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3rd International Symposium

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New Technologies and Varying Feedstocks

174th Event of the European Federation of Chemical Engineering lcp3

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Contents

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Project planning problems in ethylene production

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The large investment required at present for a new ethylene cracker calls for a careful consideration of all factors influencing the economic viability of such a project. It is therefore imperative to incorporate in the planning stage special studies of parameters which influence the economy of such a large project. In this paper a number of these factors will be highlighted.

I. ESTABLISHING THE DEMAND FOR OLEFINS

The planning cycle of an ethylene plant usually starts with an analysis of the potential down-stream demand position for its prime products in a certain area. The area being studied will be larger if the plant can be connected to an ethylene pipeline grid. While the benefits of the economy of scale will often suggest the choice of a larger size plant, it may prove to be difficult to find an indigenous consumption for the additional capacity to be installed. The study of the demand position should therefore analyse the comprehensive picture for the designated area.

II. SIZING OF THE UNIT

A competitively sized plant means a size at least similar to that of the competitors, so as to match its unit costs of production. Since the early days of ethylene manufacture in W-Europe plant sizes have increased fifty-fold and this scale-up has played a major part in reducing the ethylene manufacturing costs. Notwithstanding inflation and an enormous increase in hydrocarbon prices, ethylene costs have not increased proportionally. However, in the prevailing economic conditions it seems appropriate not to overemphasize the cost benefit of sizing above, say 350,000 mta, since other factors then will start to play a more dominant role. In this context it is evident that the decision to increase one's olefin capacity will coincide with the decision to increase the indigenous downstream consumer capacity of the location under consideration. Even so, it will often prove to be difficult to cope with the step increasing resulting from the commissioning of a cracker of 350,000 mta or larger. In such a case it seems logical to reach an agreement with other manufacturers to achieve a committed offtake for part of the production. In this way also a faster load build-up can be achieved, which is vital to the economics of such a project. However, such agreements introduce a strong interdependence into the project planning of the Petrochemical firms concerned. Sometimes therefore, a closer form of cooperation will be introduced in the form of a Joint Venture or Industry cracker concept. In such a case agreement has first to be reached between the partners on the project scope definition, the form of partnership contemplated, timing, etc. Whatever approach is used to secure a sound economic future for a project, it is clear that the initial planning stage of the decision making process can be extremely complex and time consuming.

III. FEEDSTOCK OPTIONS

An area of particular scrutiny is the feedstock selection. For the conventional tubular cracking of crude oil distillation cuts, all fractions can potentially be used as feedstock, except the residual tars. Since the choice ranges from ethane to vacuum distillate, an analysis should first be made of the longer-term supply/demand situation in the market area considered.

Assuming that free market forces determine the price levels of the various fractions, it is a difficult task for the planner to predict the hydrocarbon values five upto fifteen years into the future. However, before trying to quantify the above, it is necessary to consider the relationship between the cracking severity and the yield patterns resulting from the main feed fractions, so as to lay a proper foundation for an economic feedstock comparison.

In Figs. 1, 2 and 3 yield patterns from the main feedstocks e.g. naphtha, straight run gasoil and vacuum gasoil (also referred to as vacuum- or waxy distillate) are given. For simplicity a modern, short residence time furnace has been selected, using the tube outlet temperature as the only severity criterium. Ethane is recycled and cracked to extinction. As optimum cracking severity we have selected conditions where the sum of ethylene, propylene and butadiene production reaches a maximum, since these prime products always have a value higher than that of the other products. The severity level at which the sum of these products is maximised therefore often provides the best return per ton of feedstock.

Fig. 1, 2 and 3 on separate page.

Moreover, for SR gasoil and for vacuum gasoil it so happens that other severity criteria related to fouling rates in the transfer line heat exchanger limit the cracking severity as similar conditions.

In table I the yield data for these chosen severities are summarised.

(Table I on separate page)

Returning to the issue of feedstock selection a general relationship can be developed indicating at which price a specific feedstock is to be preferred for ethylene manufacture. To develop this relationship we postulate the feed to be hydrodesulphurised, the ethylene unit to have a capacity of 450,000 mta and to include a butadiene extraction unit and a selective pyrolysis gasoline hydrotreater (diolefinic saturation). The assumed investment costs are as follows (present day value) :

Naphtha cracker	US $ 180 million
Gasoil cracker	US $ 208 million
Vac. distillate cracker	US $ 212 million

Taking a 30 % capital charge a relationship can be worked out between the feedstock value, the fuel value and the ethylene value, by developing the following equations :

N (G, VD) = f (E, P, B, Ga, F) in which

N = naphtha value
G = gasoil value
VD = vac. dist. value
E = ethylene value
P = propylene value
B = butadiene value
Ga = gasoline value
F = fuel value

The relationship between the ethylene value and the propylene and butadiene values is rather arbitrary. However, the results are not very sensitive to the assumptions, since at the conditions taken the prime products are produced in approximately the same ratios. The relationship between gasoline, naphtha and fuel value is derived from the cost of producing gasoline of the same quality via the naphtha reforming route. The ethylene manufacturing costs can be expressed as a function of the fuel value (or fuel to naphtha ratio). Inserting the ethylene value found in the equation for gasoil and vacuum distillate, the relationship between gasoil (vacuum distillate) and fuel values can be calculated for a given naphtha value. Assuming a naphtha value of US $ 135/mt, the result is presented in fig. 4 as a function of the fuel/naphtha.

(Fig. 4 on separate page)

As can be seen from this example gasoil cracking becomes economically feasible at a gasoil value of some US $ 15-20/mt lower than naphtha. For vacuum distillate this differential ranges from US $ 20-30/mt. It is emphasized that a cracker designed for a flexible feedstock diet will demand an additional investment and therefore a higher differential between the various feedstocks indicated. However, it has the advantage of being able to benefit from the seasonal swing in hydrocarbon values experienced in Europe. Although these relationships are useful as indicators of the different price levels needed to give the Chemical Industry the incentive to change over from the conventional naphtha cracking to heavier feedstocks, it does not necessarily follow that such margins will actually materialise. In a free market economy the Oil Industry itself will look upon the heavier fractions as potential feedstock for conversion into gasoline components, since crude oil supplies will be more limited in the longer term. The ethylene cracker can be seen in this context simply as another conversion process in the same way as thermal crackers, cat. crackers and hydro-crackers. Therefore its competitive place among these processes should also be ascertained.

IV. DESIGN BASIS, CAPITAL ESTIMATION AND TIMING

1. Design basis

Even before the start of the process design phase of a new project a decision will have to be taken by the prospective manufacturer as to how much novel technology he will incorporate in his design. This question is even more important now than it used to be, due to the steep rise in feedstock and energy costs. Cost saving technology reducing energy consumption or improving the utilisation of the feedstock has met with a more acute interest from the olefin manufacturers. It is therefore necessary, carefully to consider during the project planning stage how much "calculated risk" can be accepted. Larger companies have the expertise and the facilities to carry out an independent assessment of proposed novel design features. In the pyrolysis section for example, the testing of prototype furnace designs can be done on a fairly large pilot unit before being introduced on the commercial scale. Most larger companies will also have accumulated sufficient know-how to have formulated pyrolysis computer models, which are able to give a good advance indication of probable yield gains, furnace- or TLX coking tendencies, etc. In this way the incentive for a large scale testing can already be established. In the energy saving sector for instance, know-how will have to be collected on the recovery of waste heat, particularly important for gasoil and vacuum distillate cracking.

Also of considerable importance are a number of decisions on conventional process elements, forming the design basis of the plant and influencing its operability, its flexibility, its safety and therefore its costs. Many decisions of this type will be based upon a company's experience, i.e. judgement rather than exact scientific calculations.

2. Timing and capital cost estimates

The completion of the ethylene project should be synchronised with the development of demand for derivatives. The ethylene plant being by far the largest unit of investment, and frequently entailing the most cumbersome procedure to obtain building and operating permits, is very clearly on the critical path in the development of the petrochemical complex. A rapid build-up of production after completion of the project requires a careful timeplan for the starting dates of contracts for the sale of the products of the ethylene plant and/or the initiation and execution of the downstream projects. The planner of today is particularly hampered in this respect by the difficulty of predicting the completion time of an ethylene project. Some 5 years ago ti was normal to count on completion times of 24 to 30 months after award of contract and freezing of the design basis; at present periods of 33 to 40 months are being mentioned. The delivery time of equipment is especially important and the critical items are not only the main compressors and drivers, but for instance also the steam boilers. The general turbulence in the world economy also affects the reliability of capital cost estimates. Superimposed on the strong variations in equipment prices, is a general inflationary trend of a rather unpredictable magnitude. However, this factor and the consequent difficulty of producing a credible numerical analysis of profitability is common to all sectors of the Hydrocarbon processing industry and therefore is not further discussed in this paper.

Reference

The main contents of this paper have previously been presented by J.D. van Dalen of Shell Internationale Chemie Mij., The Haghe and were published in Chemical Engineering Progress of June 1975.

Table I - Yields and properties of various feedstocks

	Naphtha	Gasoil	Vacuum distillate
Density, 15°C	0.705	0.853	0.8775 (20°C)
Mol. Wt.	95	268	367
C/H ratio	5.514	6.353	6.545
Dry gas, % wt	15.7	10.7	10.6
Ethylene*	31.5	25	23
Propylene	14.4	13.8	13
Butadiene	4.6	4.5	4
BTX	15.5	12.0	7.9
Cracking temp. °C	86	815	783

*Including ethane cracking

Figure 1: Straight run naphtha

S.g. 0.705
Mol. Wt. 95
C/H 5.514

E = ethylene
BD = butadiene P = propylene
BTX = aromatics M = methane

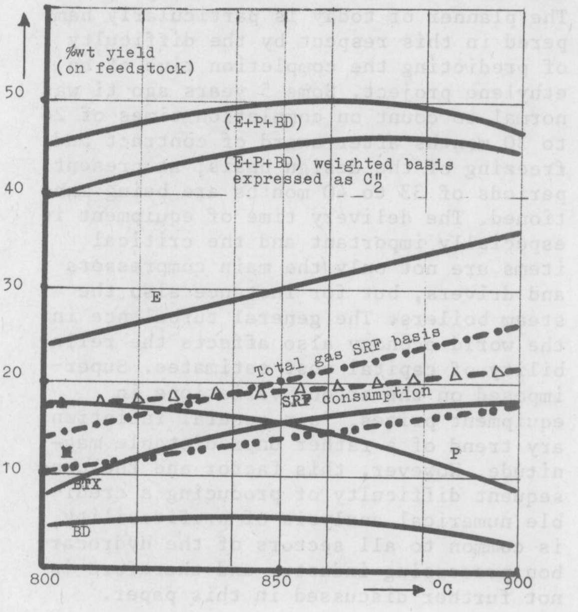


Figure 2: Straight run gasoil

S.g. 0.853
Mol. Wt. 268
C/H 6.353

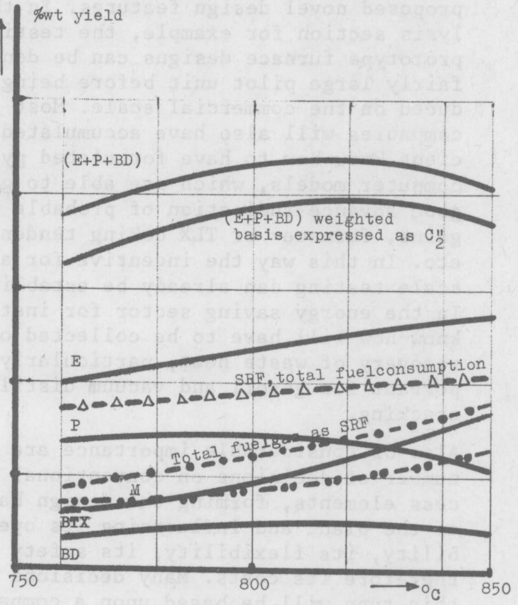


Figure 3: Vacuum distillate

S.g. 0.8775
Mol. Wt. 367
C/H 6.545

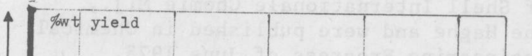
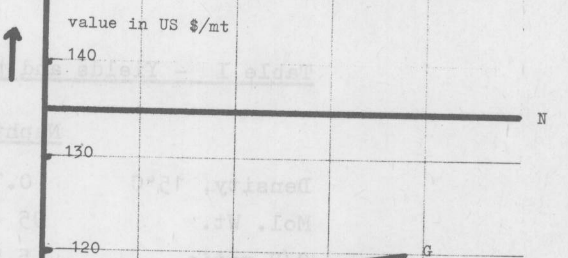


Figure 4

Break even values of hydrocarbons as ethylene unit feedstock with naphtha at US $135/mt

N = naphtha
G = gasoil
V = vacuum distillate (hydro desulphurised)
F = fuel



Flexibility of large ethylene plants

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1. ABSTRACT

The ethylene plant feedstock availability and pricing has been reviewed. As no reliable assessments can be made on the availability and pricing of liquid feedstocks, the feasibility of liquid feedstock flexibility has been studied.

It has been shown that naphtha/AGO feedstock flexibility can be achieved with some 10-15 % increase of BLCC and minor increase of utility cost. Existing naphtha crackers can be revamped for feedstock flexibility up to some 20 % gas oil capability. VGO is less suitable for feedstock flexibility and requires a major increase of BLCC and utility cost.

More reliable assessments can be made on the feedstock availability for gas crackers, reducing generally the need for feedstock flexibility. Ethane/propane cracking flexibility can be incorporated by adapting the ethylene plant recovery section already in the design stage.

2. INTRODUCTION

The so-called oil crisis has once again clearly shown that no reliable forecasts can be made with regard to feedstock availability and price over the period from plant design to taking out of operation, being some 20 years.
Ethylene plants based on liquid feedstocks give a considerable yield of byproducts. Obviously, like for the feedstock availability, no reliable forecast can be made for the products.

The purpose of this paper is to study the impact of ethylene plant feedstock and product flexibility on performance data, operability and investment.

3. FEEDSTOCK FLEXIBILITY

It is well known that the refining pattern in the USA is completely different from Europe and Japan. There are two reasons for this. Number one is that in the USA large quantities of natural gas are available, reducing the need for a high fuel oil yield. The second reason is that the motor gasoline consumption in Europe and Japan is relatively low. This has resulted up till recently in a large naphtha surplus. Obviously, the availability and pricing of crude oil fractions in Europe is more season-dependent than in the USA. Consequently, the USA ethylene plants have traditionally been based on ethane and propane feedstocks recovered from the natural gas. The European and Japanese plants were traditionally based on naphtha feedstock. Actually, this picture does not apply anymore neither for the USA nor for Europe and Japan.
The US natural gas production will slow down. Ethane and propane feedstocks can be ensured for the existing crackers by a higher recovery efficiency. New crackers should be based, at least partly, on liquid feedstocks.

Up till the oil crisis, surplus naphtha as available in Europe and Japan. However, since the oil crisis the pricing policy of naphtha has been adapted and other ethylene plant feedstocks, atmospheric and vacuum gas oil as well as gaseous feedstocks for North-West Europe are considered.

3.1 Impact of Feedstock Flexibility on Plant design

In order to study the impact of feedstock flexibility we shall review the different process steps of the ethylene plant (fig. 1, 2, 3, 4).

Feed Pretreatment

Regarding feed pretreatment there should be distinguished feed fractionation and feed hydrodesulphurisation. Feed fractionation is in question when a crude oil refinery is linked with an ethylene plant. When naphtha/kerosene and gas oil could be cocracked, savings regarding investment and utilities could be made in the topping unit. However, experiments (table 1.) clearly show that this is for the ethylene plant not a feasible solution.

Table 1. Impact of liquid feedstock fractionation on process yields

Fraction	Naphtha	Kero/ AGO	Naphtha/ Kero/ AGO
IBP	48	161	53
50 %	121	269	216
FBP	181	353	353
Spec.grav. at 15 °C	0.733	0.837	0.792
Once-through process yields, wt.%			
CH_4	16.3	11.7	12.6
C_2H_4	25.7	23.0	23.3
C_3H_6	13.0	14.0	14.4
C_4's	8.0	8.3	8.6
Pyrogasoline	23.8	19.0	21.0
Fuel oil	7.0	18.0	14.0

Note: This table shows that fractionation is attractive for high ethylene yields.

Cracking a wide range fraction leads to relativly low ethylene yields, because the lighter feedstock fraction is not cracked severe enough. Furthermore, aromatics yield is relatively low due to the same effect. Cracking of mixtures of gaseous feedstocks like ethane and propane is basically possible. Also here, co-current cracking of ethane and propane simplifies the ethane/ propane recovery from the natural gas. However, it has been shown that co-current cracking of ethane and propane negatively influences the ethylene process yield and ethane conversion. Preferably, ethane and propane should be cracked separately.

When processing atmospheric or vacuum gas oil derived of sour crudes, hydrodesulphurisation can be necessary in order to meet the fuel oil quality (sulphur content) requirements.

Cracking Heaters and Quench Coolers

Cracking heaters and quench coolers form in the ethylene plant one integrated process step regarding process and utilities (H.P. steam).

Operators basically prefer a single design for all different feedstocks. Principally, multiple feed flexible cracking heaters/quench coolers are feasible, but what is the penalty for this flexibility? Gaseous feedstocks, especially ethane, are more refractory than liquid feedstocks and require longer residence times (.5 - 1.0 sec.) in the cracking coil. The current KTI GK split coil type cracking heaters, operating in the range of 0.25 - 0.3 sec., are more suitable for liquid (less refractory) feedstocks. Heavier feedstocks than naphtha, like AGO and VGO, have a higher cracking coil coking tendency. Therefore, with heavy naphtha, AGO and VGO feedstocks no small diameter tubes should be applied near the outlet of the cracking coil in order to avoid excessive pressure drops, negatively affecting the process yields.

The ethylene yield from gas oil is considerably lower than from naphtha feedstock. Therefore, the ethylene capacity per cracking heater is also for gas oil lower than for naphtha. In order to maintain the same plant ethylene capacity on gas oil as on naphtha, more cracking heaters have to be put into operation.

As we have seen, the cracking coil is relatively flexible with regard to feedstocks, however, special care has to be taken by the designer regarding the cracking heater convection section. Problems arise from difference in heat pick-up for different feedstocks and from difference in dilution steam requirements. A centralised flue gas waste heat recovery system is more suitable for multiple feedstocks handling than an integrated or individual flue gas recovery system.

Transferline Exchangers (TLX's)

TLX'es or quench coolers generating H.P. steam have been applied for naphtha crackers since 1961. In order to limit fouling on the process side, higher steam pressure levels (80 - 125 atm) and higher quench cooler outlet temperatures (500 - 600 °C) are selected for heavier feedstocks.

For ethane crackers traditionally relative low steam pressure level has been applied for the quench coolers in the order of 40 - 60 atm. Recently, as a result of the major increase of oil price, also for ethane crackers an interest for higher steam pressure level, resulting in a higher efficiency, has developed.

KTI has pioneered application of TLX'es on atm. gas oil (AGO) with shell and tube type TLX'es with relatively small tube sizes (1 - 1.5 inch) already in 1968. Acceptable run lengths are feasible with AGO on the well-known shell and tube type as well as double-tube type TLX'es.

Processing vacuum gas oil (VGO) on cracking heaters provided with TLX'es does not seem feasible and has, according to our knowledge never been applied.

Prefractionation

The prefractionation section between the cracking heaters/TLX'es and cracked gas compression encompasses for naphtha/gas oil crackers a primary fractionator/water quench tower system. The primary fractionator/water quench tower system has basically three objectives (1) prefractionation of residue, middle oil and heavy gasoline, (2) economical heat recovery and (3) quenching with a minimum pressure drop.

The prefractionation section, especially the primary fractionator, is a critical section with regard to feedstock flexibility.

When operating on AGO, the average TLX outlet temperature is higher than on naphtha. Therefore, the heat to be recovered over the primary fractionator is considerably larger than for naphtha. Due to the difference in fuel oil yield and characteristics between naphtha and AGO, the primary fractionator can operate on AGO at a considerable higher bottom temperature. The higher bottom temperature allows the quench oil cycle rate to be kept at a reasonable level. Resulting the difference in fuel oil yield, the circulation time is lower for AGO than for naphtha.

The scheme, as shown in fig. 5, is feasible for AGO/naphtha feedstock flexibility, but not for VGO (vacuum gas oil).

As outlined before, no TLX'es can be applied on VGO-cracking and direct oil quenching is mandatory directly at the cracking coil outlet. In order to reduce the quench oil rate on VGO and optimize the waste heat recovery, oil quenching can best be performed in three stages (fig. 6). The prefractionation section of an ethane/propane cracker is much simpler than the prefractionation of a naphtha/gas oil cracker. As there is hardly any fuel oil production, the primary fractionator can be deleted (fig. 4). It is also virtually not possible to use a prefractionation section designed for naphtha/gas oil only, for ethane/propane feedstock operation. An ethane/propane feedstock prefractionation cannot be used for naphtha/gas oil operation either.

When designing a cracker for partly liquid and partly gaseous feedstocks, it is possible to use the primary fractionator/quench tower system. However, the liquid feedstock throughput should be high enough to warrant a trouble-free primary fractionator operation. For a cracker with partly liquid/partly gaseous feedstocks, separate prefractionation trains for the liquid and gaseous feedstocks are attractive, except from the investment point of view.

Cracked Gas Compression

The cracked gas from the water quench tower has to be compressed from slightly more than atmospheric pressure to about 35 atmospheres by a multi-stage centrifugal compressor driven by a steam turbine. The selection of the number of stages depends on the cooling water temperature and diene content. At higher temperatures, the dienes contained in the cracked gas tend to polymerize and deposit on the compressor internals shortening the plant run length. Maximum allowable temperatures at the exhaust are 95 - 100 °C for naphtha/gas oil feedstocks and 110 - 115 °C for ethane/propane crackers.

Resulting the process gas temperature requirements as well as cooling water temperature, the number of stages is usually 4 or 5 for naphtha/gas oil crackers and 3 or 4 for ethane/propane crackers. Nowadays, with the increased utility cost, there is a tendency for the higher number of stages in order to achieve a higher efficiency at the expense of investment.

The effect of naphtha/AGO feedstock flexibility is marginal on the cracked gas compressor, because for constant ethylene/propylene product ratio the volumetric flow to the compressor is almost identical, whereas for the naphtha high severity operation the molucular weight is lower.

Table 2. Comparison cracked gas compressors

Feedstock	Ethane/propane	Naphtha/gas oil
Max. practical exhaust temp. °C	110 - 115	95 - 100
Number of stages	4 - 3	5 - 4

Recovery Section

From the material balances (Table 3) the similarity regarding C4's and lighter yields between AGO cracking and cracking of naphtha at medium severity is striking. Therefore, the recovery section down-stream the cracked gas compressor does not present particular problems with regard to liquid feedstock flexibility.

ETHYLENE PLANT FLOW SCHEME

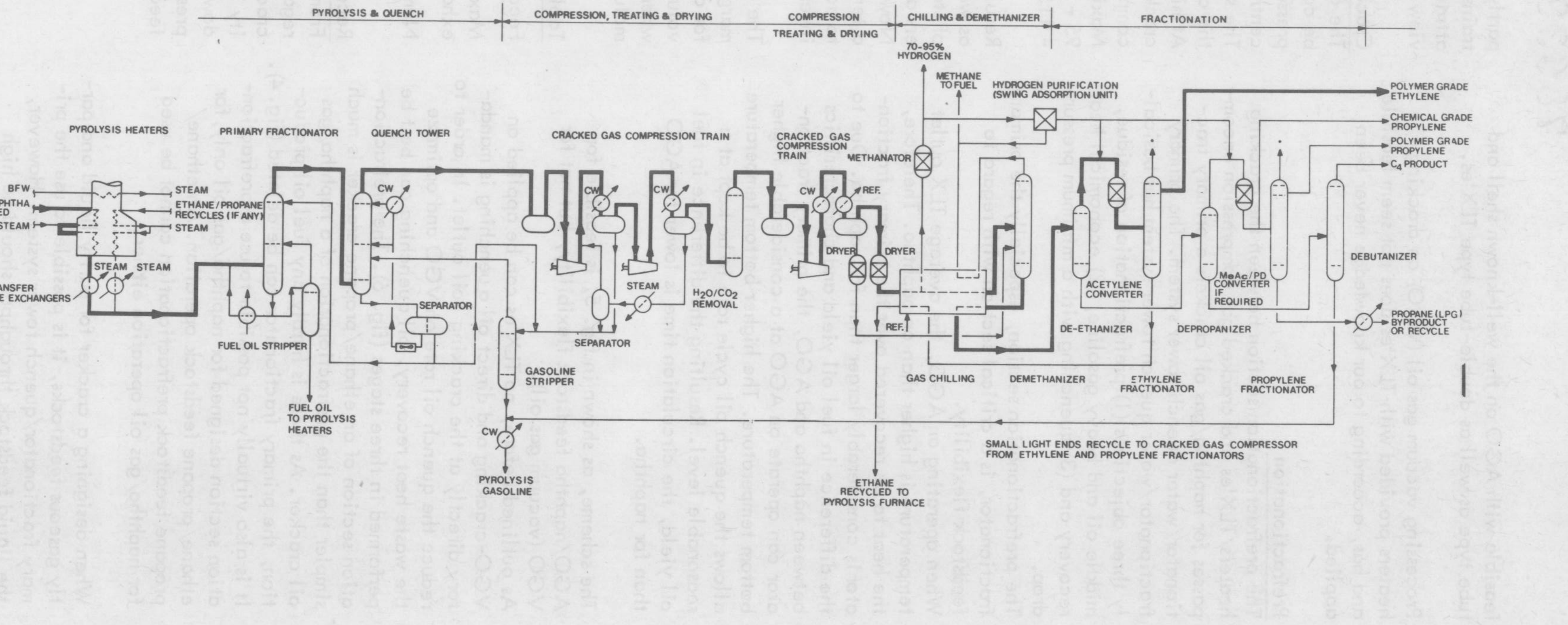

Table 3 Material Balances for different feedstocks and severity

Feedstock	Ethane	Propane	n/i Butane	Full range Naphtha High Sev.	Full range Naphtha Med. Sev.	Atm. Gas oil	Vac. Gas oil
Once-trhough ethylene yield, wt.%	48.56	34.45	30.75	28.70	25.50	23.0	18.0
Ultimate ethylene yield, wt.%	81.05	42.00	35.05	32.46	29.40	26.0	20.76
Feedstock rate, 1,000 metric ton/yr	555,210	1,071,400	1,283,800	1,386,300	1,530,600	1,730,000	2,167,630
Products, 1,000 t/yr							
Hydrogen-rich gas (90 mol.%)	61,000	29,800	27,070	26,850	26,720	26,100	25,540
Methane-rich gas	5,500	283,600	273,950	221,580	196,500	182,300	174,300
Ethylene, polymer grade	450,000	450,000	450,000	450,000	450,000	450,000	450,000
Propylene, polymer grade	9,090	179,700	201,770	169,800	252,250	252,250	296,600
Propane	1,300	-	11,150	9,800	13,000	13,200	13,650
Butadiene	11,855	28,200	38,450	64,600	75,700	84,770	118,500
Butylene/butane	5,010	15,400	110,030	55,400	89,400	88,230	137,500
Pyrolysis gasoline	10,455	79,200	145,780	325,150	381,130	321,750	411,700
Benzene	5,320	29,520	65,210	97,020	95,250	86,800	82,200
Toluene	940	7,800	18,950	48,510	61,200	62,300	65,100
C8-aromatics	120	1,100	5,800	29,250	36,300	42,200	54,200
Other pyrogasoline	4,075	40,080	55,820	60,370	188,380	130,450	210,200
Pyrolysis fuel oil	1,000	5,500	25,600	63,120	45,900	311,400	539,840

Basis : Recycle ethane cracking to extinction for all cases, recycle propane only for propane feedstock.

The analysis (Table 4) per section shows that the naphtha high severity operation is the governing design case for the demethanizer, hydrogen recovery and acetylene hydrogenation sections. For constant ethylene-to-propylene ratio the load on the de-ethanizer, C2-splitter, de-propanizer and C3-splitter (only when polymer grade propylene is required) is practically identical for naphtha and gas oil. The gas oil cracking case is the only governing design case in the recovery section for the debutanizer and even there is only a small difference with naphtha cracking at medium severity. Furthermore, gas oil cracking can be governing - dependent on the sulphur content - in the acid gas removal section.

As has been shown, there is regarding the recovery section only a small difference between naphtha and gas oil cracking, however, the picture is completely different for ethane/propane feedstock flexibility. With ethane/propane flexibility there are problems in the front end as well as in the back end of the recovery section. In the front end the completely different hydrogen - to - methane ratio as well as tail gas - to - ethylene ratio complicates the design. In the back end the difference in C3's production jeopardizes the feasibility of full ethane/propane flexibility.

Table 4 Governing Design Cases for Feedstock Flexibility

Basis : Ethane recycling and cracking to extinction.
Propane recycle only for propane feedstock.

Note : The maximum load for constant ethylene capacity is shown.

Feedstock	Ethane	Propane	Butanes	Naphtha High Sev.	Naphtha Medium Sev.	AGO	VGO
Load percentages on the columns :							
Prefractionator	-	-	88	93	100	115	175
Demethanizer	25	145	140	110	100	97	95
De-ethanizer	112	90	95	97	100	100	105
C_2-splitter	120	95	98	98	100	100	100
Depropanizer	-	105	82	70	100	100	115
C_3-splitter	-	70	80	68	100	100	115
Debutanizer	-	25	110	75	100	106	150

3.2 Revamping existing Naphtha Crackers for Naphtha/Gas Oil Flexibility

The bulk of existing European and Japanese cracking capacity is based on naphtha feedstock. In the context of this manuscript, it is worthwhile to consider the possibilities of revamping the existing crackers for naphtha/gas oil flexibility. It is obvious, that revamping existing crackers for naphtha/gas oil flexibility is not so straight forward as designing new naphtha/gas oil crackers.

Revamping existing naphtha cracking heaters for N/AGO flexibility is only occasionally feasible, usually will be chosen for new cracking heaters with feedstock flexibility. Potential pitfalls in existing heaters can be in the convection section the process- as well as the non-process (economizer and h.p. steam superheat) coils. Generally, it is easier to revamp a cracking section with central flue gas waste heat recovery, because then potential problems with regard to the non-process coils are eliminated.

The TLX'es of the naphtha cracker should be replaced by shorter TLX'es operating at lower residence times and higher outlet temperatures to avoid short run lengths. The h.p. steam pressure should be minimum 100 atmospheres to avoid quick fouling due to condensation.
The cracked gas compression, demethanizer, deethanizer, C2-splitter and depropanizer section do not present bottle-necks to achieve the same ethylene and propylene capacity on naphtha/gas oil as on naphtha only.
Due to the higher steam dilution and average TLX-outlet temperatures, the primary fractionator/water quench tower system will operate at a higher load and requires additional heat exchangers, pump capacity etc. The existing prefractionation design usually sets the maximum feasible gas oil cracking capacity.
Due to the higher C4-fraction yield, also the debutanizer will become a potential bottle-neck on

gas oil feedstock. Resulting the bottle-necks in the primary fractionator/water quench tower system as well as debutanizer the maximum feedstock flexibility will be limited to some 20 % gas oil/80 % naphtha for existing naphtha crackers.

3.3 What are the drawbacks of a cracker with feedstock flexibility?

It has been shown above that a cracker with feedstock flexibility from 100 % ethane up to 100 % gas oil is not feasible. In general will be opted for crackers with only dual feedstock flexibility i.e. naphtha/AGO crackers or ethane/propane crackers.

3.3.1 Naphtha/gas oil cracker

It has been outlined before in this paper, that feedstock flexibility does not affect the process yields or plant material balance. Regarding utilities consumption we can note that the plant's major utility, fuel, is virtually not affected. Obviously, h.p. steam generating capability of the TLX'es is worse per ton of naphtha for a naphtha/gas oil cracker than for a naphtha cracker. However, the disadvantage of smaller h.p. steam generating capability is partially offset by a larger heat recovery at lower temperature level by dilution steam generation.

The net h.p. (or m.p.) steam consumption is the balance of gross steam consumption by

1° the three main compressors, i.e. cracked gas, ethylene and propylene refrigeration;
2° the pumps in the hot section, and
3° balancing the dilution steam requirements,

and gross steam production by

1° the transfer line exchangers (h.p. steam), and
2° quench oil (dilution steam and l.p. steam).

The consumption by the three main compressors is identical because the design is set by the naphtha case. The steam or power consumption by the pumps increases because the pump capacity is considerably larger in a naphtha/AGO cracker than in a naphtha cracker. The shift of heat recovery by generating dilution steam in a naphtha/AGO cracker instead of h.p. steam generation in the TLX'es has the worst affect on the utilities. Alltogether, it has to be noted that the penalty on utilities by incorporating AGO cracking flexibility is relatively small and equals about a drop in the thermal efficiency in the cracking heaters of about 1 - 2 %. Incorporating naphtha/AGO/VGO feedstock flexibility has a much larger effect on the utilities. When there are no TLX'es (no h.p. steam generation), the heat has to be recovered at a lower temperature level (m.p. or dilution steam generation) (Table 6 and 7).

Potential pitfalls with regard to operability are mainly in the cracking heater convection section, quench coolers, primary fractionator and acid gas removal system. A well-known operator's fear is, that the run length of the cracking heaters/TLX on naphtha should be shorter in a naphtha/gas oil cracker than in a naphtha cracker. It has been shown that this is not the case. Furthermore, in case cracking heater convection section fouling occurs, there will be considerable downtime for mechanical cleaning, because provision for steam-air decoking of the convection section is usually not available.

Whereas there are some 8 - 10 cracking heaters in a large cracker and spare heater capacity is available, there is only one primary fractionator. Therefore, both units should run continuously between the scheduled bi-annual overhauls.

Comparing the BLCC of naphtha, naphtha/gas oil and gas oil crackers with identical ethylene and propylene production capacity, we can say :

a gas oil cracker requires more investment in the hot section (more cracking heaters, more high alloy piping, exchangers, bigger towers, pumps, etc.) acid gas removal and debutanizer section (higher throughput and higher C4-yield), but requires slightly less investment for the cracked gas compression, refrigeration demethanizer section and about the same investment for the de-ethanizer, C2-splitter, and depropanizer. Altogether, the BLCC of a gas oil cracker amounts to some 110 % of the BLCC of a naphtha cracker.

A naphtha/gas oil cracker obviously combines the maximum BLCC for a naphtha cracker and a gas oil cracker. For the same ethylene/propylene capacity the BLCC of N/AGO cracker is about 110 - 115 % of the naphtha cracker (Table 5).

Substantial savings (5 %) can be obtained on a N/AGO cracker when the number of cracking heaters is only capable of achieving design ethylene/propylene capacity on naphtha only.

Apart from the higher BLCC of a N/AGO cracker, also the capital expenditure for the off-sites like feed and product storage tanks are higher for a N/AGO cracker than for a naphtha cracker.

Table 5 Relationship between ethylene plant BLCC and feedstock

Feedstock	Relative BLCC (naphtha cracker high severity = 100)
Ethane	80
Propane	90
Naphtha	100
AGO	110
Naphtha/AGO	112
VGO	125

Table 6 Comparison Ethylene Production Cost for Naphtha Cracking in Naphtha/AGO Cracker and Naphtha only Cracker

All cost in U.S.$ per ton of ethylene.

Design	Naphtha Cracker	Naphtha/AGO Cracker
Feedstock	Naphtha	Naphtha
Feedstock cost	401	401
Byproducts value (credit)	(293)	(293)
Utilities	55	56
Capital charges	65	72
Operating charges	2	2
Net production cost, $/ton	230	238

Note : This table shows that ethylene production cost in naphtha/AGO crackers are higher than in naphtha only cracker, mainly due to increased capital charges.

Basis :	$/ton
Naphtha	130
Hydrogen	285
Fuel gas	90
Propylene	190
C4's	85
Butadiene	250
Gasoline	140
Fuel oil	70

Table 7 Aggregated Utilities for Different Feedstocks

Basis : 450,000 mta ethylene.
Recycle ethane cracking.
Maximum cooling water consumption. (Δt = 10 °C)
8,000 annual operating hours.

Feed	Ethane	Propane	n/i Butane	Naphtha	AGO	VGO
Fuel, MMKcal LHV/hr	260	310	330	330	370	440
H.P. Steam, ton/hr	120	80	75	70	75	170
Power, KWH	1500	2000	2500	3000	4000	5000
Cooling water, m3/hr	31000	31500	32000	32500	34000	41000