83	27.41	27.46	27.50	27.56	27.69	27.64	27.63	27.61	27.60	27.59
Diesel Fuel, .05% S					25.21	26.24	26.91	26.96	26.99	27.03	27.06	27.18	27.08	27.06	27.04	27.02
Propane	7.44	12.71	10.33	8.14	11.78	12.26	13.66	14.42	14.50	14.57	14.64	14.71	14.77	14.82	14.88	14.94
i-Butane	11.49	16.96	15.88	16.26	15.51	16.62	18.09	18.18	19.28	19.38	19.47	19.57	19.61	19.65	19.69	19.74
Field Butanes	8.18	13.62	13.04	12.18	12.28	13.09	14.81	15.70	16.80	16.90	18.00	18.09	18.14	18.18	18.22	18.26
Natural Gas ($/MMBtu)	1.53	1.30	1.14	1.13	1.24	1.35	1.64	1.68	1.78	1.80	1.84	1.88	1.90	1.93	1.96	1.97

	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018	2019	2020
World Crude Price																
Dubai FOB Mideast	18.55	19.10	19.64	20.19	20.74	21.28	21.83	22.37	22.92	23.46	24.01	24.01	24.01	24.01	24.01	24.01
Prices at Edmonton																
Mixed Sweet Blend	23.01	23.64	24.28	24.86	25.45	26.06	26.66	27.26	27.86	28.45	29.05	29.05	29.04	29.04	29.03	29.02
Synthetic Crude - ACR Quality	24.51	25.17	25.78	26.42	27.04	27.67	28.29	28.91	29.53	30.16	30.77	30.76	30.76	30.76	30.75	30.76
Bitumen	13.25	13.79	14.30	14.82	15.33	15.86	16.06	16.82	15.82	15.80	15.79	15.79	15.79	15.80	15.79	15.79
Gasoline, Regular Unleaded	28.17	28.82	29.46	30.10	30.74	31.39	32.03	32.68	33.32	33.97	34.61	34.61	34.61	34.61	34.61	34.61
Jet Fuel A	28.92	28.60	29.46	30.12	30.76	31.40	32.03	32.67	33.31	33.96	34.59	34.69	34.68	34.64	34.68	34.57
Diesel Fuel, .05% S	27.66	28.30	28.96	29.59	30.24	30.88	31.54	32.19	32.84	33.49	34.15	34.14	34.13	34.13	34.12	34.12
Propane	15.44	15.91	16.37	16.84	17.30	17.77	18.23	18.70	19.17	19.65	20.13	20.15	20.18	20.21	20.23	20.26
i-Butane	20.03	20.93	21.61	22.11	22.70	23.29	23.88	24.48	25.07	25.66	26.27	26.20	26.25	26.25	26.24	26.24
Field Butanes	16.80	17.35	17.88	18.42	18.96	19.60	20.02	20.55	21.09	21.63	22.20	22.20	22.25	22.24	22.24	22.23
Natural Gas ($/MMBtu)	2.08	2.18	2.29	2.40	2.51	2.61	2.72	2.83	2.93	3.04	3.15	3.16	3.17	3.19	3.20	3.21

February, 1993 (1)

Note: (1) Forecast is restated from October, 1992 forecast in 1993 dollars.

PRICING FORECAST IN CONSTANT 1Q93 CANADIAN DOLLARS

Stream	Mogas (Reg U/L)	Jet A-1	Diesel (.05% S)	Field Butanes	Propane	Sulphur	Natural Gas	Byproduct H2	Methanol
Year									
1993	31.68	31.96	31.51	15.35	14.70	50.00	1.55	1.56	21.00
1994	33.14	33.29	32.80	16.36	15.33	50.00	1.69	NA	21.25
1995	34.05	34.26	33.64	18.26	16.94	50.00	1.93	NA	21.70
1996	34.14	34.33	33.70	19.62	18.03	50.00	2.10	NA	22.03
1997	34.23	34.38	33.74	19.75	18.13	50.00	2.23	NA	22.27
1998	34.28	34.44	33.79	19.88	18.21	50.00	2.25	NA	22.31
1999	34.35	34.49	33.83	20.00	18.30	50.00	2.30	NA	22.41
2000	34.43	34.55	33.88	20.11	18.39	50.00	2.35	1.80	22.50
2001	34.41	34.54	33.85	20.18	18.46	50.00	2.37	NA	22.55
2002	34.41	34.51	33.83	20.23	18.53	50.00	2.41	NA	22.62
2003	34.41	34.38	33.80	20.28	18.60	50.00	2.44	NA	22.66
2004	34.41	34.49	33.78	20.33	18.68	50.00	2.46	NA	22.71
2005	35.21	35.28	34.58	21.00	19.30	50.00	2.60	NA	22.97
2006	36.03	36.08	35.38	21.69	19.89	50.00	2.73	NA	23.20
2007	36.83	36.85	36.19	22.35	20.46	50.00	2.86	NA	23.46
2008	37.63	37.65	36.99	23.03	21.05	50.00	3.00	NA	23.72
2009	38.43	38.45	37.80	23.70	21.63	50.00	3.14	NA	23.98
2010	39.24	39.25	38.60	24.38	22.21	50.00	3.26	NA	24.21
2011	40.04	40.04	39.43	25.03	22.79	50.00	3.40	NA	24.47
2012	40.81	40.84	40.24	25.69	23.37	50.00	3.54	NA	24.73
2013	41.65	41.64	41.05	26.36	23.96	50.00	3.66	NA	24.96
2014	42.46	42.44	41.86	27.06	24.56	50.00	3.80	NA	25.22
2015	43.26	43.24	42.69	27.83	25.16	50.00	3.94	NA	25.48
2016	43.26	43.24	42.68	27.81	25.20	50.00	3.95	NA	25.50
2017	43.26	43.23	42.66	27.81	25.23	50.00	3.96	NA	25.52
2018	43.26	43.23	42.66	27.80	25.26	50.00	3.99	NA	25.57
2019	43.26	43.23	42.65	27.80	25.29	50.00	4.00	NA	25.59
2020	43.26	43.21	42.65	27.79	25.33	50.00	4.01	NA	25.62
2021	43.26	43.21	42.65	27.79	25.33	50.00	4.01	NA	25.62
2022	43.26	43.21	42.65	27.79	25.33	50.00	4.01	NA	25.62
2023	43.26	43.21	42.65	27.79	25.33	50.00	4.01	NA	25.62
2024	43.26	43.21	42.65	27.79	25.33	50.00	4.01	NA	25.62
2025	43.26	43.21	42.65	27.79	25.33	50.00	4.01	NA	25.62
2026	43.26	43.21	42.65	27.79	25.33	50.00	4.01	NA	25.62
2027	43.26	43.21	42.65	27.79	25.33	50.00	4.01	NA	25.62

PRICING FORECAST IN CONSTANT 1Q93 CANADIAN DOLLARS

Stream Year	Alberta Light Sweet Crude	Raw Bitumen	Diluted Bitumen	Diluent	Intermediates	Regional Up S.C.O.	Regional Up Components	FT Mid Dist	FT Naphtha
1993	24.91	13.91	17.58	24.91	18.08	26.54	26.54	35.70	24.91
1994	26.18	14.39	18.32	26.18	18.82	27.95	27.95	36.99	26.18
1995	27.19	15.14	19.15	27.19	19.65	29.01	29.01	37.83	27.19
1996	27.29	15.15	19.20	27.29	19.70	29.08	29.08	37.89	27.29
1997	27.34	15.21	19.25	27.34	19.75	29.16	29.16	37.93	27.34
1998	27.43	15.29	19.33	27.43	19.83	29.26	29.26	37.98	27.43
1999	27.53	15.40	19.44	27.53	19.94	29.36	29.36	38.02	27.53
2000	27.65	15.51	19.56	27.65	20.06	29.49	29.49	38.07	27.65
2001	27.74	15.63	19.66	27.74	20.16	29.58	29.58	38.04	27.74
2002	27.83	15.73	19.76	27.83	20.26	29.68	29.68	38.02	27.83
2003	27.93	15.83	19.86	27.93	20.36	29.79	29.79	37.99	27.93
2004	27.96	15.89	19.91	27.96	20.41	29.83	29.83	37.97	27.96
2005	28.76	16.56	20.63	28.76	21.13	30.64	30.64	38.77	28.76
2006	29.55	17.24	21.34	29.55	21.84	31.46	31.46	39.57	29.55
2007	30.31	17.88	22.02	30.31	22.52	32.24	32.24	40.38	30.31
2008	31.06	18.53	22.70	31.06	23.20	33.03	33.03	41.18	31.06
2009	31.81	19.16	23.38	31.81	23.88	33.80	33.80	41.99	31.81
2010	32.58	19.81	24.07	32.58	24.57	34.59	34.59	42.79	32.58
2011	33.33	19.94	24.40	33.33	24.90	35.36	35.36	43.62	33.33
2012	34.08	19.78	24.54	34.08	25.04	36.14	36.14	44.43	34.08
2013	34.83	19.78	24.79	34.83	25.29	36.91	36.91	45.24	34.83
2014	35.56	19.75	25.02	35.56	25.52	37.69	37.69	46.05	35.56
2015	36.31	19.74	25.26	36.31	25.76	38.46	38.46	46.88	36.31
2016	36.31	19.74	25.26	36.31	25.76	38.45	38.45	46.87	36.31
2017	36.30	19.74	25.26	36.30	25.76	38.45	38.45	46.85	36.30
2018	36.30	19.75	25.27	36.30	25.77	38.45	38.45	46.85	36.30
2019	36.29	19.74	25.25	36.29	25.75	38.44	38.44	46.84	36.29
2020	36.28	19.74	25.25	36.28	25.75	38.45	38.45	46.84	36.28
2021	36.28	19.74	25.25	36.28	25.75	38.45	38.45	46.84	36.28
2022	36.28	19.74	25.25	36.28	25.75	38.45	38.45	46.84	36.28
2023	36.28	19.74	25.25	36.28	25.75	38.45	38.45	46.84	36.2[illegible]
2024	36.28	19.74	25.25	36.28	25.75	38.45	38.45	46.84	36.28
2025	36.28	19.74	25.25	36.28	25.75	38.45	38.45	46.84	36.28
2026	36.28	19.74	25.25	36.28	25.75	38.45	38.45	46.84	36.28
2027	36.28	19.74	25.25	36.28	25.75	38.45	38.45	46.84	36.28

APPENDIX B

CAPITAL AND OPERATING COSTS

Appendix B-1
Capital Costs
Thousands of 1Q93 Dollars

CASE NUMBER	101	201	301	401	402	405	501	601	602	603	604
DESCRIPTION	BASE	EXP BASE	BASE	PARTIAL REFINING			INTEG.	FULL REFINING			
			+F-T	SCO +	COND	SCO +	SCO+	BASE	+ F-T	+ COND	+COND
				FRACT	ONLY	FRACT+	FRACT+				& F-T
						COND	COND+F-T				
Primary / Secondary Upgrading Trains	337,333	505,999	337,333	337,333	337,333	337,333	337,333	337,333	337,333	337,333	337,333
Atmospheric Distillation	29,990	38,987	29,990	29,990	29,990	29,990	29,990	29,990	29,990	29,990	29,990
Vacuum Distillation	13,388	17,404	13,388	13,388	13,388	13,388	13,388	13,388	13,388	13,388	13,388
Sour Water Stripper	3,194	4,791	3,194	3,194	4,194	4,194	4,194	6,194	6,694	6,694	7,194
Sulphur Plant & Tail Gas Unit	9,260	13,890	9,260	9,260	9,260	9,260	9,260	9,260	9,760	9,760	9,760
Hydrogen Plant	116,207	174,311		116,207	116,207	116,207		116,207	0	116,207	0
Waste Water Treatment	7,706	10,018	10,706	7,706	7,706	7,706	10,706	9,706	9,706	10,206	10,206
Steam Plant	11,815	16,541	16,815	13,815	15,815	15,815	20,815	13,815	13,815	13,815	13,815
Pipelines	10,000	11,000	11,000	12,000	12,000	12,000	12,000	11,000	11,000	11,000	12,000
Tankage	32,902	41,786	53,427	45,357	47,112	47,112	67,637	51,747	53,027	53,027	54,913
Utilities	53,635	69,725	65,435	54,635	55,635	55,635	67,435	53,635	65,635	53,635	65,635
Flare	42,023	58,832	52,023	43,023	44,023	44,023	54,023	49,023	52,023	50,023	53,023
Raw Water Treatment	6,140	8,596	7,140	7,640	8,140	8,140	9,140	6,140	8,140	8,140	8,140
I C Pipeways	60,227	78,295	76,785	65,227	67,227	67,227	83,785	80,227	84,227	82,227	85,227
Buildings	13,984	16,781	14,984	14,984	15,484	15,484	16,484	16,484	16,984	16,484	16,984
Site Improvements	37,538	46,923	42,518	40,538	41,538	41,538	46,518	48,800	52,554	52,554	56,307
S C O Fractionation	0	0		33,988	0	33,988	33,988	29,300	32,000	29,300	32,000
Naphtha Debutanizer	0	0		1,740	1,000	2,740	2,740	1,500	3,500	2,500	4,000
Naphtha Hydrotreater & Splitter	0	0		0	0	21,228	21,228	13,200	18,300	16,700	22,400
C5 C6 Isomerization							0	9,800	19,000	14,200	19,800
Cat Reformer (modified)							0	18,200	34,300	30,800	43,600
Fluid Catalytic Cracking Unit							0	83,900	83,900	83,900	83,900
Gasoline Treating							0	900	900	900	900
C3 C4 Alkylation							0	43,800	43,800	43,800	43,800
C4 Isomerization								15,000	15,000	15,000	15,000
Tertiary Amyl Methyl Ether							0	20,600	20,600	20,600	20,600
Saturate Gas Plant							0	3,000	4,000	4,000	5,000
Light Cycle Oil Hydrotreater							0	12,200	12,200	12,200	12,200
F-T Costs			485,094				485,094		485,094		485,094
F-T Utilities			77,200				77,200		77,200		77,200
TOTAL DIRECT FIELD COSTS	785,342	1,113,879	1,306,292	850,025	826,052	883,008	1,402,958	1,104,349	1,614,070	1,138,383	1,639,409

Appendix B-1
Capital Costs
Thousands of 1Q93 Dollars

CASE NUMBER	101	201	301	401	402	405	501	601	602	603	604
DESCRIPTION	BASE	EXP BASE	BASE	PARTIAL REFINING			INTEG.	FULL REFINING			
			+F-T	SCO +	COND	SCO +	SCO+	BASE	+ F-T	+ COND	+COND
				FRACT	ONLY	FRACT+	FRACT+				& F-T
						COND	COND+F-T				
GENERAL FIELD EXPENSE:											
Direct Hire Support	143,618	203,709	209,135	155,454	151,070	161,486	227,003	177,666	249,924	183,368	254,536
Construction Management	28,723	40,742	41,917	31,091	30,214	32,297	45,401	35,533	49,985	36,674	50,907
Bussing / Travel	23,560	33,416	42,224	25,561	25,751	26,550	45,214	32,752	46,073	33,804	46,923
SUBTOTAL	195,901	277,854	293,276	212,106	207,036	220,334	317,618	245,951	345,982	253,846	352,366
ENGINEERING & PROCUREMENT	133,508	189,359	222,070	144,504	140,429	150,111	238,503	187,739	274,392	193,525	278,700
TOTAL CONSTRUCTED COST	1,114,751	1,581,092	1,821,638	1,206,636	1,173,516	1,253,453	1,959,079	1,538,040	2,234,444	1,585,754	2,270,475
OTHER CAPITAL COSTS:											
Initial Catalyst & Chemicals (estimated)	7,078	10,617	63,778	7,078	10,617	8,578	65,278	15,078	71,778	16,078	72,778
Owner's Budget	111,807	145,349	133,007	115,807	117,707	119,607	140,807	131,807	153,007	133,807	155,007
Startup Budget	55,904	78,265	76,104	58,904	59,654	60,404	80,604	68,904	89,104	69,904	90,104
Capital Spares	8,564	11,990	14,270	9,564	10,064	10,564	16,270	3,500	9,200	3,800	9,500
SUBTOTAL	183,353	246,221	287,159	191,353	198,042	199,153	302,959	219,289	323,089	223,589	327,389
ALLOWANCE FOR OMMISSIONS	129,810	182,731	210,880	139,799	137,156	145,261	226,204	175,733	255,753	180,934	259,786
CONTINGENCY	142,791	201,005	231,968	153,779	150,871	159,787	248,824	193,306	281,329	199,028	285,765
TOTAL ESTIMATED COST	1,570,706	2,211,049	2,551,644	1,691,566	1,659,586	1,757,653	2,737,066	2,126,368	3,094,615	2,189,305	3,143,415
G.S.T.	109,949	154,773	178,615	118,410	116,171	123,036	191,595	148,846	216,623	153,251	220,039
TOTAL INSTALLED PLANT COST	1,680,655	2,365,822	2,730,259	1,809,976	1,775,757	1,880,689	2,928,661	2,275,213	3,311,238	2,342,556	3,363,454
										MARCH 24,1993	

Appendix B-2
Operating Costs
Thousands of 1Q93 Dollars

CASE NUMBER			101	201	301	401	402	405	501	601	602	603	604
DESCRIPTION			BASE	EXP BASE	BASE	PARTIAL REFINING			INTEG.	FULL REFINING			
					+F - T	SCO +	COND	SCO +	SCO+	BASE	+ F - T	+COND	+ COND
						FRACT	ONLY	FRACT+	FRACT+				& F - T
								COND	COND+F-T				
VARIABLE COSTS:													
Natural Gas Req'd (10x9 BTU/CD)			51.68	77.52	241.05	53.20	52.40	53.90	241.05	57.80	241.05	58.50	241.05
Natural Gas @	$1.55	/MMBTU	29,238	43,857	136,374	30,098	29,645	30,494	136,374	32,700	136,374	33,096	136,374
Electric Power (MW)			33.6	50.4	-18.8	34.6	35.0	35.6	-16.8	51.6	-0.8	52.6	0.2
@ .90 load factor			8,273	12,358	-4,674	8,516	8,613	8,759	-4,188	12,649	-247	12,892	65
Catalyst & Chemicals			8,902	13,353	16,630	8,902	9,302	9,302	17,030	13,500	21,130	14,600	22,230
Pitch Disposal			4,022	6,033	4,022	4,022	4,022	4,022	4,022	4,022	4,022	4,022	4,022
	SUBTOTAL		50,435	75,600	152,352	51,538	51,583	52,577	153,238	62,872	161,279	64,611	162,691
SEMI-VARIABLE COSTS:													
Operating Labour (additional people)			0	25	35	10	10	12	47	48	75	50	75
Operating Labour @	$76,000	/Person	0	1,900	2,660	760	760	912	3,572	3,648	5,700	3,800	5,700
OPERATING LABOUR TOTAL COST			12,920	14,820	15,580	13,680	13,680	13,832	16,492	16,568	18,620	16,720	18,620
Maintenance Labour			22,295	31,622	36,433	24,133	23,470	25,069	39,182	30,761	44,689	31,715	45,410
Maintenance Material			22,295	31,622	36,433	24,133	23,470	25,069	39,182	30,761	44,689	31,715	45,410
Misc Operating Supplies			550	700	550	550	550	550	550	800	800	800	800
Administration & Support			4,560	5,472	4,560	4,560	4,560	4,560	4,560	6,900	6,900	6,900	6,900
Office Costs & Misc			550	700	550	550	550	550	550	1,000	1,000	1,000	1,000
Insurance			2,787	3,953	4,554	3,017	2,934	3,134	4,898	3,845	5,586	3,964	5,676
Local Taxes			5,574	7,905	9,108	6,033	5,868	6,267	9,795	7,690	11,172	7,929	11,352
Interest on Working Capital			1,924	2,886	2,281	2,412	2,017	2,505	3,138	3,389	3,572	3,517	3,694
	SUBTOTAL		73,455	99,680	110,049	79,067	77,099	81,536	118,346	101,714	137,028	104,260	138,862
	TOTAL		123,890	175,280	262,400	130,605	128,682	134,113	271,584	164,586	298,307	168,871	301,552
												MARCH 25,1993	

APPENDIX C
CASE COMPARISONS

CASE No. ===>	BASE 101	201	DEL BASE	301	DEL BASE	401	DEL BASE	405	DEL BASE
VOLUMES									
Feedstocks									
Bitumen	60000	90000	30000	60000	0	60000	0	60000	0
Condensate	0	0	0	0	0	0	0	12000	12000
Field Butanes	0	0	0	0	0	0	0	0	0
Methanol	0	0	0	0	0	0	0	0	0
Subtotal	60000	90000	30000	60000	0	60000	0	72000	12000
Products									
Diluent			0		0		0		0
Intermediates			0		0		0		0
S.C.O.	63894	95841	31947	63894	0	57894	-4000	69894	6000
Pitch (Long tons)	290	435	145	290	0	290	0	290	0
FT Middle Distillate			0	9673	9673		0		0
FT Naphtha			0	5767	5767		0		0
Gasoline (Regular U/L)			0		0		0		0
Jet A-1			0		0	2000	2000	2000	2000
Diesel (.05% S)			0		0	4000	4000	4000	4000
Propane			0		0		0		0
n-Butane			0	1295	1295		0		0
Sulphur (Long tons)	422	634	212	422	0	422	0	428	6
Subtotal	64184	96276	32092	80919	16735	64184	0	76184	12000
REVENUES									
Products									
S.C.O.	$618.9	$928.4	$309.5	$618.9	$0.0	$560.8	($58.1)	$677.1	$58.1
FT Middle Distillate	$0.0	$0.0	$0.0	$126.0	$126.0	$0.0	$0.0	$0.0	$0.0
FT Naphtha	$0.0	$0.0	$0.0	$52.4	$52.4	$0.0	$0.0	$0.0	$0.0
Gasoline (Regular U/L)	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Jet A-1	$0.0	$0.0	$0.0	$0.0	$0.0	$23.3	$23.3	$23.3	$23.3
Diesel (.05% S)	$0.0	$0.0	$0.0	$0.0	$0.0	$46.0	$46.0	$46.0	$46.0
Propane	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Sulphur	$7.7	$11.6	$3.9	$7.7	$0.0	$7.7	$0.0	$7.8	$0.1
n-Butane	$0.0	$0.0	$0.0	$7.3	$7.3	$0.0	$0.0	$0.0	$0.0
TOTAL REVENUES	$626.6	$940.0	$313.3	$812.4	$185.7	$637.9	$11.2	$754.2	$127.6
COSTS									
Feedstocks									
Bitumen	$304.6	$456.9	$152.3	$304.6	$0.0	$304.6	$0.0	$304.6	$0.0
Condensate	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$109.1	$109.1
Field Butanes	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Methanol	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Subtotal	$304.6	$456.9	$152.3	$304.6	$0.0	$304.6	$0.0	$413.7	$109.1
Operating									
Variable	$50.4	$75.6	$25.2	$152.4	$101.9	$51.5	$1.1	$52.6	$2.1
Semi-Variable	$73.5	$99.7	$26.2	$110.0	$36.6	$79.1	$5.6	$81.5	$8.1
Subtotal	$123.9	$175.3	$51.4	$262.4	$138.5	$130.6	$6.7	$134.1	$10.2
TOTAL ANNUAL COSTS	$428.5	$632.2	$203.7	$567.0	$138.5	$435.2	$6.7	$547.8	$119.3
GROSS MARGIN	$198.1	$307.8	$109.6	$245.4	$47.2	$202.6	$4.5	$206.4	$8.2
TOTAL CAPITAL COSTS	$1,680.7	$2,365.8	$685.2	$2,730.3	$1,049.6	$1,810.0	$129.3	$1,880.9	$200.2
Ratio GM:TCC	11.79%	13.01%	16.00%	8.99%	4.50%	11.20%	3.48%	10.97%	4.12%
TOTAL CONSTRUCTED COST	$1,114.8	$1,581.1		$1,821.6		$1,206.6		$1,253.5	

CASE No. ===>	501	DEL BASE	601	DEL BASE	602	DEL BASE	603	DEL BASE	604	DEL BASE	408	DEL BASE
VOLUMES												
Feedstocks												
Bitumen	60000	0	60000	0	60000	0	60000	0	60000	0	60000	0
Condensate	12000	12000	0	0	0	0	12000	12000	12000	12000	12000	12000
Field Butanes	0	0	1180	1180	0	0	582	582	0	0	0	0
Methanol	0	0	482	482	482	482	482	482	482	482	0	0
Subtotal	72000	12000	61662	1662	60482	482	73064	13064	72482	12482	72000	12000
Products												
Diluent		0		0		0		0		0		0
Intermediates		0		0		0		0		0		0
S.C.O.	69894	6000		-63894		-63894		-63894		-63894	75894	12000
Pitch (Long tons)	290	0	290	0	290	0	290	0	290	0	290	0
FT Middle Distillate	9673	9673		0		0		0		0		0
FT Naphtha	5767	5767		0		0		0		0		0
Gasoline (Regular U/L)		0	23373	23373	27905	27905	31952	31952	36831	36831		0
Jet A-1	2000	2000	9754	9754	12172	12172	10234	10234	12652	12652	0	0
Diesel (.05% S)	4000	4000	29261	29261	36516	36516	30701	30701	37956	37956	0	0
Propane		0	593	593	683	683	707	707	797	797		0
n-Butane	1295	1295		0	189	189		0	774	774		0
Sulphur (Long tons)	428	6	422	0	422	0	428	6	428	6	428	6
Subtotal	92919	28735	63271	-913	77755	13571	73884	9700	89300	25116	76184	12000
REVENUES												
Products												
S.C.O.	$677.1	$58.1	$0.0	($618.9)	$0.0	($618.9)	$0.0	($618.9)	$0.0	($618.9)	$735.2	$116.2
FT Middle Distillate	$126.0	$126.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
FT Naphtha	$52.4	$52.4	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Gasoline (Regular U/L)	$0.0	$0.0	$270.3	$270.3	$322.7	$322.7	$369.5	$369.5	$425.9	$425.9	$0.0	$0.0
Jet A-1	$23.3	$23.3	$113.8	$113.8	$142.0	$142.0	$119.4	$119.4	$147.6	$147.6	$0.0	$0.0
Diesel (.05% S)	$46.0	$46.0	$336.5	$336.5	$420.0	$420.0	$353.1	$353.1	$436.5	$436.5	$0.0	$0.0
Propane	$0.0	$0.0	$3.2	$3.2	$3.7	$3.7	$3.8	$3.8	$4.3	$4.3	$0.0	$0.0
Sulphur	$7.8	$0.1	$7.7	$0.0	$7.7	$0.0	$7.8	$0.1	$7.8	$0.1	$7.8	$0.1
n-Butane	$7.3	$7.3	$0.0	$0.0	$1.1	$1.1	$0.0	$0.0	$4.3	$4.3	$0.0	$0.0
TOTAL REVENUES	$939.9	$313.3	$731.5	$104.8	$897.1	$270.4	$853.6	$226.9	$1,026.4	$399.8	$743.0	$116.4
COSTS												
Feedstocks												
Bitumen	$304.6	$0.0	$304.6	$0.0	$304.6	$0.0	$304.6	$0.0	$304.6	$0.0	$304.6	$0.0
Condensate	$109.1	$109.1	$0.0	$0.0	$0.0	$0.0	$109.1	$109.1	$109.1	$109.1	$109.1	$109.1
Field Butanes	$0.0	$0.0	$6.6	$6.6	$0.0	$0.0	$3.3	$3.3	$0.0	$0.0	$0.0	$0.0
Methanol	$0.0	$0.0	$3.7	$3.7	$3.7	$3.7	$3.7	$3.7	$3.7	$3.7	$0.0	$0.0
Subtotal	$413.7	$109.1	$314.9	$10.3	$308.3	$3.7	$420.7	$116.1	$417.4	$112.8	$413.7	$109.1
Operating												
Variable	$153.2	$102.8	$62.9	$12.4	$161.3	$110.8	$64.6	$14.2	$162.7	$112.3	$51.6	$1.1
Semi-Variable	$118.3	$44.9	$101.7	$28.3	$137.0	$63.6	$104.3	$30.8	$138.9	$65.4	$77.1	$3.6
Subtotal	$271.6	$147.7	$164.6	$40.7	$298.3	$174.4	$168.9	$45.0	$301.6	$177.7	$128.7	$4.8
TOTAL ANNUAL COSTS	$685.3	$256.8	$479.5	$51.0	$606.6	$178.1	$589.6	$161.0	$719.0	$290.5	$542.4	$113.9
GROSS MARGIN	$254.6	$56.5	$251.9	$53.8	$290.4	$92.3	$264.0	$65.9	$307.5	$109.3	$200.6	$2.5
TOTAL CAPITAL COSTS	$2,928.7	$1,248.0	$2,275.2	$594.6	$3,311.2	$1,630.6	$2,342.6	$661.9	$3,363.5	$1,682.8	$1,775.8	$95.1
Ratio GM:TCC	8.69%	4.53%	11.07%	9.05%	8.77%	5.66%	11.27%	9.95%	9.14%	6.50%	11.30%	2.59%
TOTAL CONSTRUCTED COST	$1,959.1		$1,538.0		$2,234.4		$1,585.8		$2,270.5		$1,173.5	

NOTE ==> Pitch disposal costs have been allocated to the Variable operating cost account.