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FEASIBILITY STUDY OF A LOW SEVERITY HYDROGENATION PROCESS FOR UPGRADING JET AND DIESEL FUELS

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ABSTRACT

Possible mid 1990's specifications limiting the aromatic content of Canadian jet and diesel fuels will severely strain Canadian refiners. Increasing use of synthetic crude oils will worsen this problem. This study shows that a low severity hydrogenation process has possible commercial applications in both the short term (1992) and long term (2005). CANMET has done extensive test work with conventional high severity aromatic hydrogenation catalysts and recently has been developing a low severity supported nickel catalyst route. The study compares these two routes for four feedstocks representing Canadian applications - synthetic crude middle distillates from each of Suncor Inc. and Syncrude Canada Ltd., a light cycle oil and a blend of light cycle oil with conventional middle distillate. Preliminary capital costs of both routes are found to be equal, but the operating costs of the nickel catalyst route are marginally higher. The development of a sulphur tolerant nickel catalyst with regenerability would greatly improve the economics of the low severity process.

INTRODUCTION

Future projections for petroleum refining in Canada indicate a decline in feedstock qualities, an increase in synthetic crude production and greater environmental constraints. In particular, middle distillate refining will be affected by the higher aromaticity and sulphur content of Canadian crudes. High levels of aromatics degrade diesel ignition and jet fuel combustion quality and may also increase engine particulate emissions (Wilson, 1981). Future syncrudes are expected to be somewhat lower than their current levels in both aromatics and sulphur, but anticipated changes to jet and diesel fuel specifications are likely to create problems for existing hydroprocessing operations in Canada. Improved hydrogenation processes, solvent extraction of aromatics or a combination of both may be required for quality improvement of future middle distillate fuels.

- not before 1995

← to the point of no problem

This study assesses the competitive aspects of using a low severity catalytic hydrogenation process for saturation of aromatics in Canadian middle distillates. The process employs supported nickel metal catalysts and is based upon laboratory and pilot work carried out by CANMET/Energy Research Laboratories, Energy, Mines and Resources Canada (Wilson et al., 1985, 1986, 1987, 1988). The study compares the low severity route with "conventional" high pressure hydrogenation using a sulphided Ni-Mo hydrotreating catalyst. The nickel catalyst route was also studied as one approach to offset the high and rising cost of

Ni Co ?

platinum used in other low pressure processes. The work to date has used commercially available catalysts.

The study examines the Canadian setting for middle distillate aromatics hydrogenation and notes relative economics and prospective applications for short term (1992) and long term (2005). It also identifies areas for further research and development.

CURRENT AND FUTURE CONCERNS FOR PRODUCTION OF MIDDLE DISTILLATE FUELS IN CANADA

Canadian refineries which produce Jet A fuel are forced to restrict their intake of synthetic crude oil because of the present specifications limit of 25 vol % of aromatics and/or a smoke point of at least 18. Synthetic crude kerosene fractions typically have greater than 25 vol % aromatics and poor, i.e. 16-18 smoke points (Steere et al., 1981). Synthetic crude diesel fractions have unacceptably low cetane numbers between 30 and 35, well below the specification cetane rating of 40 minimum. Refineries in Western Canada must also blend very low quality light cycle oil (cetane number 25) from catalytic cracking units into diesel fuel at 15-20 vol % of the total. Only two Canadian refineries out of a total of 27 feature specific aromatic hydrogenation units to improve the quality of diesel fuel. Most refineries have to use cetane improvers, which become very expensive beyond a 5 to 6 cetane number increase. As a result, the intake of syncrude is restricted to about 15 vol % of refinery capacity.

Currently, middle distillate production represents about 40% of all Canadian liquid fuels. This proportion is expected to increase to 44% by 1995 and reach 45% by 2005 (Devai, 1987). All projections indicate an increase in syncrude production in Canada which should be less aromatic due to added processing in primary upgrading. Some domestic crude oils, such as those from the Beaufort Sea, are also likely to have high aromatic contents and hence yield distillates low in cetane number. Simultaneously, the light fuel oil market in Eastern Canada is forecast to decline thus decreasing the outlet for light cycle oil. At the same time a rise in Jet A and diesel fuel demand is expected across the country (Canadian National Energy Board, 1986).

Specifications of 0.05 wt % sulphur and a 20 vol % aromatics have been proposed for diesel fuel in the U.S. to meet 1995 engine emission standards (McCullough et al., 1987). In Canada the odds for gradual implementation of new sulphur and aromatic specifications by the year 1996 are high (Yoes and Asim, 1987). Scientific evidence, which correlates aromatic content and other fuel parameters with engine emissions, will be available as early as mid-1989. Generally, it is believed by Canadian petroleum refiners that a 1996 aromatics specification of 25 vol % would be appropriate for diesel fuel. However, for the purpose of the present study the worst possible case (20 vol %) has been chosen.

Table I shows a preliminary estimate of Canadian middle distillate specification changes between 1987 and 1996.

Aromatics Hydrogenation Capabilities of Canadian Refineries

Only two Canadian refineries, Shell Scotford and Petro-Canada Edmonton, have significant capability for middle distillate aromatics hydrogenation. Two other units are now under construction: New Grade Energy Inc., Regina, Saskatchewan for the primary processing of heavy oils, and Syncrude Canada Ltd., Fort McMurray, Alberta for hydrogenation of oil sand derived middle distillates. Anticipated reductions of middle distillate sulphur levels to 0.05 wt % by 1996 will require additional hydrotreating capacity in Canada. However, it appears unlikely that such new facilities will have a significant enough impact on improving diesel fuel cetane number so that the current specification is met.

Imposition of a diesel fuel aromatics specification in the 20 to 25 vol % range would necessitate the installation of hydrogenation units in about 12 Canadian refineries. The introduction of new hydrogenation capacity at one or both existing oil sands bitumen upgrading plants in Alberta could displace up to four of such units in downstream operations. Nevertheless, aromatics hydrogenation in virgin and syncrude distillates, in some hydrocracked distillates (especially where syncrudes are processed) and in catalytic cracked light cycle oils may all be required. To ease Jet A fuel production problems due to syncrude and other future high aromatic crudes, additional kerosene hydrogenation units may also be needed. Thus a significant Canadian market is foreseen for aromatic hydrogenation units if environmental constraints are imposed in Canada. In the short term, to 1995, up to three kerosene units are possible, and in the long term, beyond 1995, another five diesel fuel units are likely with the introduction of a new aromatics specification.

SUMMARY OF CATALYTIC HYDROPROCESSING ALTERNATIVES FOR MIDDLE DISTILLATE UPGRADING

Considering the above analysis one may assume that new aromatic hydrogenation facilities are likely to be needed in Canada by 1996. An alternative processing route which could be used in isolated cases is solvent extraction. Recent work carried out in this program has explored the possibility for combined aromatics hydrogenation and extraction (Wilson et al., 1987). However in all situations involving extraction, disposal of the aromatic extract presents a major problem.

In this study, catalytic hydroprocessing alternatives are conveniently grouped into six classes. They are based on the type of catalyst used, which is the major factor in determining process operating conditions, and are as follows:

Platinum Sulphide Catalyst (Type I)

In this process the feedstock sulphur is converted to H_2S before passing through a bed of platinum sulphide. The latter hydrogenation stage operates at a modest pressure (6.9 MPa). However operating temperatures are higher than needed for platinum and nickel supported catalysts (processes Type II and III). In spite of some hydrocracking the product yield is still about 100 vol %. The ability to handle feeds up to 10,000 ppm of sulphur is claimed.

Supported Platinum Metal Catalysts (Type II)

Commercially available platinum catalysts require feedstock sulphur levels below 15 ppm (Bartholomew et al., 1982). It is anticipated that a newly developed catalyst tolerating up to 150 ppm sulphur could enter the market within two years. Current processes operate at mild hydrogen pressures (5.0 MPa) and below 200°C (Cobb and Chapel, 1985). Shell's syncrude middle distillate hydrogenation unit at Scotford, Alberta falls into this category. Virtually all platinum catalysts are regenerable and sulphur and nitrogen are temporary poisons only.

Supported Nickel Metal Catalysts (Type III)

Commercial nickel catalysts are currently very competitive with platinum for hydrogenation of feedstocks containing less than 2-3 ppm sulphur. Nickel catalysts are not regenerable and sulphur accumulates as a poison (Bartholomew et al., 1982). Guard beds of special disposable nickel "adsorbents" placed in front of high activity nickel (or platinum) catalysts can considerably extend, at a cost, their life cycle. Operating conditions for nickel catalysts are very similar to those for platinum. Nickel catalysts have also been used in a slurry mode for benzene hydrogenation. In one commercially available catalyst, added copper and cobalt slightly improved the sulphur tolerance and reduced the hydrocracking tendencies of pure nickel (Corbett, 1987).

Newly Developed Catalysts (Type IV)

This category covers hydrogenation catalysts adapted for petroleum refining after they proved successful in other industries. Capabilities of these proprietary, often trimetallic, catalysts incorporate high activities for desulphurization (cobalt) and the hydrogenation capabilities of other metals usually in association with molybdenum.

Conventional Sulphided Ni-Mo and Ni-W Catalysts (Type V)

These typical hydrotreating catalysts are used for aromatic saturation and operate at moderate temperatures (340-380°C) with both minimal hydrocracking and thermodynamic equilibrium effects on aromatics conversion (Wilson et al., 1985; Wilson and Kriz, 1984, 1985). The required hydrogen pressures range from 10 to 20 MPa. In spite of significant sulphur and nitrogen tolerance, a dual catalyst or high temperature sulphided Co-Mo hydrotreating reactor unit may be required for high sulphur containing feedstocks. This type of process has been used extensively outside Canada and the U.S. for kerosene hydrogenation and may be regarded as the current conventional process. All catalyst types in this group are regenerable and have the lowest cost over all other hydroprocessing catalyst types.

Mild Middle Distillate Hydrocracking Catalysts (Type VI)

This group includes Type V catalysts, or specifically formulated sulphided metal catalysts with added fluoride and other promoters on a silica-alumina support. For middle distillate upgrading, the first stage catalyst in a conventional gas oil hydrocracking unit is used almost exclusively. The process operates at high temperatures (400°C +) and hydrogen pressures ranging between 12.2 and 13.6 MPa. Compared with the other processes that use catalysts of Types I through V this operation gives lower yields of middle distillates ranging from 92 to 94%, but enhances cetane number more through cracking of partially saturated polycyclic compounds in the feed to lighter monoaromatics and more paraffinic products (Hill and Peterson, 1982). The loss in middle distillate yield shows up as good quality naphtha. The process is used by Petro-Canada in Edmonton, Alberta, for cetane improvement of syncrude middle distillate (Brunn et al., 1984). Two similar units are under construction, the first as part of the New Grade Energy project in Regina, Saskatchewan and the second, for selected middle distillates, at Syncrude Canada Ltd., Fort McMurray, Alberta, to be completed in 1988. Gulf/Chevron and Unocal are the major promoters of this approach.

Generally, as one follows through from Type II to Type VI processes capital costs increase because of rising operating pressures and temperatures. However, through this sequence, the sulphur tolerance also rises markedly. In Canadian middle distillate scenarios the lack of sulphur tolerance of platinum and nickel metal catalysts necessitates an in-depth hydrodesulphurization (HDS) of the feedstock. This HDS step is not well defined since there has been little demand for ultra low sulphur middle distillates - except in certain solvents.

So far, aromatics hydrogenation remains the most viable route for the short and the long term Canadian middle distillate situation. The following sections give a preliminary assessment and comparison of two hydrogenation processes: the first uses a Type IV catalyst in a "conventional" Ni-Mo route and the second a Type III catalyst in a low severity Ni route.

TECHNICAL AND COST COMPARISON OF LOW SEVERITY Ni ROUTE AND HIGH SEVERITY Ni-Mo ROUTE FOR HYDROGENATION OF MIDDLE DISTILLATES

TERMS OF REFERENCE AND PROCESS ARRANGEMENTS

Selected Canadian Middle Distillate Feedstocks

For comparison of the low and conventional high severity processes, the following four Canadian feedstocks were selected for analysis. They cover the range of middle distillates which

Canadian refiners would normally expect to process:

(i) Suncor Inc. Kerosene

Boiling marginally higher than conventional kerosene (from 190 to 340°C), this feedstock is a major synthetic crude component derived from oil sands. The jet fuel fraction generally has an aromatic content from 23 to 30 vol % which does not meet the product specification, and hence the Jet A smoke point is below standard. Suncor kerosene also has a cetane number which is well below the specification of 40. The feedstock considered for this study had 200 ppm sulphur and 30 vol % aromatics.

(ii) Syncrude Canada Ltd. Middle Distillate

This feedstock represents the largest output of synthetic middle distillate from oil sands. The quality is sufficiently low to require construction of additional hydroprocessing units for improvement of jet and diesel fuel fractions. The feedstock is expected to increase in availability by 50 to 70% over the next 10 years. At present only a single hydrotreater is under construction. However, the quality of the distillate may improve in the mid 1990's if added deep hydrotreating is installed as part of a major expansion. In this study only the addition of a single hydrotreater is assumed together with a feedstock having 400 ppm sulphur and 36 vol % aromatics.

who is building?

(iii) Blend of Virgin Diesel (70%) and Light Cycle Oil (30%)

If a stringent diesel aromatic specification is imposed, it is expected that this blend would become the major feedstock requiring aromatics hydrogenation in Canada. In this study the distillate is assumed to contain 2000 ppm sulphur and 41 vol % aromatics.

low sulphur

(iv) Light Cycle Oil

This is the poorest quality feedstock considered. In an average Canadian refinery with a catalytic cracking facility, the availability represents only about 20% of total middle distillates processed hence a smaller hydrogenation unit is deemed appropriate for the purpose of the study. A feedstock containing 3000 ppm sulphur and 70 vol % aromatics is considered.

low sulphur

In cases (iii) and (iv), existing low pressure hydrotreating facilities have been assumed available. Additional conventional hydrotreating capacities are expected in Canadian refineries due to probable reduction in the allowable sulphur content of middle distillates in the early 1990's. The facilities will provide bulk desulphurization to the 2000 to 3000 ppm level and olefin saturation at lower costs than if performed in added equipment attached to aromatic saturation units. Aromatic and sulphur contents of all four feedstocks are shown in Tables 2 and 3.

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Product Quality

A product quality target of below 20 vol % total aromatics and below 0.05 wt % sulphur is assumed in all cases studied. Since the Ni Route has sufficient activity to generate a product with a much lower aromatic content, to minimize costs for both syncrude distillates, a portion of the feed is assumed to by-pass the hydrogenation reactor. Such an arrangement allows some polycyclic aromatics to enter the final product. Therefore such products, for any given total aromatic content, could be of poorer quality than those obtained by passing the total feed over the nickel catalyst. Nickel catalysts tolerate approximately 1 ppm each of sulphur and nitrogen. In this analysis all four Ni Route cases have the necessary facilities to remove catalyst poisons either from the total feed or from streams passing only through the hydrogenation reactors. No cost credit has been given to products containing less than 0.05 wt % sulphur.

General Process Arrangements

The study considers a throughput of 2000 m^3/d for feedstocks (i) through (iii). A throughput of 1000 m^3/d would better suit the scenario for (iv) the light cycle oil feedstock. The number of reactors and/or catalyst beds varies from case to case. In principle, the temperature of reactors is controlled by a cooled quench gas in the conventional Ni-Mo Route and by a cooled liquid recycle in the low severity Ni Route. The process battery limits are delineated by feedstock pump systems, make-up hydrogen compressors and facilities for product recovery.

Based on different sulphur tolerance of catalysts a complete middle distillate upgrading can be accomplished in one reaction stage for the Ni-Mo Route but in two stages - sulphur/nitrogen removal and aromatic hydrogenation - for the Ni Route. Only in feed cases with very low sulphur content can the sulphur removal stage in the Ni Route be modified as demonstrated below.

COMPARISON OF LOW AND HIGH SEVERITY HYDROGENATION PROCESSES

Conventional High Severity Process (Sulphided Ni-Mo Catalyst)

Table II summarizes the principal operating parameters for each feedstock case for a process based on sulphided Ni-Mo catalyst, e.g. Katalco NM-506 or equivalent). Table II also gives details of sulphur and aromatic contents of feedstocks and products. It should be noted that, as the compositions of the feedstocks change from (i) light Suncor kerosene to (iv) light cycle oil, the aromatic hydrogenation and hydrotreating severities increase. Consequently, the process severity expressed in terms of liquid hourly space velocity (LHSV) varies from 2.0 to 0.8, the reactor temperature from 280 to 335°C and the pressure from 10.1 to 15.2 MPa.

Figure 1 shows a simplified process flow sheet. Apart from reactors with specially designed multiple catalyst beds and recycle gas quench systems for close temperature control the Ni-Mo Route would be identical to any commercial hydrotreating process. The number of reactors and catalyst beds vary from case to case, which is indicative of the high exothermic heat of aromatics hydrogenation, particularly with light cycle oil.

Low Severity Process (Nickel Catalysts) with HDS/HDN First Stage

Table III summarizes, for each feedstock case, the principal parameters for both process stages. Stage 1 is a hydrodesulphurization/hydrodenitrogenation (HDS/HDN) reactor utilizing a sulphided Co-Mo catalyst (Akzo Chemie 742 or UOP S12 or equivalent). Stage 2 is the aromatics hydrogenation step and employs a $Ni/SiO_2-Al_2O_3$ supported catalyst (Harshaw Ni-3288 or equivalent). Table III also gives the reactor system throughput and by-pass for each feedstock and the number of reactors for each stage. Also shown are the aromatic and sulphur contents of the feedstocks and of the intermediate and final products. The process parameters do not vary significantly from case to case. For the HDS/HDN stage, the LHSV varies between 0.8 and 2.0, the temperature from 349 to 366°C and hydrogen pressure from 8.4 to 10.5 MPa. For the hydrogenation process the LHSV varies between 4.0 and 1.5, the temperature from 160°C (start of run) to 250°C (end of run) and hydrogen pressure from 3.2 to 4.2 MPa.

Figure 2 presents a simplified process flow sheet. Compared with the conventional Ni-Mo Route presented above, the two-stage Ni Route is more complex. The HDS/HDN stage is itself similar to the Ni-Mo Route, but operates at a lower pressure and with a reduced catalyst inventory. The hydrogenation stage includes guard bed reactors operating in the liquid phase between 160 and 220°C and in the absence of hydrogen. With the exception of Suncor kerosene, in all cases the hydrogenation stage features a feed split between multiple reactors. The hydrogenation reactor temperature is controlled by the cooled liquid recycle as shown in Figure 2. For this study, fixed bed hydrogenation reactors are considered, but for better temperature control slurry or ebullated bed reactors may be useful.

Nitrogen and water are nickel catalyst poisons at a level of approximately 1 ppm, hence efficient denitrogenation and water stripping are required in the first stage. To remove most of the feed sulphur the sulphided Co-Mo catalyst has to operate at relatively high temperatures at which traces of olefins may form. Newly formed organic sulphur compounds resulting from a recombination of olefins with H_2S may pass through the stripper. It is therefore assumed that the product from the HDS/HDN stage contains 5 ppm of sulphur. To maintain the life of the nickel catalyst for a reasonable period, guard beds are placed in front of the hydrogenation reactors. Fresh and deactivated nickel catalysts have significant capacity to capture organic sulphur compounds. However spent nickel catalyst cannot be a self-sustained sulphur adsorbent since it "bleeds" through a portion of sulphur from the HDS/HDN stage. The study considers the use of an adsorbent, Harshaw Ni-0750, a cheaper version of the hydrogenation stage Ni catalyst, or equivalent. This adsorbent retains about 10 times as much sulphur as the fresh nickel catalyst. It is capable of complete removal of organic sulphur compounds from the HDS/HDN stage product. Use of the spent Ni hydrogenation catalyst for capture of residual sulphur in combination with the above adsorbent could however reduce total costs.

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Stand Alone Ni Route

In some refinery operations, kerosene streams which are hydrotreated along with naphtha may yield kerosene products with maximum nitrogen and sulphur contents of approximately 1 and 10 ppm respectively. Kerosenes from hydrocracking of syncrude heavy gas oils have about these levels of sulphur but higher amounts of nitrogen. These kerosenes have aromatic contents of about 30 vol % and are not acceptable as Jet A fuel products. The aromatics could be reduced to a very low level in a modified Ni Route process unit without the HDS/HDN stage provided that the excess feed sulphur is removed by a nickel adsorbent in multiple guard beds.

Figure 3 shows the modified Ni Route process without the HDS/HDN stage. For a feedstock containing 10 ppm sulphur and 30 vol % aromatics the study considers a reactor system throughput of 800 m^3/d. This unit could deliver at the rate of 2000 m^3/d of kerosene product with 20 vol % aromatics and no sulphur. A third guard bed has been added in addition to the two units employed in the above case with prior HDS/HDN.

COST COMPARISON OF HIGH AND LOW SEVERITY PROCESSES

Ni-Mo Route versus Ni Route

Cost estimates for the four middle distillate feedstock cases applied to the two process routes are based on 1987 Western Canadian prices. The estimates omit a slightly lower liquid yield for the Ni Route caused by the higher temperature of the HDS/HDN stage which leads to more hydrocracking. Also the hydrogen demand for the Ni Route is slightly higher due to additional losses in gas/liquid separation in the hydrogenation stage.

Table IV presents cost comparisons for the Ni-Mo and Ni Routes. In all feedstock cases studied, the capital costs of the Ni Route are more or less equal with Ni-Mo. However, the operating costs are marginally higher due to nickel catalyst non-regenerability, marginally poorer energy efficiency and greater process complexity. It is important to note that the major expenses of the Ni Route lie in the HDS/HDN stage which carries approximately 75% of the capital and 65% of the operating costs (exclusive of hydrogen make-up costs).

Current (1987) prices of the non-regenerable, low sulphur and low nitrogen tolerable nickel catalyst and the nickel-based sulphur sorbent are about \$37 and \$27/kg(Can.), respectively. It is believed that price reduction due to increased production volumes of both the nickel catalyst and the adsorbent would significantly improve the economic feasibility of the Ni Route. Other factors having a positive effect would be an improved sulphur and nitrogen tolerance and catalyst regenerability.

The capital cost estimates of the study case for light cycle oil hydrogenation, a 1000 m^3/d throughput unit, was compared with a similar estimate made for the AH Unibon process provided by UOP Inc. This is a similar unit using a platinum catalyst. A small (5%) difference in the cost estimates for nickel and platinum was found and confirmed the adequacy of the methods used in this study.

Cost of the Stand Alone Ni Route

Including costs for the nickel catalyst and adsorbent the estimated capital costs of a 2000 m^3/d stand alone unit for processing the low sulphur feedstocks is estimated at $10 million Can. and operating costs, exclusive of hydrogen make-up costs, at $2.1/m^3(Can.).

CONCLUSIONS

Figure 4 provides a simplified overview of the "fit" of various middle distillate aromatics hydrogenation routes as developed in this study of four representative Canadian feedstocks. The work concludes that a low severity hydrogenation route using certain types of commercial nickel catalysts will be competitive with platinum for distillate aromatics conversion when feedstock sulphur is below 5 ppm. In certain circumstances a nickel route will also be competitive with platinum in the 5 to 15 ppm range with enhanced guard beds. When the feedstock sulphur level is above 5000 ppm and an HDS stage is required by all available routes, the nickel route may compete effectively with sulphided Ni-Mo and Ni-W catalysts. All hydrogenation routes are very capital intensive especially above 15 ppm sulphur, with about 75% of the capital costs for Pt and Ni routes being in the HDS/HDN stage.

The ability of mild middle distillate hydrocracking to improve cetane number beyond that possible through aromatics hydrogenation alone is to be noted. However, the mild hydrocracking route may not provide enough aromatics saturation to attain Jet A or future diesel fuel aromatic specifications. The requirement for installation of new aromatics hydrogenation plants in Canada by 1996 will be determined by the imposition of a new aromatic specification for jet and diesel fuels.

It is important to note the major role which sulphur tolerance plays in the selection of an aromatics hydrogenation process. The non-regenerability of current nickel catalysts is also a major disadvantage. There is a need to investigate optimum HDS/HDN conditions and catalysts to determine if the high capital costs of this stage can be reduced.

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FIGURE CAPTIONS

1. Simplified flow diagram of a conventional aromatics hydrogenation process using a sulphided Ni-Mo catalyst.

2. Simplified flow diagram of a low severity aromatics hydrogenation process using a supported Ni catalyst combined with an HDS/HDN step.

3. Simplified flow diagram of a stand alone aromatics hydrogenation process using a supported Ni catalyst.

4. Approximate economic fits of middle distillate aromatics hydrogenation processes.

Table I - Anticipated Changes in Canadian Middle Distillate Fuel Specifications

Product	Component	Year 1987	Year 1992	Year 1996
Road diesel fuel	Sulphur, wt %	0.5(a)	0.15	0.05
	Aromatics, vol %	(b)	(b)	25 (c)
	90% point, °C	360	360	340
Railway diesel fuel	Sulphur, wt %	(b)	0.15	0.05
Jet fuel type A, A1 and A2	Sulphur, wt %	0.2	0.1	0.05
	Aromatics, vol %	22(d)	22(e)	22(e)
Light fuel oil No. 2	Sulphur, wt %	0.5(a)	0.15	0.15(f)

(a) An internal refinery specification is generally in use in Canada.

(b) No standard.

(c) May be stated as the equivalent of alkylbenzenes with a penalty for polynuclear aromatics.

(d) Can be used with notification up to 25 vol % aromatics.

(e) No allowance above 22 vol % aromatics.

(f) a combustion quality specification may be added to reduce aromatic content.

Table II - Conventional Ni-Mo Route Hydrogenation Process Parameters

Feedstock	Suncor kerosene	Syncrude middle distillate	Blend of conv. diesel oil & LCO[a]	LCO[a]
Feed rate, m^3/d	2000	2000	2000	1000
Sulphur in/out, ppm	200/ 100	400/ 200	2000/ 500	3000/ 500
Aromatics in/out, vol %	30/20	36/20	41/20	70/20
LHSV	2.0	1.5	1.0	0.8
Average reactor temperature, °C	280	280	315	335
Pressure, MPa	10.1	11.5	13.1	15.2
Catalyst	Katalco NM-506 (Ni-Mo) or Shell 4545 (Ni-W) or equivalent			

[a] Light cycle oil

Table III - Low Severity Ni Route Hydrogenation Process Parameters

Feedstock	Suncor kerosene	Syncrude middle distillate	Blend of conv. diesel oil & LCO[a]	LCO[a]
Total feed rate, m^3/d	2000	2000	2000	1000
Reactor system				
throughput, m^3/d	800	1100	2000	2000
by-pass, m^3/d	1200	900	0	0
<u>Hydrodesulphurization/hydrodenitrogenation stage</u>				
Sulphur in/out, ppm	200/5	400/5	2000/5	3000/5
Aromatics in/out, vol%	30/30	36/36	41/41	70/70
LHSV	2.0	1.5	1.2	0.8
No. of reactors	1	1	1 or 2	2
SOR[b] reactor temperature, °C	349	354	354	354
EOR[c] reactor temperature, °C	360	366	366	366
Pressure, MPa	8.4	10.5	10.5	10.5
Catalyst	Co-Mo on alumina - Akzo 742, UOP S12 or equivalent			
<u>Hydrogenation stage</u>				
Sulphur in/out, ppm	5/0	5/0	5/0	5/0
Aromatics in/out, vol%	30/5	36/5	41/20	70/20
LHSV	4.0	3.0	3.0	1.5
No. of reactors	1	2	2	3
SOR[b] reactor temperature, °C	160	170	170	170
EOR[c] reactor temperature, °C	250	250	250	250
Pressure, MPa	3.2	4.2	4.2	4.2
Catalyst	Ni on silica alumina - Harshaw Ni-3288 or equivalent			

[a] LCO - light cycle oil
[b] SOR - start of run
[c] EOR - end of run

Table IV - Cost Comparison of Ni-Mo and Ni Routes (1987 Canadian dollars)

Feedstock	Suncor kerosene	Syncrude middle distillate	Blend of conv. diesel oil & LCO[a]	LCO[a]
Total feed rate, m^3/d	2000	2000	2000	1000
	Capital costs, $10/$m^3$/d			
Ni-Mo Route	10.2	12.0	14.8	29.5
Ni Route	10.8	13.2	17.2	32.4
	Operating costs, $/$m^3$[b]			
Ni-Mo Route	2.6	2.7	3.6	3.5
Ni Route	3.2	4.1	6.2	5.6

[a] LCO - light cycle oil

[b] Operating costs exclusive of hydrogen make-up costs

FIG. 1

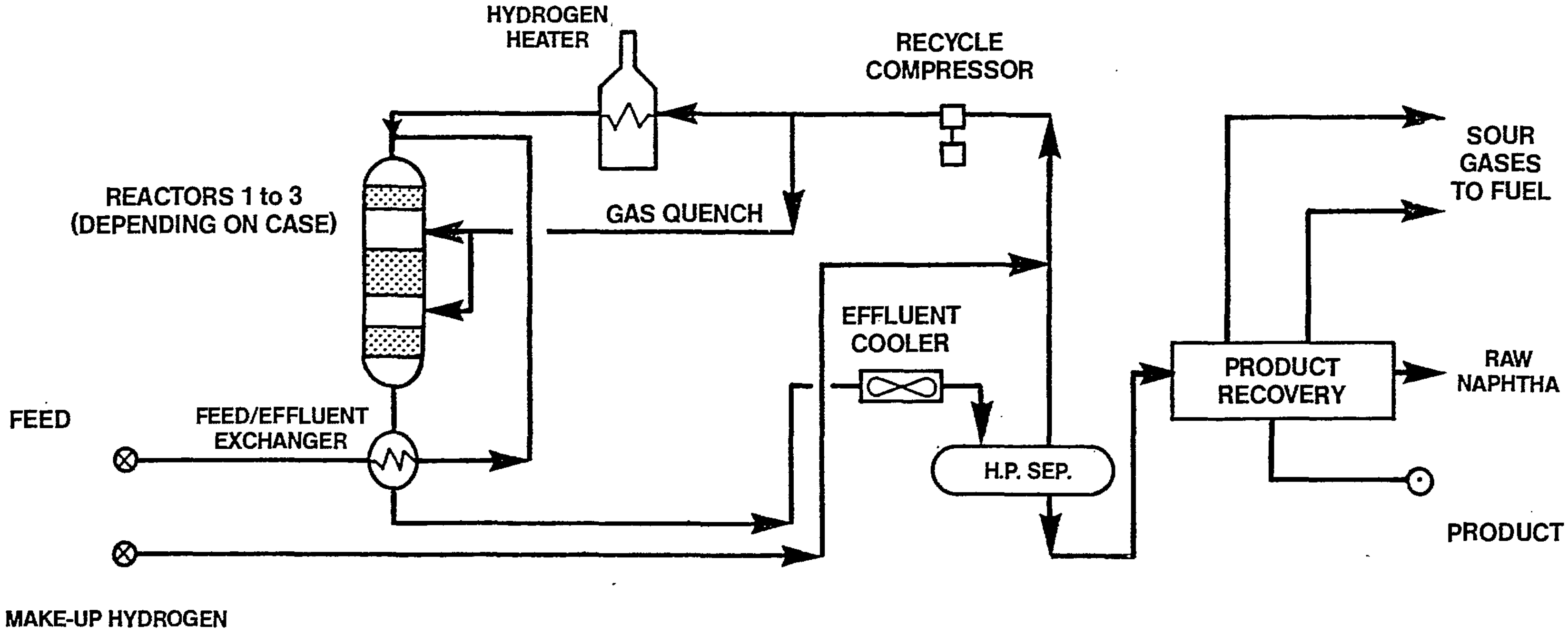

HYDROGEN HEATER
RECYCLE COMPRESSOR
REACTORS 1 to 3 (DEPENDING ON CASE)
GAS QUENCH
SOUR GASES TO FUEL
EFFLUENT COOLER
PRODUCT RECOVERY
RAW NAPHTHA
FEED
FEED/EFFLUENT EXCHANGER
H.P. SEP.
PRODUCT
MAKE-UP HYDROGEN

FIG.2

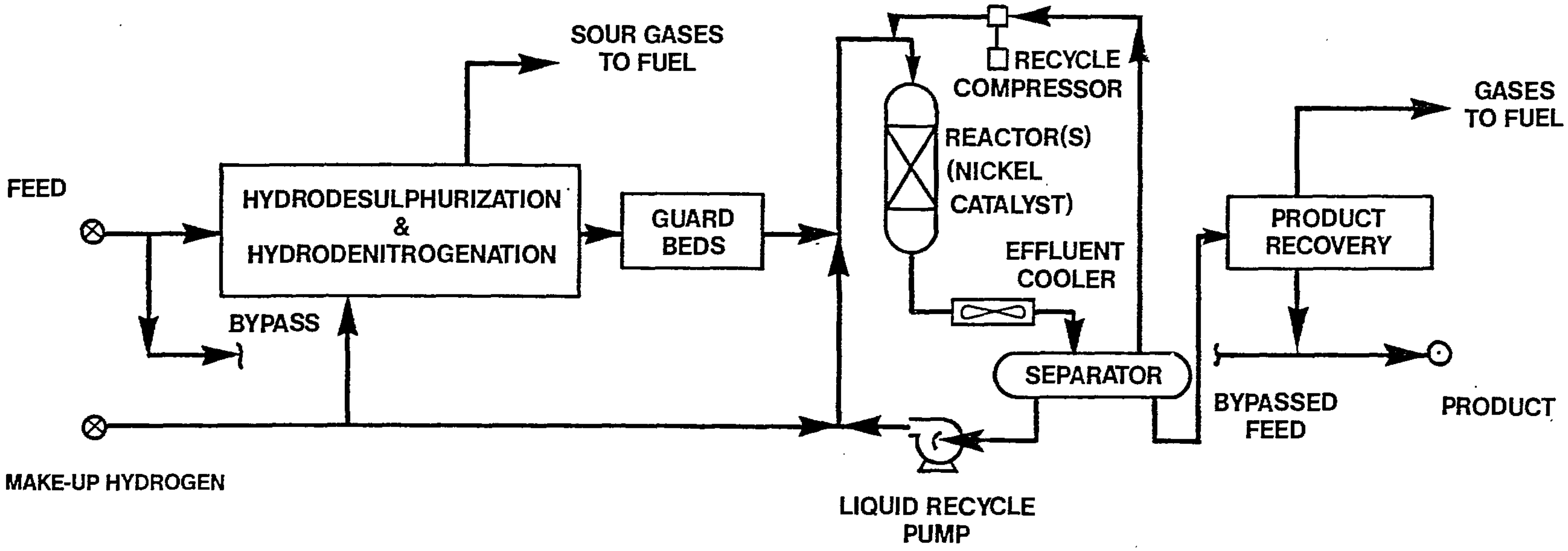

SOUR GASES TO FUEL
RECYCLE COMPRESSOR
GASES TO FUEL
REACTOR(S) (NICKEL CATALYST)
FEED
HYDRODESULPHURIZATION & HYDRODENITROGENATION
GUARD BEDS
PRODUCT RECOVERY
EFFLUENT COOLER
BYPASS
SEPARATOR
BYPASSED FEED
PRODUCT
MAKE-UP HYDROGEN
LIQUID RECYCLE PUMP

FIG. 3

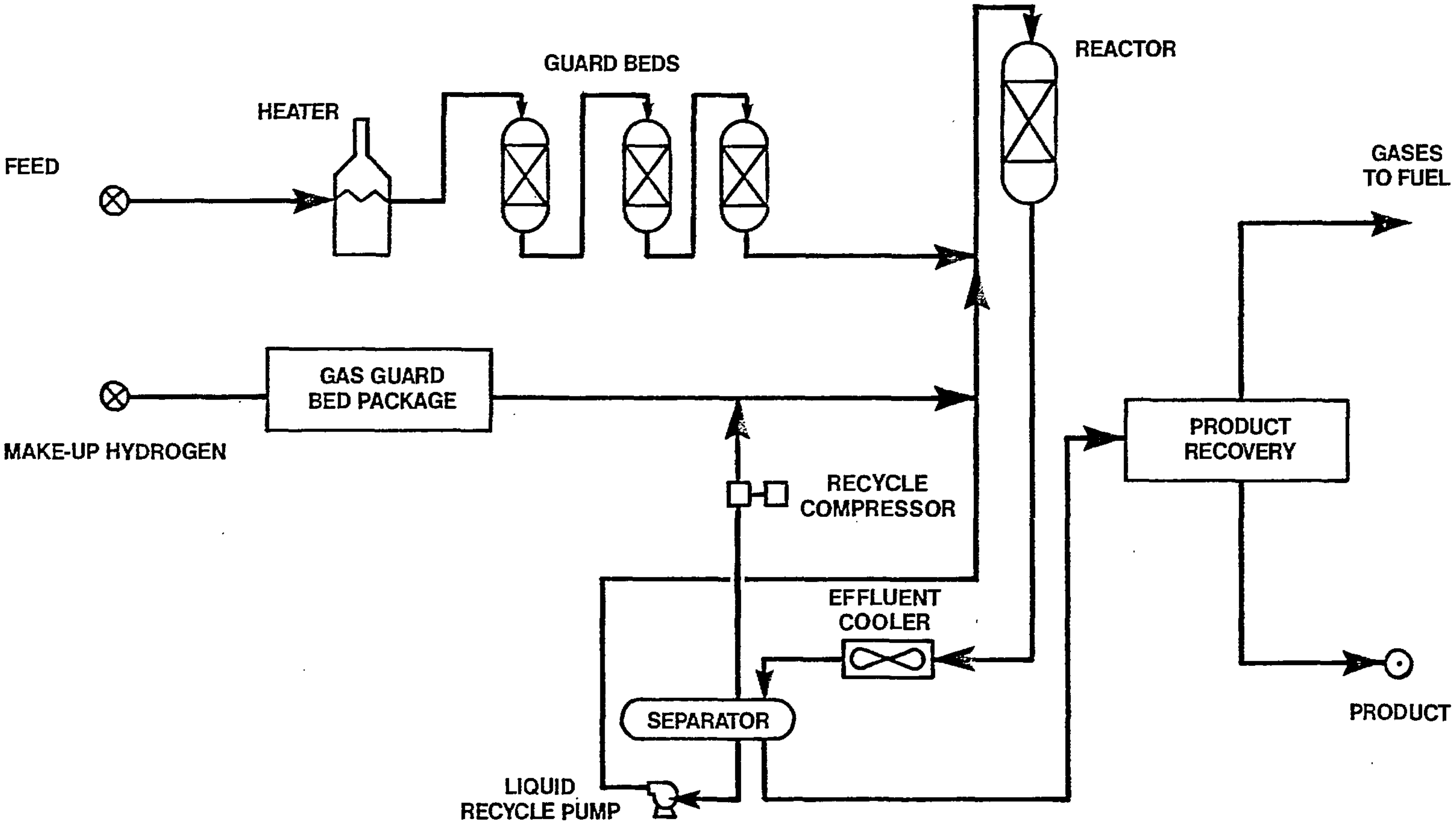
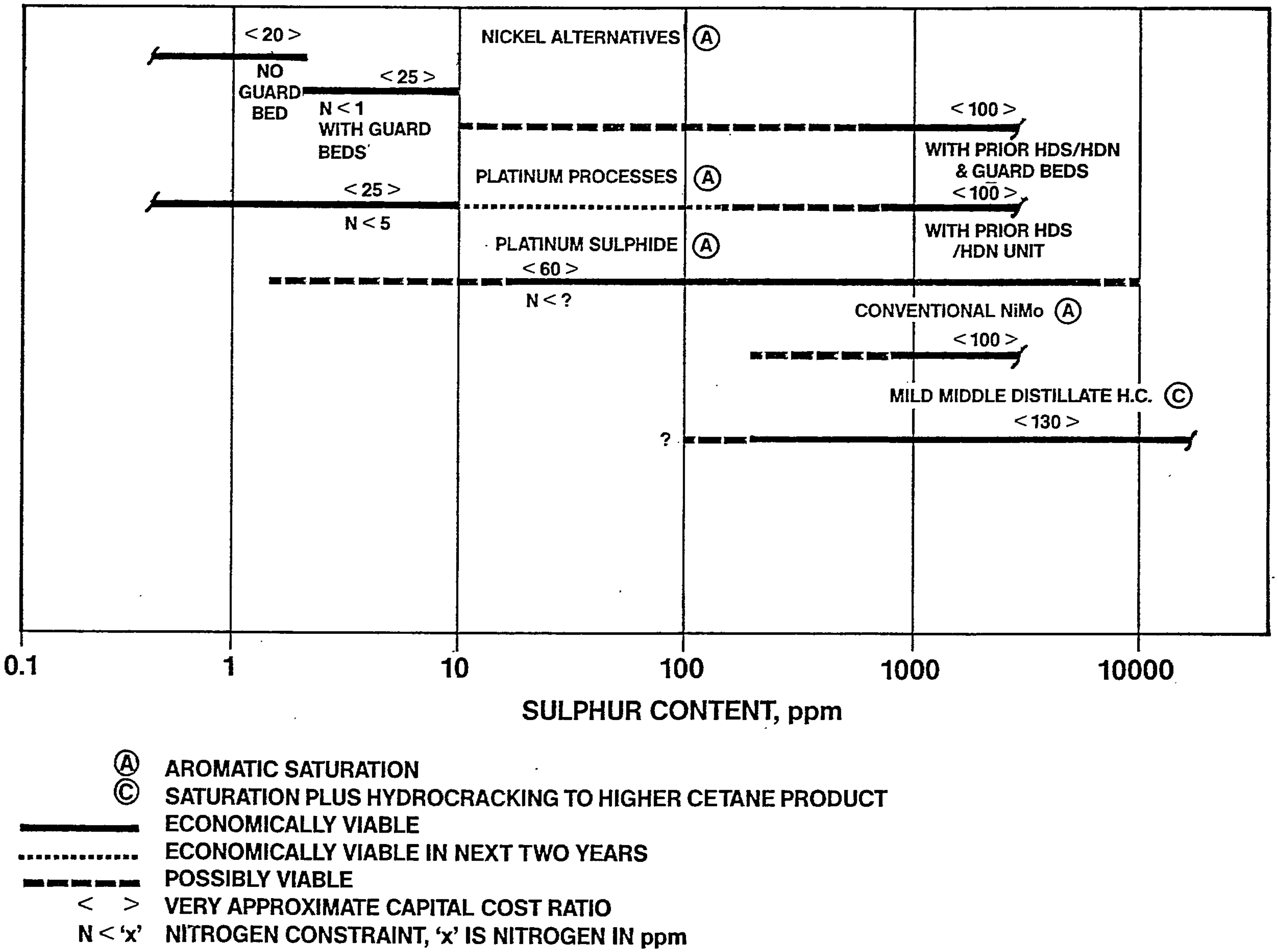

GUARD BEDS
REACTOR
HEATER
FEED
GASES TO FUEL
GAS GUARD BED PACKAGE
MAKE-UP HYDROGEN
PRODUCT RECOVERY
RECYCLE COMPRESSOR
EFFLUENT COOLER
SEPARATOR
PRODUCT
LIQUID RECYCLE PUMP

FIG. 4



NICKEL ALTERNATIVES Ⓐ
< 20 >
NO GUARD BED
< 25 >
N < 1
WITH GUARD BEDS
< 100 >
WITH PRIOR HDS/HDN & GUARD BEDS
PLATINUM PROCESSES Ⓐ
< 25 >
N < 5
< 100 >
WITH PRIOR HDS /HDN UNIT
PLATINUM SULPHIDE Ⓐ
< 60 >
N < ?
CONVENTIONAL NiMo Ⓐ
< 100 >
MILD MIDDLE DISTILLATE H.C. Ⓒ
< 130 >
?
0.1
1
10
100
1000
10000
SULPHUR CONTENT, ppm
Ⓐ AROMATIC SATURATION
Ⓒ SATURATION PLUS HYDROCRACKING TO HIGHER CETANE PRODUCT
ECONOMICALLY VIABLE
ECONOMICALLY VIABLE IN NEXT TWO YEARS
POSSIBLY VIABLE
< > VERY APPROXIMATE CAPITAL COST RATIO
N < 'x' NITROGEN CONSTRAINT, 'x' IS NITROGEN IN ppm