DEPARTMENT OF MINES AND ENERGY

SOUTH AUSTRALIA

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SOUTH AUSTRALIAN DEPARTMENT OF MINES AND ENERGY

AMDEL Report

No. 1473

UTILIZATION OF COOPER BASIN LPG FOR THE MANUFACTURE OF GASOLINE

by

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SUMMARY

Background

Large quantities of LPG will soon become available for domestic use in South Australia from the liquids fractionation plant currently under construction at Stony Point. One possible use which could be of distinct benefit to South Australia is the manufacture of high octane gasoline additives by an alkylation process. Alkylates can be produced from LPG by firstly converting propane and butane to propylene and butylene and then reacting these products with isobutane.

The South Australian Department of Mines and Energy commissioned AMDEL to carry out cost studies for converting some of the Cooper Basin LPG to high octane alkylates.

Objectives

The aims of this project were as follows:

1. To investigate the economic viability of producing high octane alkylates from LPG and blending these with condensate to produce premium grade gasoline.
2. To determine the capital and operating costs for a plant using butane only as the raw material and for a plant using both butane and propane.
3. To demonstrate the sensitivity of the cost of finished gasoline to changes in the price of raw materials.
4. To calculate the overall energy efficiency.

Summary of Work Done

Three stages are involved in the overall alkylation process, viz, isomerization, dehydrogenation and alkylation. Isomerization is used to convert n-butane to its isomer isobutane which is an essential reagent in the alkylation stage. Dehydrogenation was developed in the late 1970s to convert propane and butane to their equivalent mono-olefins (propylene and butylene). Although it is a relatively recent development, it is based on similar technology which has been in commercial use for many years. In the alkylation stage, isobutane is reacted with the mono-olefins.

Processing and economic details of the dehydrogenation process have been published by two of developers and licensors, UOP and Houdry Division of Air Products and Chemical Inc. Integrated flowsheets incorporating the dehydrogenation stage have also been presented. These were used to calculate mass balances, raw material and utility requirements, together with capital and operating costs for plants located at Stony Point and Port Stanvac.

A summary of the costs obtained is presented below:

Capital Costs

Case 1 - A plant to process 150,000 tonnes/a butane.

	Stony Point	Port Stanvac
On-sites	$\$54.1 \times 10^6$	$\$46.7 \times 10^6$
Off-sites	$\$18.0 \times 10^6$	$\$15.5 \times 10^6$
Total	$\$72.1 \times 10^6$	$\$62.2 \times 10^6$

Case 2 - A plant to process 200,000 tonnes/a LPG consisting of 130,800 tonnes/a butane and 69,200 tonnes/a propane.

	Stony Point	Port Stanvac
On-sites	$\$80.6 \times 10^6$	$\$69.6 \times 10^6$
Off-sites	$\$21.4 \times 10^6$	$\$18.5 \times 10^6$
	$\$102.0 \times 10^6$	$\$88.1 \times 10^6$

Production Costs (Stony Point)

	Costs, \$/a	
	Case 1	Case 2
LPG feedstock	33,000,000	44,000,000
Net manufacturing costs	22,445,000	27,046,000
Production cost	55,445,000	71,046,000
Condensate	10,725,000	13,600,000
Gasoline	66,170,000	84,646,000
Gasoline, \$/tonne	365	368
Gasoline, ¢/litre	27.0	27.2

The required wholesale price in Adelaide would be made up as follows:

	Costs, ¢/litre			
	Case 1		Case 2	
	Stony Point	Port Stanvac	Stony Point	Port Stanvac
Gasoline - refinery gate	27.0	28.0	27.2	28.3
Return on capital	4.4	3.8	4.9	4.3
Transport to Adelaide	1.5	-	1.5	-
Government taxes	6.5	6.5	6.5	6.5
	39.4	38.3	40.1	39.1

Graphs were constructed to illustrate the sensitivity of this price to changes in the price of LPG over the range $180 to $280/tonne and the price of condensate over the range $210 to $310/tonne. The cost of raw materials account for around 70% of the refinery gate cost of gasoline.

Conclusions

(1) The production of gasoline from LPG appears to be marginally viable. Economic viability depends on the cost of raw materials, applicability of Government taxes, and the wholesale price of gasoline in Adelaide.

(2) Total capital costs for the two processes considered are:

Stony Point

(a) $72.1 million to treat 150,000 tonnes/a butane

(b) $102 million to treat 200,000 tonnes/a butane plus propane.

Port Stanvac

(a) $66.2 million to treat 150,000 tonnes/a butane

(b) $88.1 million to treat 200,000 tonnes/a butane plus propane.

(3) The cost of gasoline delivered to Adelaide, including 15% p.a. profit and assuming feedstock costs of $220/tonne for LPG and $250/tonne for condensate are:

(a) Stony Point - 39.4 and 40.1¢/litre respectively for the two processes.

(b) Port Stanvac - 38.3 and 39.1¢/litre respectively.

(4) The required wholesale price in Adelaide to return 15% p.a. on capital is about 40¢/litre if Government taxes apply.

Recommendations

(1) Firm production schedules should be obtained to ensure that there will be adequate supplies of butane for at least 10 years.

(2) Negotiations should be commenced with Federal and State Governments to seek exemption from the relevant taxes.

(3) Full feasibility studies should be carried out with direct input from the licensors of the processes.

1. INTRODUCTION

The Cooper Basin, situated in the north-eastern corner of South Australia and extending into Queensland, contains hydrocarbon gases and liquids ranging in composition from methane to crude oil. Production of natural gas has so far been from relatively 'dry' wells, and the small amounts of hydrocarbon liquids produced have been flared off. However, it will soon become necessary to utilise the 'wet' wells, and a liquids pipeline has recently been constructed from Moomba to Stony Point to cater for this stage of production. Once the pipeline and fractionating plant have been commissioned, LPG from the Cooper Basin will be available for domestic use and export.

Catalytic dehydrogenation processes have recently been developed in the US for processing saturated light hydrocarbons such a propane and butane to the corresponding mono-olefins (propylene and butylene). These can then be alkylated by reacting them with isobutane to form high octane alkylates which can be used to replace tetra ethyl lead in gasolines. The isobutane is either available from the LPG feedstock or is produced from n-butane by isomerization. It is assumed that there would be sufficient flexibility in the distillation and alkylation processes to ensure that the quality of the alkylates and condensate would be such that the resulting gasoline would meet all the normal quality specifications.

The South Australian Department of Mines and Energy (SADME) has suggested that if Cooper Basin LPG were converted to motor alkylate, South Australian motorists could benefit from lower cost gasoline and security of supply from a local source. Accordingly, AMDEL was commissioned to carry out cost studies to determine the capital and operating costs for converting some of the Cooper Basin LPG to high octane alkylates at two plant sites, viz. Stony Point and Port Stanvac.

Two alternatives were to be investigated:

1. Processing of 150,000 tonnes/a butane.
2. Processing of 200,000 tonnes/a butane plus propane.

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2. COOPER BASIN LPG

LPG recovered from the Cooper Basin liquids is expected to have the following composition:

	%
Propane	61
Iso-Butane	13
n-Butane	26
	100

Butane production is expected to be of the order of 150,000 tonnes/a, and this figure is assumed for the cost studies detailed in this Report. The wholesale price of propane and butane in South Australia as from 1 July 1982, was $220. This price is used in this Report and the effect of price variations is also examined.

3. REFINERY FLOWSHEETS

3.1 Introduction

Conversion of LPG to alkylate involves three major processes, viz., dehydrogentation, isomerization and alkylation. The last two are well established commercial processes, but the former was developed in the late 1970s when US oil producers began to examine ways of using surplus LPG to supplement gasoline supplies. At least three companies, viz. Phillips Petroleum, UOP, and the Houdry Division of Air Products and Chemical Inc., have developed dehydrogenation processes to convert LPG to propylene and butylene.

Little is known of the Phillips process except for a recent announcement (Anon, 1982) which stated that the process had been tested on a semi-commercial scale and was ready for licensing on a commercial scale. The process was isobutane feedstock to produce isobutylene for alkylation. It has been estimated that the capital cost of a dehydrogenation plant to treat 15,000 bpd isobutane feed to obtain 6,100 bpd isobutylene product would be $US41 m, US Gulf Coast, first quarter 1982 costs.

The other two processes have been described in the literature in more detail and are outlined in the following sections.

3.2 UOP Dehydrogenation Process

The UOP process was described by Berg, Vora, and Mowry (1980) and is illustrated diagrammatically in Fig. 1. The fresh feed, either propane or butane, is mixed with recycled hydrocarbons and hydrogen, heated to the reactor inlet temperature of 600°C to 700°C, and reacted in a series of stacked reactors. The reactors use the moving bed continuous catalyst regeneration concept which has been used successfully in the UOP Platforming process. The reaction is endothermic and activity is maintained by supplying the heat of reaction through interheaters.

The product from the reactors passes through a series of heat exchangers, compression and turbo-expansion to separate the relevant C_3 or C_4 hydrocarbons from a combination of recycle gas and net hydrocarbon off gas. The liquid from the separator is pumped to a light end stripper. The bottoms contain the product olefin and any unconverted feed.

Each pass through the process achieves about 40% conversion and hence there is a large recirculating load. However, catalytic dehydrogenation is a highly selective process giving high percentage yields of the desired products. This contrasts with other dehydrogenation processes such as steam cracking which give a wide scatter of products. UOP figures comparing the yields obtained from catalytic dehydrogenation and thermal cracking from propane, isobutane and n-butane feedstocks are listed in Table 1.

3.3 Houdry Catofin Process

The Houdry catalytic dehydrogenation process, known as Catofin, was described by Craig and White (1980) and Gussow, Spence and White (1980) and is illustrated diagrammatically in Fig. 2. The process was based on the well established Catadiene process, described by Craig and Dufallo (1979), which is another dehydrogenation process used to manufacture butadiene.

The Catofin process is very similar to the UOP process except in the reactor design. In place of the UOP moving bed concept, Houdry used multiple reactors to allow for three different phases of operation, viz, on stream, on purge and on catalyst regeneration.

3.4 Integrated Flowsheets

High octane alkylate is produced by reacting isobutane with propylene and/or butylene. Hence the initial feedstock can be butane only or butane plus propane in appropriate proportions, but not propane on its own. UOP have presented flowsheets with overall mass balances for the two configurations of direct interest, viz.

1. alkylate from mixed butanes
2. alkylate from butanes and propane.

They have also presented capital and operating cost information for the first flowsheet.

On the other hand, Houdry have presented cost information on a number of related flowsheets but not on the actual flowsheets required in this Report. Cost information is available for the following flowsheets:

1. alkylate from n-butane only
2. alkylate plus propylene from n-butane and propane
3. propylene from propane
4. isobutylene from n-butane.

In general, UOP flowsheets and costs have been used in this Report because they refer to the actual flowsheets required. However, cost information for the flowsheet involving both propane and butane has been deduced on a pro-rata basis from the Houdry information and from articles giving cost and flowsheet details on alkylation processes (Anon, 1980 - Refining Process Handbook and Ewing 1971) and isomerization (Ewing, 1971).

3.4.1 Mixed Butane Feed

A block diagram of the process for producing alkylate from mixed butanes is presented in Fig. 3. The approximate mass balance data included on the diagram have been calculated from the UOP and Houdry information and from the alkylation and isomerization references mentioned above. The mixed butanes are charged to a deisobutanizer column where the isobutane is taken off overhead and the n-butane is withdrawn from the column and fed to an isomerization unit. Any C_5+ components in the butane feed are removed from the base of the column.

About 60% of the n-butane is isomerized in a single pass through the isomerization unit to provide additional isobutane for the alkylation reaction. The deisobutanizer overhead and the isomerization product are combined and fed to the dehydrogenation unit where about 40% of both the normal and isobutanes are converted to butylene. The mixture of butylene and unreacted butanes is fed to the alkylation unit where the butylene and isobutane react, while the n-butane passes unchanged through the process and after fractionation is returned to the deisobutanizer.

An overall mass balance for the process is given in Table 2 together with the properties of the alkylate produced. These have been taken directly from UOP data. A more detailed mass balance has been calculated for plant sizing and cost estimating purposes, and this is shown in Fig. 3. A feedrate of 150,000 tonnes/a mixed butanes would result in the production of 138,500 tonnes/a high octane alkylate.

3.4.2 Butane and Propane Feed

A block diagram of the process for producing alkylate from propane and mixed butanes is presented in Fig. 4. In order to utilize the maximum amount of propane, the mono-olefin is derived entirely from the propane, while butane is used solely as a source of isobutane. The mixed butanes are fed to a deisobutanizer column as before, but in this instance the isomerization unit is in closed circuit with the deisobutanizer and hence only isobutane is used in later processing stages.

The propane is fed to the dehydrogenation unit and the product is reacted with isobutane in the alkylation unit. Unreacted propane (about 60% of the total feed) passes through the alkylation unit unchanged and after fractionation is recycled to dehydrogenation stage.

In both alkylate processes, hydrogen and light ends are produced. Some hydrogen is used in the process while the remainder and the light ends are available for sale or for use as fuel or chemicals elsewhere in the refinery complex.

An overall mass balance for the process is given in Table 3 together with the properties of the alkylate produced. A more detailed mass balance has been calculated and the data appear in Fig. 4. A total feedrate of 200,000 tonnes/a would comprise 130,800 tonnes/a butane and 69,200 tonnes/a propane. This would result in the production of 175,600 tonnes/a high octane alkylate.

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4. COST ESTIMATES

4.1 Capital Costs

Capital costs have been presented in both the UOP and Houdry papers in US currency referring to second quarter 1980, US Gulf Coast. They have been converted to late 1982 costs in Australian currency for plant sites at Stony Point and Port Stanvac using the following assumptions:

(i) One US dollar equals one Australian dollar.

(ii) Capital costs in Adelaide are 20% higher than US Gulf Coast costs, and at Stony Point the costs are 16% higher than in Adelaide.

i.e. Stony Point costs = 1.39 times US Gulf Coast costs.

These ratios are based on experience at the Port Stanvac refinery and the Stony Point liquids plant.

(iii) The Nelson Cost Indexes (Farrar, 1982) have been used to update costs. They are available only to the end of the first quarter 1982 and have been extrapolated to the end of 1982. December 1982 costs = 1.24 × mid 1980 costs.

The resulting factor to update published US Gulf Coast figures in 1980 to Stony Point costs in late 1982 is 1.72.

All figures quoted in the following sections are for 1982 at Stony Point.

4.1.1 Mixed Butane Feed

The capital cost for a UOP plant to treat butane only is $\$104.7 \times 10^6$ for a butane feed rate of 13,430 bpsd, or 408,600 tonnes/a, assuming 330 stream days per year. Exponential factors for cost versus size for many types of processing plant were published by Guthrie (1970). These included:

Alkylation	0.6
Isomerization	0.65
Dehydrogenation (Catadiene)	0.68

Houdry figures indicate that the costs of these units in a butane alkylation plant are in the ratio 2:1:7. A weighted average of the exponential factors is 0.66 for the overall process, and this factor has been used to calculate the costs of different sized plants.

Hence, the capital cost of a plant to treat 150,000 tonnes/a butane is $54.1 × 10^6 at Stony Point or $46.7 × 10^6 at Port Stanvac. However, this cost refers to battery limits only. The cost of off-sites, including storage for feed and products, would be a further $18.0 × 10^6 at Stony Point or $15.5 × 10^6 at Port Stanvac based on figures published by Houdry. This represents total investments of $72.1 × 10^6 and $62.2 × 10^6 respectively for the two plant sites.

4.1.2 Butane and Propane Feed

No capital cost has been given directly for a plant to produce alkylate from both butane and propane. However, the costs can be deduced from throughputs and other cost data given by UOP and Houdry.

Alkylation

The alkylation unit produces 175,600 tonnes/a compared with 135,800 tonnes/a in the previous flowsheet. Using an exponential factor of 0.6, the capital cost should be increased by a factor of 1.15.

Isomerization

The isomerization unit treats a total feedrate of 133,000 tonnes/a compared with 124,000 tonnes/a in the previous flowsheet. Using an exponential factor of 0.65, the capital cost should be increased by a factor of 1.05.

Dehydrogenation

The propane dehydrogenation unit receives a total feedrate of 173,000 tonnes/a compared with a feedrate of 168,000 tonnes/a to the butane unit. Using an exponential factor of 0.68, the capital cost should be increased by a factor of 1.02. However, figures published by Houdry show that the capital cost of a propane dehydrogenation unit is 1.62 times the cost of a butane dehydrogenation unit for plants of equal capacity. Therefore, the capital cost of the propane unit in Fig. 4 is 1.65 times that of the butane unit in Fig. 3.

As mentioned previously, the ratio of the capital costs of alkylation, isomerization and dehydrogenation units for the butane flowsheet was 2:1:7. From the factors given above, the corresponding figures for the butane plus propane plant are 2.30:1.05:11.55. This represents a total cost of 14.9 units compared with 10.0 for the butane flowsheet.

The total battery limits cost of the butane plus propane plant is therefore 1.49 times the cost of the butane plant, i.e. $80.6 × 10^6 for the Stony Point and $69.6 × 10^6 for Port Stanvac. If the cost of off-sites increases in relation to the total feed and product rate, then the costs of off-sites are 21.4×10^6 and 18.5×10^6 respectively, and the total investments are 102.0×10^6 and 88.1×10^6 respectively.

4.2 Production Costs

Production costs for alkylate at Stony Point have been calculated in Appendix A and are summarised below:

	Butane only	Butane plus Propane
Net production costs, $/a	55,445,000	71,046,000
Net costs, $/tonne alkylate	400	405

Production costs at Port Stanvac would differ only in those items related to capital cost such as maintenance, depreciation etc. and in the cost of transporting the LPG from Stony Point to Port Stanvac (say $20/tonne). It should be noted that a credit has been claimed for the sale of excess hydrogen. It is assumed that this would be used elsewhere in the refinery complex at either site. The credit would not necessarily be valid if the plant were located at another metropolitan site such as Port Adelaide. The Port Stanvac production costs are summarised below:

	Butane only	Butane plus Propane
Net production costs, $/a	57,107,000	73,162,000
Net costs, $/tonne alkylate	412	417

The alkylate would be blended with Cooper Basin condensate (RON ≈80) to form gasoline with an octaine rating of 97. The proportions of alkylate and condensate used in blending would be as follows:

	Alkylate tonnes/a	Condensate tonnes/a	Gasoline tonnes/a
Butane only	138,500	42,900	181,400
Butane plus propane	175,600	54,400	230,000

Using the alkylate costs given above and a condensate cost of $250/tonne at Stony Point and $270/tonne at Port Stanvac (including $20/tonne transport from Stony Point), the refinery gate cost of gasoline would be as follows:

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	Butane only		Butane plus Propane	
	Stony Point	Port Stanvac	Stony Point	Port Stanvac
Alkylate, \$/a	55,445,000	57,107,000	71,046,000	73,162,000
Condensate, \$/a	10,725,000	11,583,000	13,600,000	14,688,000
Gasoline, \$/a	66,170,000	68,690,000	84,646,000	87,850,000
Gasoline, \$/tonne	365	379	368	382
Gasoline, ¢/litre (Sp.Gr. = 0.74)	27.0	28.0	27.2	28.3

Most of the State's gasoline is sold in Adelaide where the break-even selling price must also include transport from Stony Point (say \$20/tonne or 1.5¢/litre). It is assumed that the Federal Tax of 5.155¢/litre and the State Government franchise fee of 1.33¢/litre will also apply. These fees apply to gasoline but not to LPG. Since the gasoline considered here is manufactured mainly from LPG, there may be a case for seeking partial or total exemption from these fees.

The operating company would require a return on capital of about 15% p.a. simple interest, which is equivalent to 4.4 and 4.9¢/litre respectively for the two processing routes at Stony Point, and 3.8 and 4.3¢/litre respectively at Port Stanvac. The final wholesale price in Adelaide would then be made up as follows:

	Butane only		Butane plus Propane	
	Stony Point	Port Stanvac	Stony Point	Port Stanvac
Refinery gate cost, ¢/litre	27.0	28.0	27.2	28.3
Return on capital	4.4	3.8	4.9	4.3
Transport to Adelaide	1.5	-	1.5	-
	32.9	31.8	33.6	32.6
Fees and taxes (if applicable)	6.5	6.5	6.5	6.5
Total	39.4	38.3	40.1	39.1

The present retail price in Adelaide is 42.4¢/litre. When the retailer's margin of 3.6¢/litre is deducted, this leaves a wholesale price (including Government taxes) of 38.8¢/litre. This is only marginally less than the selling prices derived in the tabulation above.

4.3 Cost Sensitivity Analysis

Two of the major items in the total cost of gasoline manufactured by the processes described in this Report are the costs of LPG feedstock and condensate blending stock. The cost of LPG represents about 60% of the total production cost of alkylate, and hence the combined costs of LPG and condensate account for almost 70% of the total refinery gate cost of gasoline. The effect of variations in these raw materials costs for Stony Point is summarised in Tables 4 and 5, and shown graphically in Figs 5 and 6. The gasoline costs referred to are the total wholesale prices in Adelaide including Government fees and taxes. For Port Stanvac, the gasoline prices would be 1¢/litre cheaper than those shown in the graphs for Stony Point.

A change of $20/tonne in the price of LPG products a change of 1.22 and 1.29¢/litre respectively in the prices of gasoline in the two processes. A change of $20/tonne in the price of condensate produces a change of 0.35¢/litre in the price of gasoline in either process.

4.4 Energy Efficiency

The overall energy efficiency of the processes has been determined by comparing the gross specific energy of the gasoline produced with that of the raw materials. The gross specific energy values used are as follows:

Hydrocarbon	Gross Specific Energy (GJ/tonne)
Propane	50.0
Butane	49.5
Condensate	48.9
Gasoline	46.5

Case 1 - Butane only

		GJ/a
Raw Materials:		
Butane = 150,000 tonnes/a @ 49.5 GJ/tonne	=	7,425,000
Condensate = 42,900 tonnes/a @ 48.9 GJ/tonne	=	2,097,810
Total		9,522,810
Product: 181,400 tonnes/a @ 46.5 GJ/tonne	=	8,435,100
Overall Energy Efficiency	=	88.6%

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Case 2 - Butane plus Propane

			GJ/a
Raw Materials:			
Propane	- 69,200 tonnes/a @ 50.0 GJ/tonne	=	3,460,000
Butane	- 130,800 tonnes/a @ 49.5 GJ/tonne	=	6,474,600
Condensate	- 54,400 tonnes/a @ 48.9 GJ/tonne	=	2,660,160
	Total		12,594,760
Product:	230,000 tonnes/a @ 46.5 GJ/tonne	=	10,695,000
Overall Energy Efficiency		=	84.9%

Energy efficiency can also be calculated on a number of other bases as follows:

(1) Excess hydrogen is produced and is available for use as a fuel. The energy efficiencies based on total products and total feed are 91.0% and 86.7% respectively for Cases 1 and 2.

(2) The processes also consume electrical power and hydrocarbon fuels. Energy efficiencies based on total products and total energy input (fuel, power and raw materials) are 76.6% and 76.5% respectively.

5. DISCUSSION

Although the technology used in the dehydrogenation unit is relatively new, it is based on commercial processes and equipment which have been used for decades. The other stages involved in the overall flowsheets are established commercial processes. As a result, no really new or experimental technology would be required in an alkylation plant.

Capital costs and operating requirements have been deduced from the overall plant figures presented in the literature by UOP and Houdry. Their figures for most of the units are based on commercial operating experience, while those for dehydrogenation are engineering estimates based on long experience with similar equipment. In general, it appeared that the UOP plant would be slightly cheaper and less expensive to operate, but the two sets of data were presented on different bases and were not directly comparable. For the costs presented in this Report, it was necessary to use figures from both sets of data, and hence the net result is an 'average' of the two. More definite figures could only be obtained by commissioning UOP and Houdry to carry out detailed studies and provide their own set of costs.

The present cost study indicates that the production of gasoline from LPG could be economically attractive. A final assessment of its viability will depend on three major factors:

(1) The total cost of raw materials (both LPG and condensate) represents about 70% of the total refinery gate cost of the gasoline and hence has a major influence on total costs.

(2) Federal and State taxes on gasoline, amounting in total to 6.5¢/litre, could make the difference between economic viability or otherwise. Exemption from these taxes should be sought on the basis that the gasoline is manufactured from LPG which itself is exempt.

(3) The current wholesale and retail prices of gasoline in Adelaide at any time will naturally determine the final profitability of the project.

The costing has assumed that 150,000 tonnes/a butane will be available as feedstock on a long term basis. This is approximately equal to the predicted maximum production rate of butane, and if this rate can be maintained for only a few years and then decreases, the economic viability of the project will suffer accordingly. A straight line depreciation of 10% p.a. has been used, and this implies the assumption that the plant will operate <u>at full production</u> for a minimum of ten years.

6. CONCLUSIONS

(1) The production of gasoline from LPG appears to be marginally viable. Economic viability depends on the cost of raw materials, applicability of Government taxes, and the wholesale price of gasoline in Adelaide.

(2) Total capital costs for the two processes considered are:

Stony Point

(a) \$72.1 million to treat 150,000 tonnes/a butane

(b) \$102 million to treat 200,000 tonnes/a butane plus propane.

Port Stanvac

(a) \$66.2 million to treat 150,000 tonnes/a butane

(b) \$88.1 million to treat 200,000 tonnes/a butane plus propane.

(3) The cost of gasoline delivered to Adelaide, including 15% p.a. profit and assuming feedstock costs of \$220/tonne for LPG and \$250/tonne for condensate are:

(a) Stony Point - 39.4 and 40.1¢/litre respectively for the two processes

(b) Port Stanvac - 38.3 and 39.1¢/litre respectively.

(4) The required wholesale price in Adelaide to return 15% p.a. on capital is about 40¢/litre if Government taxes apply.

7. RECOMMENDATIONS

(1) Firm production schedules should be obtained to ensure that there will be adequate supplies of butane for at least 10 years.

(2) Negotiations should be commenced with Federal and State Governments to seek exemption from the relevant taxes.

(3) Full feasibility studies should be carried out with direct input from the licensors of the processes.

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APPENDIX A

ALKYLATE PRODUCTION COSTS

A1. PROCESS REQUIREMENTS AND PRODUCTS

A1.1 Raw Materials

Butane only:	150,000 tonnes/a butane
Butane plus propane:	130,800 tonnes/a butane
	69,200 tonnes/a propane
	200,000 tonnes/a total LPG feed

A1.2 Utilities

Process	Power kWh/a	Net Fuel GJ/a	Boiler Feed Water kl/a	Make-up Water kl/a
A. Butane feed only				
Isomerization	1.26×10^6	113,000	-	34,000
Dehydrogenation	4.00×10^6	639,000	34,000	607,000
Alkylation	7.71×10^6	992,000	23,000	1,050,000
Total	12.97×10^6	1,744,000	57,000	1,691,000
B. Butane plus propane				
Isomerization	1.35×10^6	121,000	-	36,000
Dehydrogenation	9.61×10^6	238,000	43,000	709,000
Alkylation	9.78×10^6	1,258,000	29,000	1,331,000
Total	20.74×10^6	1,617,000	72,000	2,076,000

A1.3 Products

	Butane only	Butane plus Propane
Alkylate, tonnes/a	138,500	175,600
Hydrogen: Gross, tonnes/a	2,100	2,100
Reused, tonnes/a	500	500
Net, tonnes/a	1,600	1,600
Light ends*, gross, tonnes/a	9,400	22,300

*All of the light ends are reused in the process as fuel. This has been allowed for in calculating the net fuel figure listed in Section 1.2 "Utilities".

A2. TOTAL PRODUCTION COSTS (STONY POINT)

The total production costs for alkylate at Stony Point are summarised below:

	Butane only $'000/annum	Butane plus Propane $'000/annum
1. Raw Materials		
LPG @ $220/tonne	33,000	44,000
2. Utilities		
Power @ 5¢/kWh	649	1,037
Fuel (propane) @ $4.90/GJ	8,546	7,923
Boiler feed water @ $1.30/kl	74	94
Make-up water @ 40¢/kl	676	830
Total utilities	9,945	9,884
3. Direct Labour		
5 men/shift @ $28,000/a	560	560
4. Maintenance		
3% of fixed capital	2,163	3,060
5. Supervision		
20% of direct labour	112	112
6. Operating Supplies		
Chemicals, catalysts, etc.	733	755
7. DIRECT MANUFACTURING COSTS (Items 1 to 6)	46,513	58,371
8. Payroll Overhead		
(20% of direct labour)	112	112
9. Plant Overhead		
(125% of direct labour)	700	700
10. Process Control		
(25% of direct labour)	140	140
11. INDIRECT MANUFACTURING COSTS (Items 8 to 10)	952	952
12. Depreciation		
(10% of fixed capital)	7,210	10,200
13. Property Taxes and Insurance		
(1% of fixed capital)	721	1,020

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		Butane only $'000/annum	Butane plus Propane $'000/annum
14.	FIXED MANUFACTURING COSTS (Items 12 and 13)	[illegible]	11,220
15.	TOTAL MANUFACTURING COSTS (Items 7, 11 and 15)	55,396	70,543
16.	Administrative Expenses (3% of manufacturing costs)	1,662	2,116
17.	TOTAL PRODUCTION COSTS	57,058	72,659
18.	Sale of Hydrogen by-product @ 9¢/Nm^3	1,613	1,613
19.	NET PRODUCTION COSTS	55,445	71,046
20.	Net Cost, $/tonne Alkylate	400	405

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TABLES 1 to 5

FIGS 1 to 6

TABLE 1: ESTIMATED YIELDS - CATALYTIC DEHYDROGENATION VERSUS THERMAL CRACKING

	Weight/100 Weights of Feed Converted	
	Catalytic Dehydrogenation	Thermal Cracking
A. Propane		
Hydrogen	3	1.5
Methane	7	23
Ethylene	3	36
Ethane	10	3
Propylene	77	33
Other	0.1	3.5
B. Isobutane		
Hydrogen	3	2
Propylene	3	30
Isobutylene	80	42
Normal Butylenes	9	1
Butadiene	0.3	1
Normal Butane	1.7	1
Other	3	23
C. n-Butane		
Hydrogen	3	2
Ethylene	2	35
Propylene	2	28
Butylenes	81	20
Butadienes	2	4
Isobutane	2	-
Others	8	11

TABLE 2: PRODUCTION OF ALKYLATE FROM MIXED BUTANES

	Weight/100 Weights of Alkylate	Weight tonnes/a
Feed:		
Mixed butanes	108.3	150,000
Products:		
Motor alkylate	100.0	138,500
Hydrogen plus light ends	8.3	11,500
	108.3	150,000
Typical alkylate properties:		
Specific gravity @ 60°F	0.7405	
Research clear octane No.	95.0	
Motor octane No.	93.0	
RON + 3cc TEL	106.4	

TABLE 3: PRODUCTION OF ALKYLATE FROM MIXED BUTANES AND PROPANE

	Weight/100 Weights of Alkylate	Weight tonnes/a
Feed:		
Mixed butanes	74.5	130,800
Propane	39.5	69,200
	113.9	200,000
Products:		
Motor alkylate	100.0	175,600
Hydrogen plus light ends	13.9	24,400
	113.9	200,000
Typical alkylate properties:		
Specific gravity @ 60°F	0.695	
Research clear octane No.	91.6	
Motor octane No.	90.7	
RON + 3cc TEL	103.2	

TABLE 4: GASOLINE PRICES VERSUS RAW MATERIAL COSTS - BUTANE ONLY

LPG, $/tonne	Gasoline Prices*, ¢/litre					
	180	200	220	240	260	280
Condensate, $/tonne						
210	36.3	37.5	38.7	39.9	41.1	42.4
230	36.6	37.8	39.1	40.3	41.5	42.7
250	37.0	38.2	39.4	40.6	41.8	43.1
270	37.3	38.5	39.8	41.0	42.2	43.4
290	37.7	38.9	40.1	41.3	42.5	43.8
310	38.0	39.2	40.5	41.7	42.9	44.1

*Price includes return on investment, transport to Adelaide and Government taxes.

TABLE 5: GASOLINE PRICES VERSUS RAW MATERIAL COSTS - BUTANE PLUS PROPANE

	Gasoline Prices*, ¢/litre					
LPG, $/tonne	180	200	220	240	260	280
Condensate, $/tonne						
210	36.8	38.1	39.4	40.7	42.0	43.3
230	37.2	38.5	39.8	41.0	42.3	43.6
250	37.5	38.8	40.1	41.4	42.7	44.0
270	37.9	39.2	40.5	41.7	43.0	44.3
290	38.2	39.5	40.8	42.1	43.4	44.7
310	38.6	39.9	41.2	42.4	43.7	45.0

*Price includes return on investment, transport to Adelaide and Government taxes.

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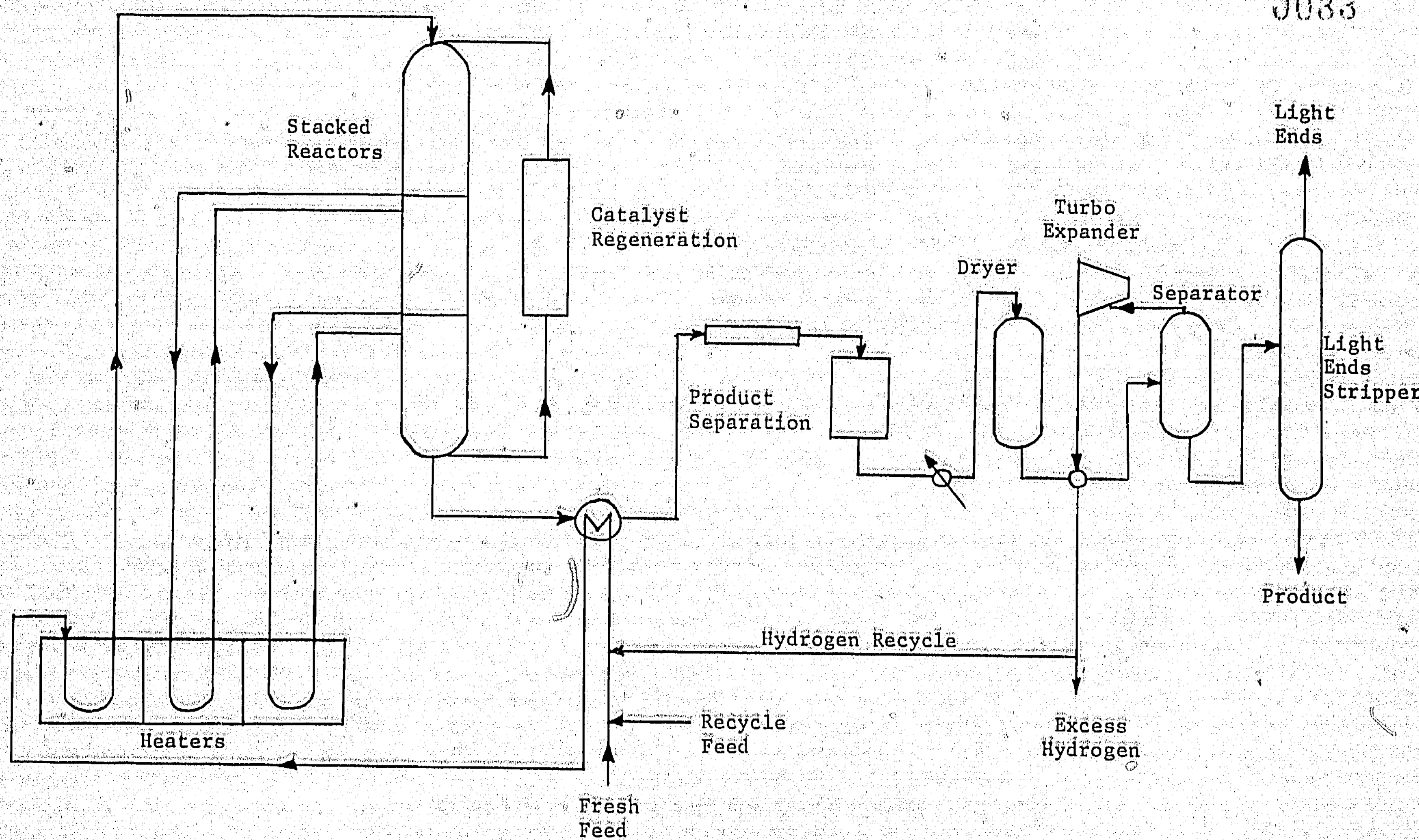

FIG. 1: UOP CATALYTIC DEHYDROGENATION PROCESS

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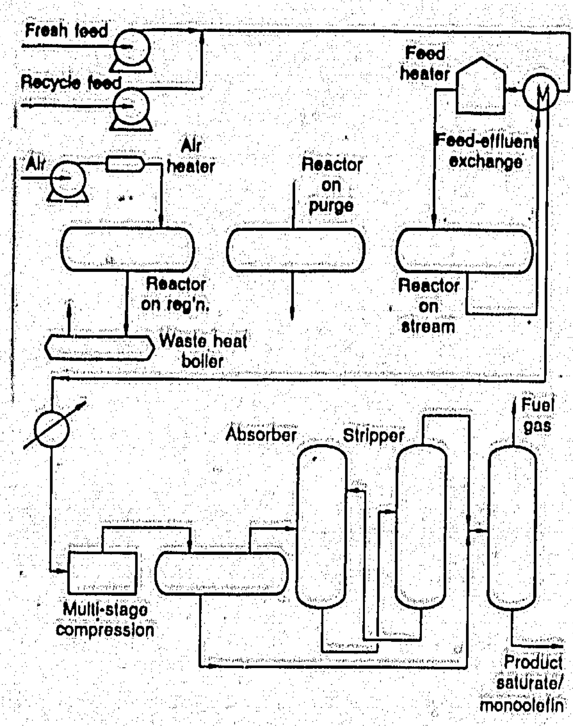

FIG. 2: HOUDRY CATOFIN PROCESS

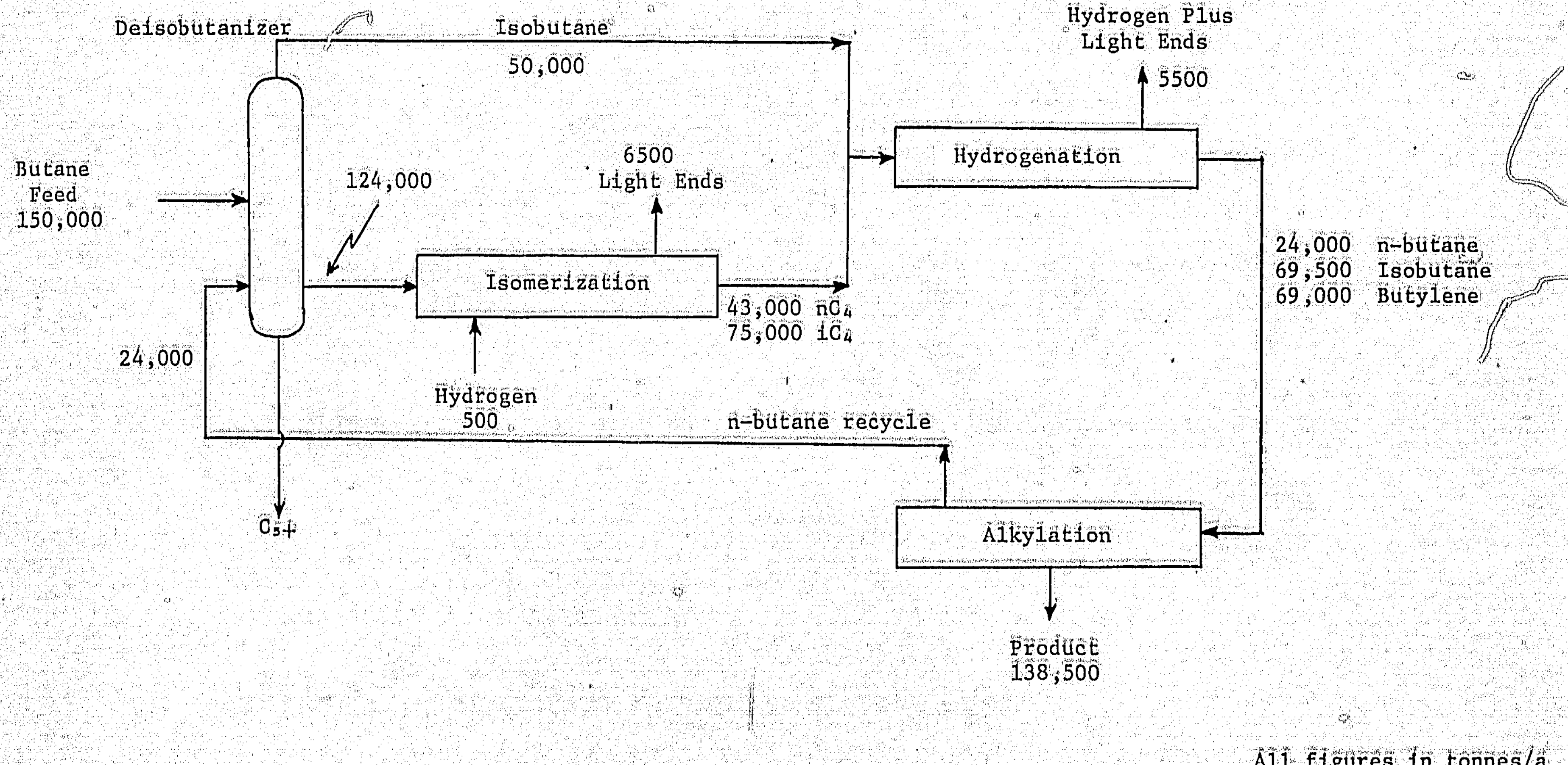

FIG. 3: PRODUCTION OF ALKYLATE FROM BUTANE

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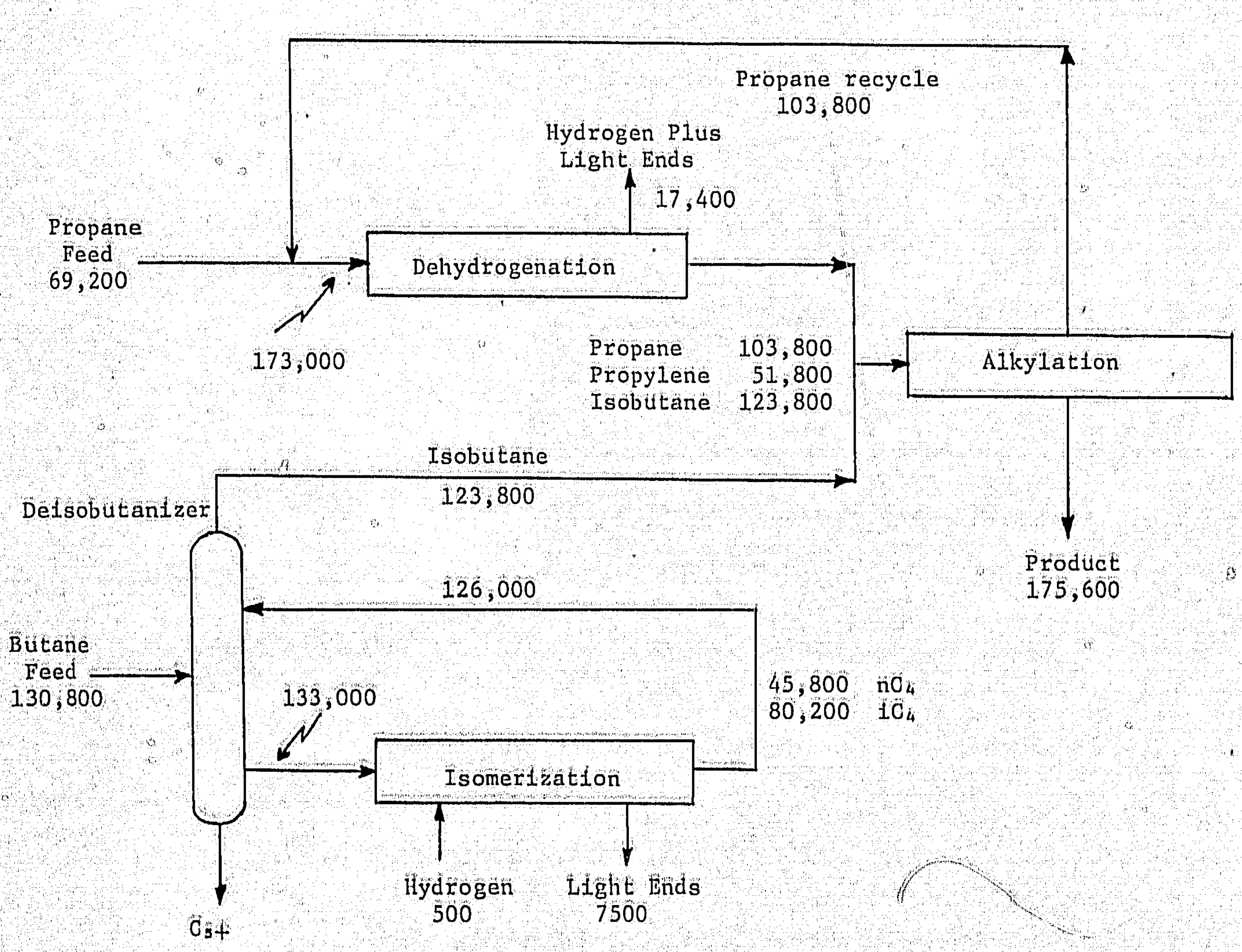

All figures in tonnes/a

FIG. 4: PRODUCTION OF ALKYLATE FROM BUTANE AND PROPANE

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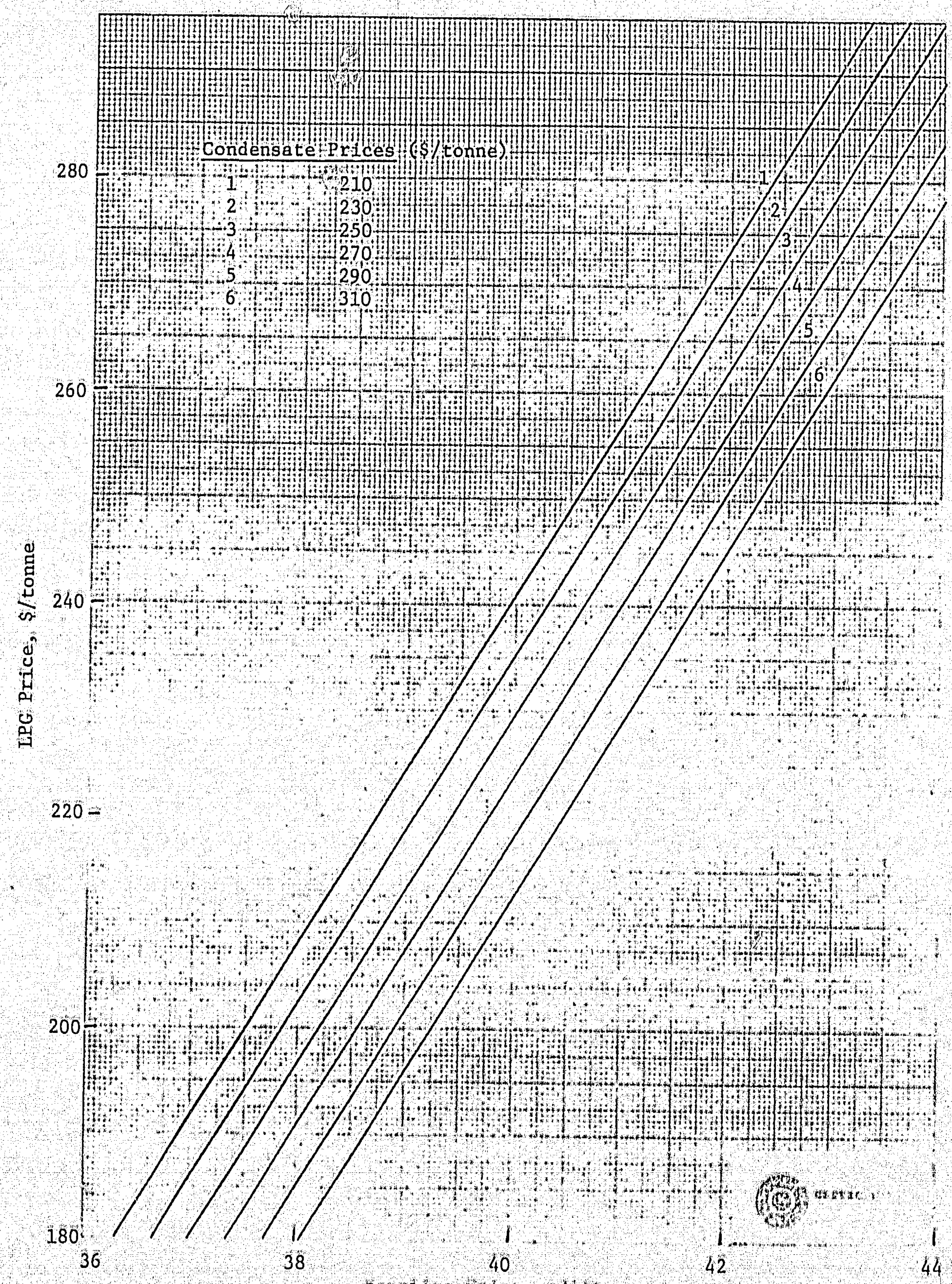

FIG. 5: GASOLINE PRICE VERSUS RAW MATERIALS COSTS (BUTANE ONLY)

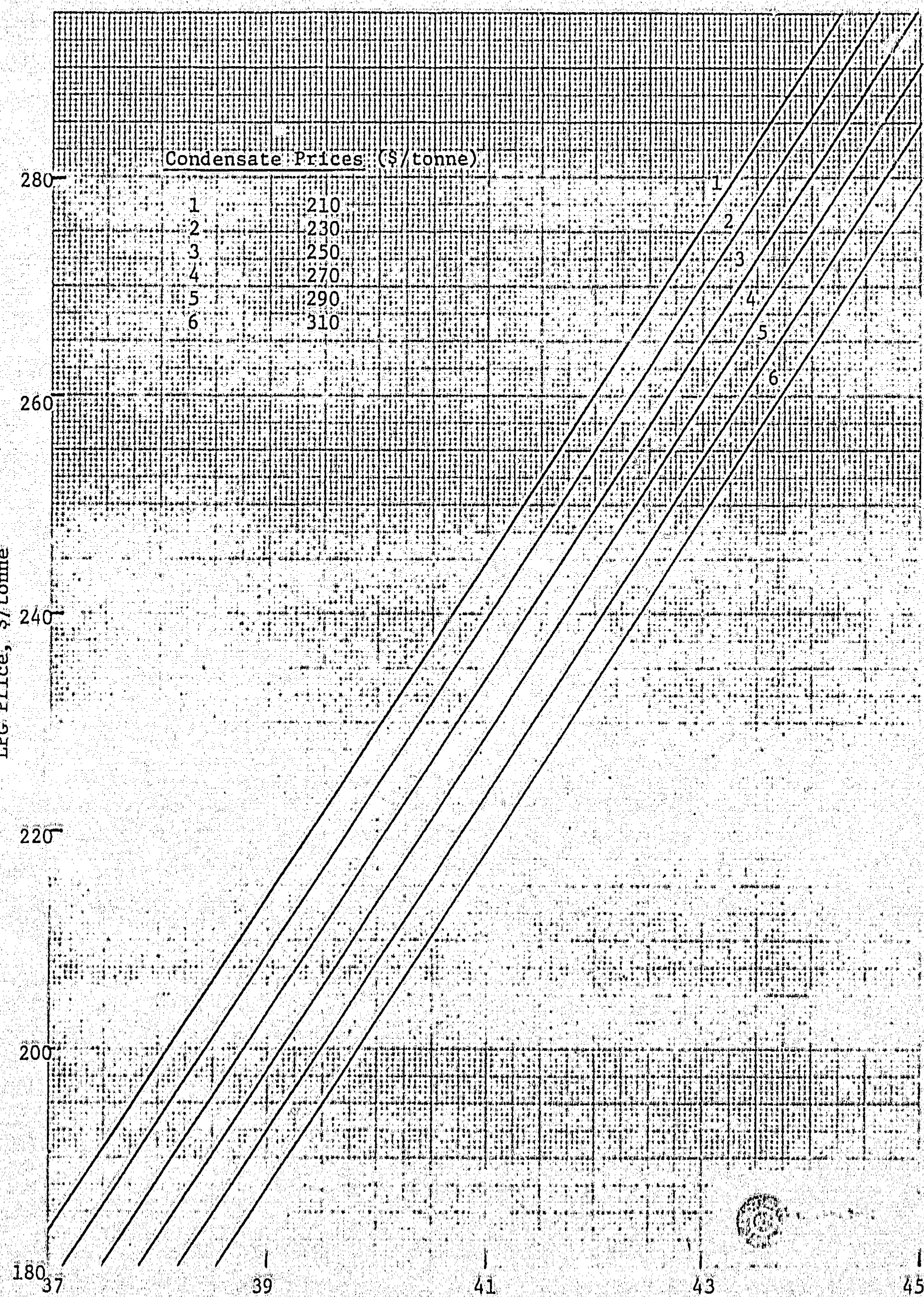

FIG. 6: GASOLINE PRICE VERSUS RAW MATERIAL COSTS (BUTANE PLUS PROPRANE)