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COAL LIQUEFACTION

KERNFORSCHUNGSANLAGE JUELICH G.M.B.H.
(GERMANY, F.R.)

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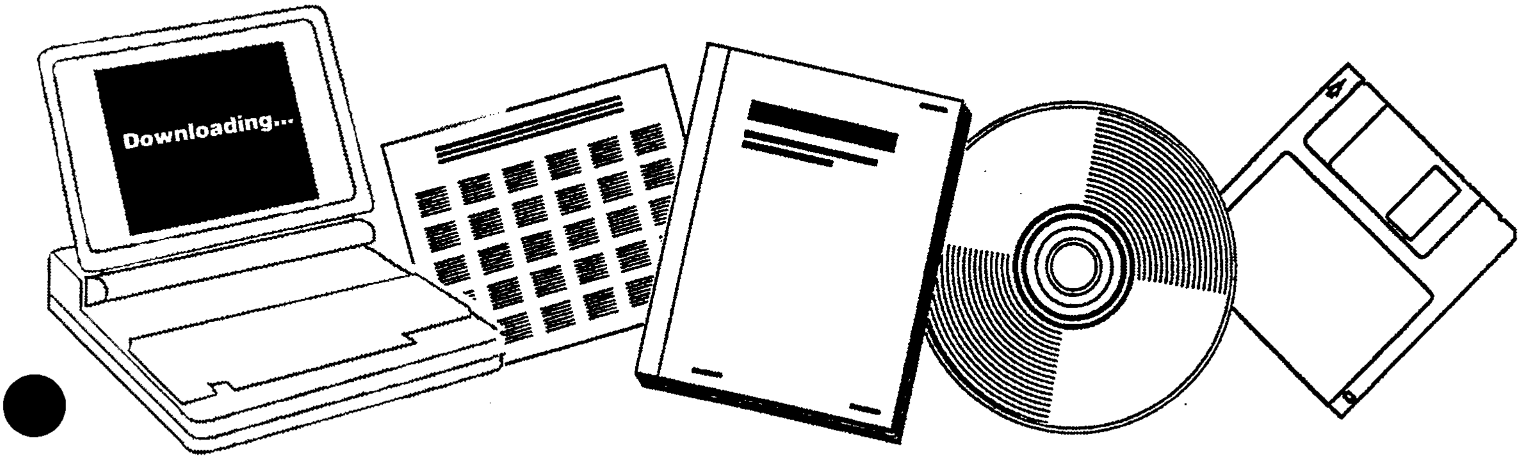

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KERNFORSCHUNGSANLAGE JÜLICH GMBH

WORKSHOP

"COAL LIQUEFACTION"

18./19. SEPTEMBER 1978

VERANSTALTET VON DER
PROJEKTLEITUNG ENERGIEFORSCHUNG

Workshop "Coal Liquefaction" 18./19. September 1978

Teilnehmerliste

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Dr. Schnur	Ruhrchemie AG
Dr. Hubert	Schering AG
Dr. Schramm	Schering AG
Dr. Wolowski	Ruhrkohle AG
Dr. Klusmann	Ruhrkohle AG
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Dr. Knudsen	DOE/USA
Dr. McNeese	ORNL/USA
Dr. Hill	Escoe/USA
Mr. Baker	IEA/Großbritannien

KFA-Angehörige, die an der Konferenz teilnahmen:

Dr. Stöcker	KFA/PLE
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Herr Bertram	KFA/PLE
Dr. Manthey	KFA/STE
Dr. Eickhoff	KFA/STE

WORKSHOP COAL LIQUEFACTION

K.F.A., 18 - 19 September, 1978

MINUTES

HOLIGHAUS: During our first session, I will chair the workshop. I would like to introduce Dr. Engelmann, who is a member of the board at K.F.A., Jülich.

ENGELMANN: I would like to welcome you to the K.F.A. I hope that your workshop on coal liquefaction will result in a good assessment of this technology. Coal liquefaction is increasing in importance; this will be discussed in greater detail by Dr. Ziegler. We are now eyperiencing a "coal renaissance", coal being a primary source of energy which is readily available in large quantities. A potential use of coal is to replace, to some extent, natural gas.
We as a nuclear research centre hope that nuclear energy will also have its share, and we see that coal liquefaction and coal gasification can make use of nuclear power. At Jülich, we are very much engaged in developing the high - temperature gas - cooled reactor as the source of both electric power and nuclear process heat for the purpose of coal gasification and perhaps, later on, also for coal liquefaction.
I hope the meeting will be successful; I will now call upon Dr. Ziegler.

HOLIGHAUS: Our next speaker is Dr. Ziegler, from the Federal Ministry of Research and Technology, where he is director of the office responsible for the non-nuclear energy R+D. programme.

ZIEGLER: Ladies and Gentlemen, I am glad to welcome you at the Kernforschungsanlage Jülich, too. This is the first time that we execute a workshop under the IEA Agreement on national planning cooperation in the field of coal hydrogenation technology.

Coal liquefaction is probably one of the most important technologies for the conversion of coal in the future besides carbonization. Both techniques lead to products which cannot be replaced by other products using today's technologies. There is no real alternative to using gasoline in cars or airplanes. Coke hase to be used in the blast furnace for the near future.

The projects for the future supply of mineral oil show a broad spectrum of possibilities. Some tell us that within the next decade we shall fall short of oil, others promise, that oil will be available beyond the year 2000 in sufficient quantities. These uncertainties are called risks in normal life. The response of most people to such risks is to go to an insurance company. But where can the nations go and insure themselves against risks of damage to their conditions of life? Well, they can only take their fortune in their own hands. They can look for alternative ways of life. To allow cars and airplanes to disappear from the surface of the earth is surely not a possible alternative. Therfore we have to look for an alternative supply of fuel. Coal liquefaction is the way to do this.

If one considers coal liquefaction as an insurance-like tool, one has to look carefully at the economics of that tool. During the last five years, the funds spent on or committed for coal liquefaction research tasks in the Federal Republic of Germany amounted to more than DM 300 million. This is nearly 1 % of the production value of gasoline. In terms of insurance mathematics, this means a mean life for time of more than 100 years. This is surely not the period of life for our present supply paths for gasoline. It will be much shorter.

This insurance aspect is, of course, not the main reason for considering the economics of coal liquefaction. Economic consideration are an important part of technology assessment. They show the economic perspective of the technology to be developed. In the case of coal liquefaction I have the impression that during the last five years the following situation has not changed much:

The costs of the conversion of coal to liquid products are just as high as the world market prices for crude oil, but the energy price of the feed coal divided by the conversion efficiency has to be added to the production costs.

This role will not change as long as the oil-producing countries adjust the price for crude oil to the index of industrial good. Thus, the prospects for the market for coal liquefaction technology are not very good. But it might be that my somewhat pessimistic view concerning the commercialization of coal liquefaction is the result of too conservative cost estimates. I am, therfore, glad that this workshop offers an opportunity for exchanging opinions and views on preconditions and prospective results of the economic of coal liquefaction. I am sure that we all will learn from each other how to approach the target of a realistic assessment of coal liquefaction economics.

In that sense I wish the workshop a good success.

HOLIGHAUS: I would like to welcome our guests from the U.S. and the U.K., whose Mr. Baker is concerned with the economic aspects of coal conversion technology at an international level. Dr. Knudsen will deliver the first paper.

FOSSIL FUEL ECONOMICS

by

Christian W. Knudsen
and
Paul O. Hedman

ABSTRACT

A large number of fossil energy processes are now in various stages of research and development around the world to produce substitute fuels for conventional oil and gas. Process design and cost estimation of new processes is an invaluable part of the development process to guide R&D to the most promising processes and to place experimental emphasis on technical problems of greatest priority. Types of design and cost estimation are described as well as the uncertainties involved in the resulting estimates as they depend on data quality and the level of estimate detail. Project and process contingencies are given which have been found to be appropriate to account for the expected under-estimation.

Cost evaluations are described for coal gasification processes taken from the recent C.F. Braun & Co. report which compares new process developments with commercial Lurgi coal gasification. Costs of approximately $5 per million BTU are indicated. Coal liquefaction costs for processes currently at the pilot plant stage of development are discussed. Liquid product costs are indicated between about $3.50 and $5.00 per million BTU. Power generation is examined

on the basis of near-term new and retrofitted plants as well as the longer range potential of combined cycle technology.

INTRODUCTION

Preliminary design and cost estimating of fossil energy processes is the principal means of determining the practical advantages and disadvantages that a given process has compared with others which produce similar products. The results of such comparisons are of particular importance to research and development. They not only indicate those processes which offer promise of technical and economic feasibility in a future market, but also those sections of a process flow scheme which should receive the greatest attention during further development. It becomes quickly apparent that certain unit operations create the heaviest economic burdens on plant investment and product selling price. These areas then become prime targets for innovative engineering.

Successful process-related companies rely greatly on such process analysis to guide their development efforts and to point to new research projects. Inventors pay close attention as well since the royalty they will receive on a new patent will be negotiated as a portion of the savings created relative to the next best alternative.

U.S. Government research and development activities in fossil energy have grown beyond $500 million annually and decisions about program and project direction are strongly influenced by process analyses.

PROCESS DEVELOPMENT AND ANALYSIS

New heavy-industry process development is an expensive and risky enterprise usually conducted by large companies and governments, sometimes in joint venture. The 15 to 20 year development time to first commercialization which has been estimated for new coal conversion processes, for example, practically mandates government-industry cost sharing.

An example of liberal government cost sharing with industry to induce steady development of new coal conversion processes is illustrated by Figure 1. It represents a logical developmental sequence for a hypothetical case. Although no specific case would necessarily follow this example closely, perhaps the composite of a number of cases would be reasonably close.

The example indicates that after conceptual work, exploratory research follows to test scientific feasibility in a unit capable of about one ton of daily coal throughput. Over a period of one to four years for this phase, $10 million or more may be consumed. Next, a process development unit (PDU) is shown to gather the necessary physical, chemical and engineering data. About five years and $20 to $30 million is required for this phase. A large pilot plant is typically the next phase of development and requires about seven years to complete. Project cost for a 100 ton per day plant may approach $100 million. Finally, the last two stages shown by

Figure 1 represent successively larger commercial prototype plants in final preparation for a full-sized 50,000 barrel per day plant (or its thermal equivalent if the product is other than oil). This development scheme is admittedly conservative and perhaps for some cases the exploratory research and PDU phases could be combined. Likewise the pilot plant and demonstration plant phases might be accomplished jointly by a plant size of several hundred tons per day capacity. Nevertheless, the time to reach commercialization would still be almost 15 years.

Guiding process development by design and cost engineering analysis is very important, but complicated by the need to compare estimates taken from various sources. Engineering design and cost estimating procedures and data will differ somewhat when different process groups have been involved. Any significant differences usually can be resolved when the material is well documented. However, two other factors must be considered when two or more estimates are to be compared. The first concerns the degree of engineering effort expended in the design and costing of each estimate. Greater engineering effort generally produces more accurate estimates. The second concerns the quality or reliability of the data being used for the design. Data from the demonstration or commercial development phase is obviously more accurate than that taken from smaller units such as PDU-sized equipment.

These two sources of inconsistencies in estimates can be resolved by means of project and process contingencies. These are allowances to account for differences in the level of engineering effort and in data reliability, respectively. Application of these contingencies adjusts an estimate to a value equivalent to the completion of development when full data is available for all sections of the plant and an accurate detailed estimate can be made.

Project and process contingencies which are being used to compare and resolve process estimates in the Fossil Energy Program, U.S. Department of Energy, are shown in Figure 2. The process contingency is calculated as a percentage of the onsite portion of the plant and represents the additional investment necessary to improve or expand process equipment to reach design conditions, since data taken while developing a process tend to be optimistic. Project contingency is calculated as a percentage of the total onsite (including process contingency) and offsite investment and is then added to obtain the final investment. It allows for errors in cost estimating due to design assumptions, labor productivity and rate assumptions, late delivery of construction materials, and the like. Therefore, it reflects only the uncertainty of constructing a given plant for a given cost and does not depend on the uncertainty of the technical data. It does depend on the type of estimate made as shown in the figure. Typical engineering costs of producing these estimates for a 50,000 barrel per day coal conversion plant are given in parentheses.

The contingency figures shown in Figure 2 resulted from discussions with large U.S. processing firms over the last two years and are based on their process development and plant construction experience. Major contribution was received from Exxon Corporation.

A better understanding of various levels of cost estimates and the accuracy which can be expected from them can be gained by considering Figures 3, 4 and 5. Together these figures describe the basic differences between preliminary, definitive and detailed estimates.

The first step in developing an estimate is setting the design basis. All three estimate types require the same type of design basis information, with the exception that the site specification for the three differs. For example, a detailed design including detailed mechanical drawings requires specification of an actual site and core drillings may be necessary to determine foundation design.

The next step in process estimating is the process design itself (Figure 4). Differences in estimate accuracy are most obvious from consideration of the varying efforts expended in this step. In a preliminary design the effort ends with an equipment list, while in a definitive design detailed specifications are prepared, including piping and instrumentation specifications. This additional information requires a great deal more engineering effort to develop, but it is important to accuracy since process plants contain piping and instrumentation that may represent up to 40 percent of the plant capital investment. A detailed design includes the latter elements plus detailed engineering drawings and plans which may require hundreds of thousands of man-hours to produce. Of course, this effort is appropriate only when actual construction is planned.

The last step is the cost estimating process itself. For preliminary estimates, cost curves, experience factors, and rules of thumb are used, whereas for a definitive estimate, a more detailed estimating procedure is required. Vendor quotes, specific cost indexes, and projected financial conditions are appropriate. For a detailed study, one seeks vendor bids, finances under actual conditions, and studies actual labor rates and productivity for the area in question. Actual labor costs and productivity are extemely important factors which are generally overlooked. The availability of skilled craftsmen and the specifics of union rules vary in different parts of the United States and can have a large effect on the final plant cost.

Reconsidering Figure 2, it is clear that a final investment estimate varies a great deal as a result of the contingencies applied to it. Consider, for example, a coal liquefaction plant producing

50,000 barrels of product oil daily. Onsite investment might be roughly $750 million and offsite investment about $250 million. If these investments had been calculated using data of PDU quality by a preliminary type of estimate, process and project contingencies would be taken as 25 and 20 percent, respectively. Applying these contingencies results in a total investment estimate of $1,425 million or an increase of about 43 percent above the investment base of $1,000 million without contingencies.

COAL GASIFICATION ESTIMATES

Consistent cost estimates for coal gasification processes which are now under development have been made by C.F. Braun & Co. using western U.S. subbituminous coal with 250 million standard cubic feet per day of substitute natural gas production assumed as the standard plant size. The study examines the investments, operating costs, and the resulting prices of the HYGAS, BI-GAS, CO_2 Acceptor and Synthane processes compared with similar figures for the presently-commercial Lurgi gasification technology. Another phase of the same study which will soon be published examines the same processes using eastern U.S. coals.

Figure 6 is a plot of product costs for the various processes calculated by Braun for western coal, assuming 100 percent equity financing, 12 percent discounted cash flow (DCF) rate of return, and 1976 constant dollars. Braun used a 15 percent project contingency for all of these cases, but included no process contingencies in the onsite investments. Note that product costs can be plotted as straight lines when annual operating costs are plotted against total capital requirement.

From the figure one sees that the HYGAS case with the residual char gasified using a steam-oxygen gasifier appears to be the most attractive process at approximately $4.25 per million BTU of product cost. The Lurgi process is about $5.50 per million BTU as is the case for Synthane where excess char is sold outside the plant and slurry coal feeding to the gasifiers is used. BI-GAS and CO_2 Acceptor approach the low-cost HYGAS case. However, the HYGAS case with residual char gasified using a steam-iron gasifier is less attractive than Lurgi, as are two Synthane cases which export electrical power for sale outside the plant.

The type of cost estimate performed by the Braun study is equivalent to a preliminary study and the 15 percent project contingency used is reasonable. However, no process contingencies were used to reflect the differing data quality available for the individual estimates. Given the PDU and pilot data quality of all of the data except Lurgi, process contingencies of 15 to 25 percent are indicated. A value of five percent is suitable for the Lurgi estimate. Application of these additional factors to Lurgi and the three estimates on the figure which are lower cost than Lurgi narrows their cost advantage over Lurgi by about 50 cents per million BTU. This has the result that only the HYGAS process retains an apparent advantage over Lurgi technology. Other processes appear marginal or higher cost compared with Lurgi technology.

COAL LIQUEFACTION ESTIMATES

At present several coal liquefaction processes are under development. These include such processes as Exxon Donor Solvent (EDS), H-Coal, and Solvent Refined Coal (SRC). Each of these processes makes liquid fuels with different physical properties. However, each of the processes has some flexibility to operate over a range between a heavier boiler fuel type of primary product and a lighter synthetic crude primary product, depending on liquefaction reactor space velocity.

A recent paper by Gulf (2) concerning the SRC process operated to produce a synthetic crude (although they view its best use as fuel to a boiler) indicates a price of $3.21 per million BTU assuming 100 percent equity financing, 12 percent DCF and 1976 constant dollars. A 20 percent project contingency is included, but no process contingency was applied. Including a 20 percent process contingency increases the cost to about $3.60 per million BTU. This is equivalent to about $22 per barrel.

Preliminary estimates of other liquefaction processes within Fossil Energy indicate prices of $30 per barrel and greater when using this same economic basis to produce a synthetic crude. However, since the various designs and cost estimates have been made by different concerns, it is not clear whether these cost differences are due to true process differences or merely to design philosophy differences among the various firms involved. This matter is currently under study.

POWER GENERATION ESTIMATES

New electric generation facilities can be based on a number of liquid and solid alternative fossil fuels. Figures 7 and 8 contrast various base load alternatives, showing the capital, operation and maintenance (O&M), and fuel components of total cost expressed as mills per kilowatt-hour of power generated. These power costs were derived from recent work done by Gilbert Associates (3) which determined capital and O&M costs for various alternatives. The fuel component was added to these by choosing recent cost ranges for the basic fuels used (Table I). An 800 megawatt electric plant size operating at 70 percent capacity factor is assumed and the basis is utility economics equivalent to a 10 percent DCF rate of return in 1975 constant dollars. A 15 percent project contingency was used in all cases with no process contingency.

In Figure 7, the No. 6 fuel oil case shows a variation in power cost of 28 to 33 mills per kilowatt-hour (the variation in the fuel component of this and all other cases represents the range shown in Table I). The natural gas case is less, but this fuel is now in scarce supply in the United States. SRC hot liquid refers to the Solvent Refined Coal liquefaction process operated so as to make a heavy liquid product which would solidify if cooled. This case and that for heavy synthetic coal liquid both indicate a significant cost increase compared to No. 6 fuel oil. The dashed area is added to emphasize the relative uncertainty of these estimates. Finally, medium BTU gas made off site and bought by the power plant at the range shown by Table I is also relatively expensive. Note that the capital and O&M components for all of these liquid cases are substantially the same and only the fuel components vary.

The solid fuel cases shown in Figure 8 show some interesting variations. Low sulfur coal without flue gas desulfurization (FGD) is very attractive and compares favorably with the use of natural gas on the previous figure. The high sulfur coal case with FGD illustrates the fact that the additional capital and O&M components due to the FGD equipment are not offset by the lower fuel cost of high sulfur coal. Similarly, installation and operation of an on site low BTU gas plant using high sulfur coal is not offset by the cheaper fuel.

The solid SRC case without FGD has the same low capital and O&M components as the low sulfur coal case but the expensive fuel prices this alternative well above the others. Next, cleaned high sulfur coal without FGD appears competitive with low sulfur coal. Finally, the two high sulfur coal cases using fluidized bed combustion and a low BTU gas, combined cycle system both look very competitive.

Retrofit of base load electric utilities is illustrated by Figure 9 using the same economic basis as before. Here the incremental cost of modifying solid and liquid fuel plants is shown by the three cost components. FGD adds only about 10 mills per kilowatt-hour but solid SRC adds over 20 mills. Among alternatives for retrofitting solid fuel plants, cleaned high sulfur coal adds the least or about five mills. For liquid plants, the heavy synthetic coal liquid and the medium BTU gas off site cases add about 10 mills per kilowatt-hour or more. The low BTU gas on site case adds nothing because the savings in fuel cost by using high sulfur coal to generate the gas offsets the capital and O&M components. The coal oil slurry case indicates a reduction, since the needed capital and O&M are not large and the savings in No. 6 fuel oil substituted by less expensive low sulfur coal more than offsets them.

The economics of steam generation by fluidized bed combustion (FBC) have recently been studied (4). Figure 10 contrasts FBC with conventional firing (CF) for both high and low sulfur coal;

conventional firing with low sulfur fuel oil is shown for comparison. These costs show capital, O&M and fuel components (see Table I) calculated in 1975 constant dollars at a 10 percent DCF rate of return for a 100,000 pound per hour boiler. No process contingency was assumed, but a 20 percent project contingency was used.

For high sulfur coal, the FBC case is definitely lower cost than conventional firing with FGD. There is no relative improvement when using low sulfur coal, however. Note that the capital and O&M costs for a boiler based on low sulfur fuel oil is much less than the other cases. Of course, this is fully offset by the relatively higher cost of the fuel oil.

SUMMARY

Consistent process design and cost estimating procedures play an important role in guiding research and development. Application of proper process and project contingencies is a key element in obtaining realistic and comparable estimates.

Preliminary estimates have been made for many of the coal conversion and power generation alternatives now under development in the United States. Coal gasification and power generation economics are presently the most fully developed, but a number of studies are planned to better define the prospects for coal liquefaction.

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3. "Assessment of Fossil Energy Technologies for Electric Power Generation," Vol. 1, prepared for the Office of Program Planning and Analysis, Fossil Energy, by Gilbert Associates, Inc., October 1976.

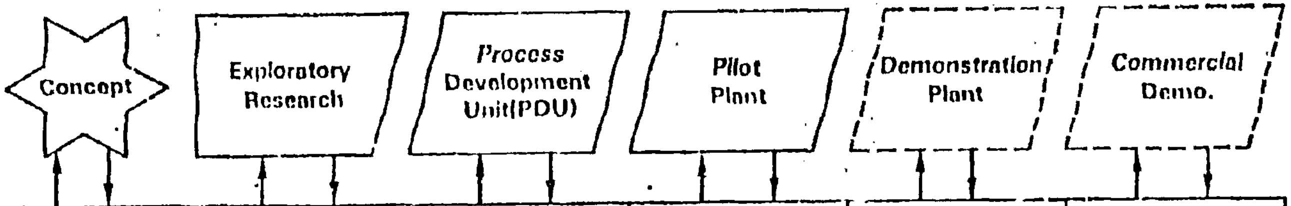
4. Farmer, M.H., E.M. Magee, and F.M. Spooner, "Application of Fluidized Bed Technology to Industrial Boilers," prepared for U.S. FEA/ERDA/EPA by Exxon Research and Engineering Company, Linden, N.J., January 1977.

TABLE I

FUEL COST TO POWER GENERATION

	Dollars per Million BTU
Liquid Fuels	
No. 6 Fuel Oil	2.12 - 2.86
Natural Gas	0.52 - 2.00
SRC Hot Liquid	3.00 - 5.00
Heavy Synthetic Coal Liquid	3.00 - 5.00
Medium BTU Gas	3.00 - 4.00
Solid Fuels	
Low Sulfur Coal	1.00 - 1.25
High Sulfur Coal	0.75 - 1.00
Solid SRC	3.00 - 5.00

Purpose, Size, Cost of Individual Coal Conversion Units



Purpose:	Discovery	Scientific Feasibility	Technical Feasibility	Economic Feasibility	Commercial Feasibility	Resolve Investment Uncertainties
Information:	Theory	Concept Proof	Physical, Chemical, Engineering Data	Engineering Parameters of Scale-up	Validate Process Economics, and Environmental/ Socioeconomic Impacts	Capital and Other Resource Requirements, Marketability of Products
Typical Size: 1/ (Tons/Day)	0 to 0.1	1.0	10	100	1,000	10,000
Capital Cost: 1/ (1976 Million $)	N.A.	$3 to $5	$10 to $15	$20 to $30	$100 to $200	$400 to $800
Annual Operating Cost, 1/ (1976 Million $)	N.A.	$3 to $5	$5 to $10	$10 to $15	$25 to $50	$80 to $160
Government Share: 1/ (Percent)	N.A.	100%	100%	66%	50%	0 to 50% (Cost if Venture Fails)

1/ Typical Values; each process is different & must be individuallly estimated.

Figure 1

PROJECT AND PROCESS CONTINGENCIES*

TYPE OF COST ESTIMATE

DEVELOPMENT PHASE

PROJECT % PROCESS %	STUDY ($\$2-5 \times 10^4$)	PRELIMINARY ($\$2-5 \times 10^5$)	DEFINITIVE ($\$2-5 \times 10^6$)	DETAILED ($\$2-5 \times 10^7$)
PDU	25 50	20 25		
PILOT	25 25	20 15	15 10	
DEMONSTRATION	25 15	20 10	15 5	10 5
COMMERCIAL	25 5	20 5	15 5	10 5

*PROCESS CONTINGENCY IS APPLIED TO ONSITES; OFFSITES ARE THEN ADDED AND PROJECT CONTINGENCY IS APPLIED TO THE TOTAL

Figure 2

DESIGN BASIS

PRELIMINARY ($\$0.2\text{-}0.5 \times 10^6$)	DEFINITIVE ($\$2\text{-}5 \times 10^6$)	DETAILED ($\$20\text{-}50 \times 10^6$)
PRODUCT SPECS	DO	DO
FEED SPECS	DO	DO
DESIGN ASSUMPTIONS	DO	DO
PROCESS DESCRIPTION	DO	DO
UTILITY SPECS	DO	DO
GENERAL SITE	HYPOTHETICAL SITE	ACTUAL SITE

Figure 3

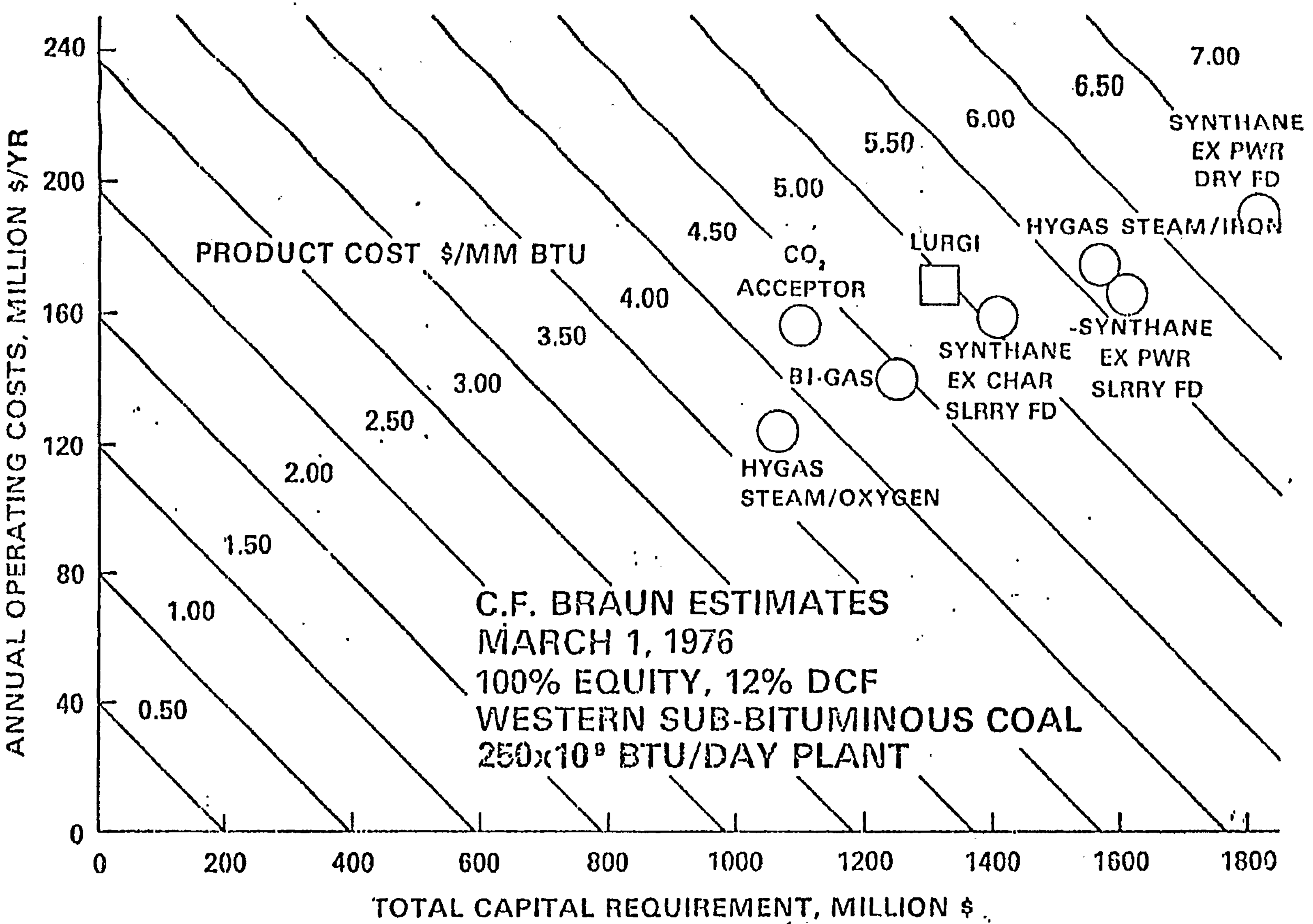
PROCESS DESIGN

PRELIMINARY (\$0.2-0.5 $\times 10^6$)	DEFINITIVE (\$2-5 $\times 10^6$)	DETAILED (\$20-50 $\times 10^6$)
• FLOW DIAGRAM	• DO	• DO
• MATERIAL BALANCE	• DO	• DO
• ENERGY BALANCE	• DO	• DO
• OPERATING CONDITIONS	• DO	• DO
• PLOT PLAN	• DO	• DO
• ENVIRONMENTAL ASSESSMENT	• DO	• ENVIRONMENTAL IMPACT STATEMENT
• MAJOR EQUIPMENT SIZED	• ALL EQUIPMENT SIZED	• DO
• EQUIPMENT LIST	• EQUIPMENT LIST AND DETAILED SPECS	• DO
	• P AND I DIAGRAMS	• DO
	• PIPING SPECS	• DO
	• PROCESS RELATED STRUCTURAL SPECS	• COMPLETE STRUCTURAL DRAWINGS
		• DETAILED ENGINEERING DRAWINGS
		• PLANT ELEVATION DRAWINGS
		• PROCUREMENT AND CONSTRUCTION PLAN

Figure 4

PROCESS ECONOMICS

PRELIMINARY ($\$0.2\text{-}0.5 \times 10^6$)	DEFINITIVE ($\$2\text{-}5 \times 10^6$)	DETAILED ($\$20\text{-}50 \times 10^6$)
○ COST CURVES	● DO	● VENDOR BIDS
○ EXPERIENCE FACTORS	● VENDOR QUOTES ON MAJOR ITEMS	○ ACTUAL LABOR COSTS AND PRODUCTIVITY
○ RULES OF THUMB	● EXPERIENCE FACTORS BASED ON MORE DETAILED DRAWINGS	● DETAILED ENGINEERING EVALUATION
○ GENERAL COST INDEXES	○ SPECIFIC COST INDEXES	○ FINANCING UNDER ACTUAL CONDITIONS
○ ASSUMED FINANCIAL CONDITIONS	○ PROJECTED FINANCIAL CONDITIONS	



C.F. BRAUN ESTIMATES
MARCH 1, 1976
100% EQUITY, 12% DCF
WESTERN SUB-BITUMINOUS COAL
250x10^9 BTU/DAY PLANT

Figure 6

NEW ELECTRIC UTILITIES

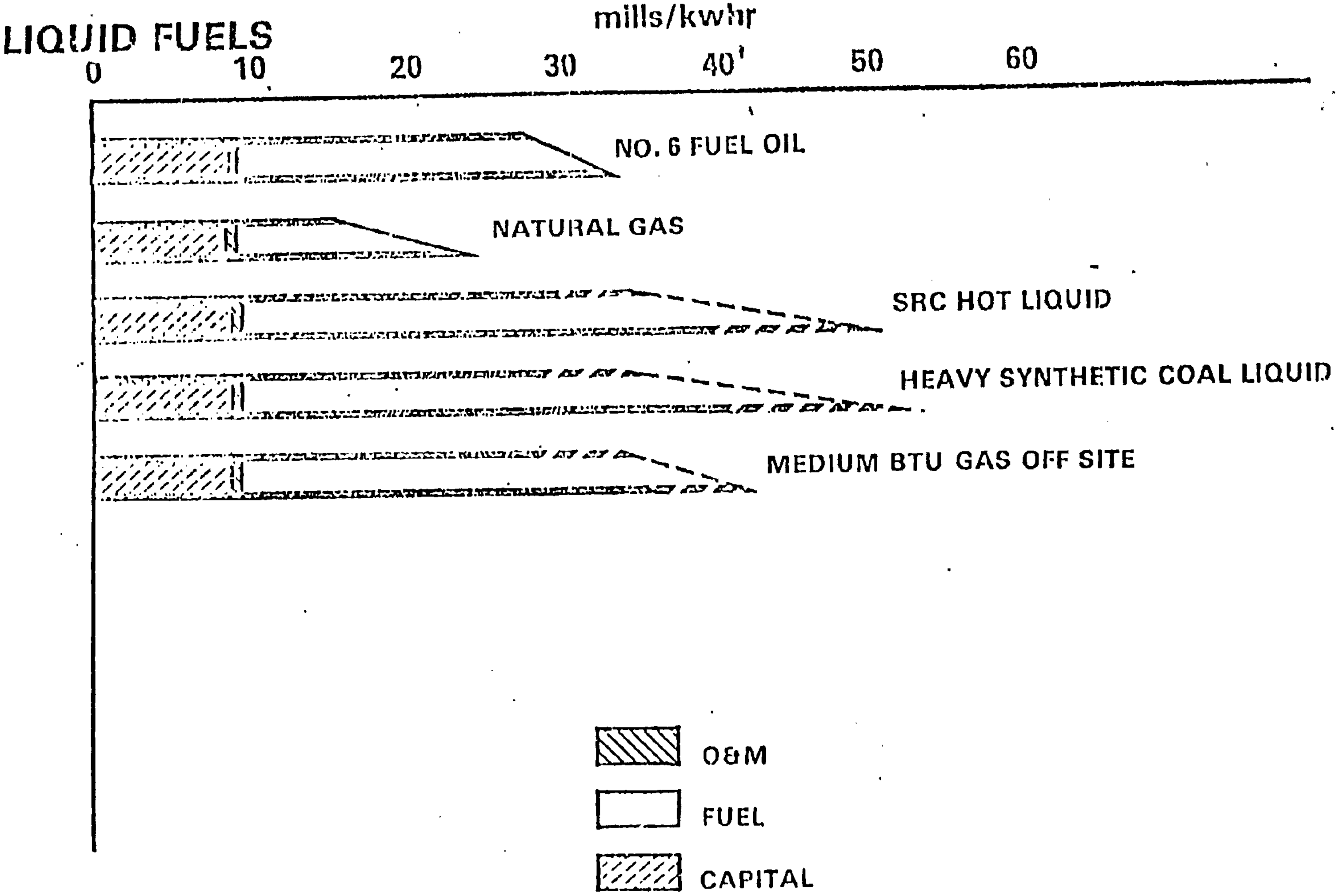


Figure 7

NEW ELECTRIC UTILITIES

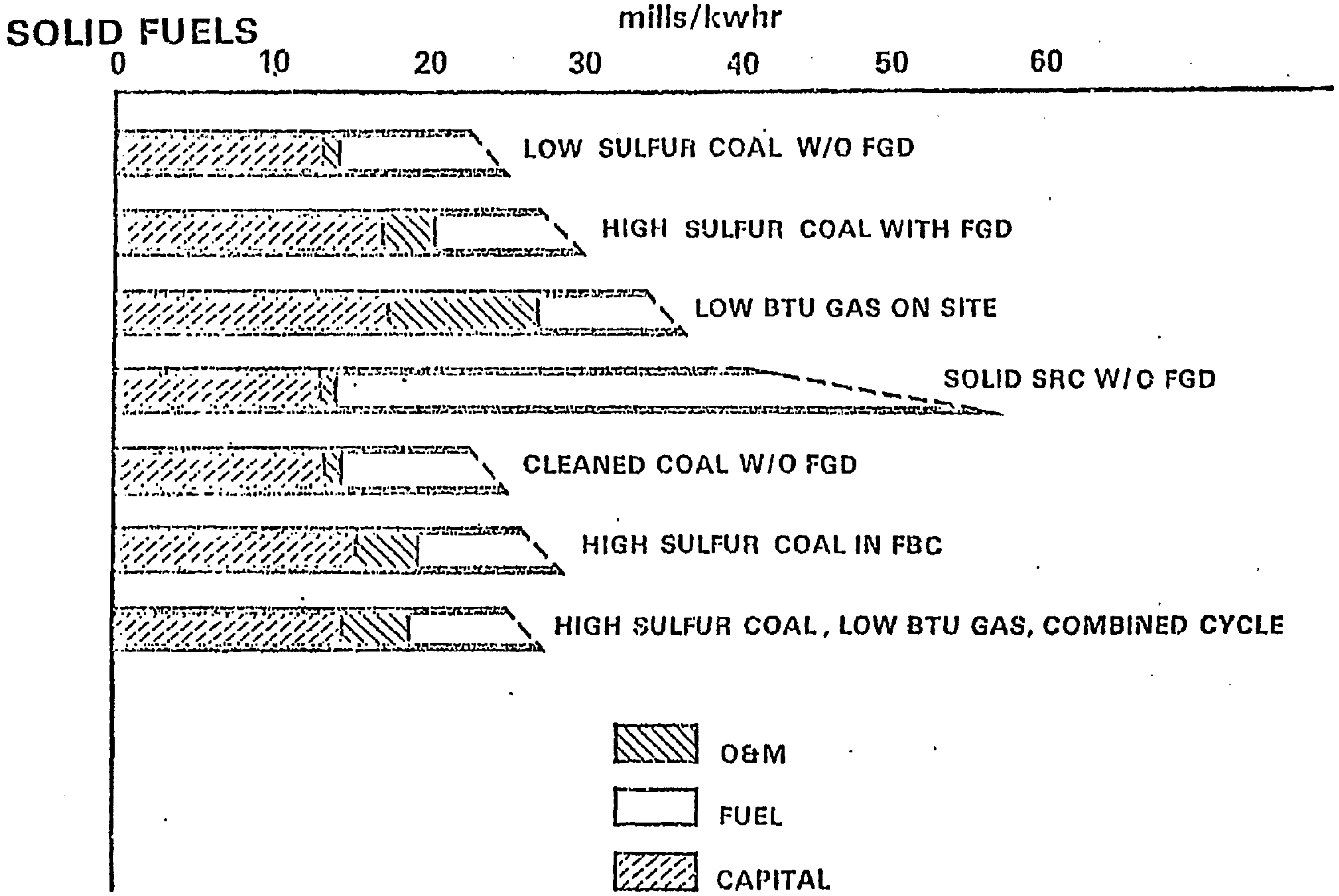


Figure 8

RETROFIT OF ELECTRIC UTILITY

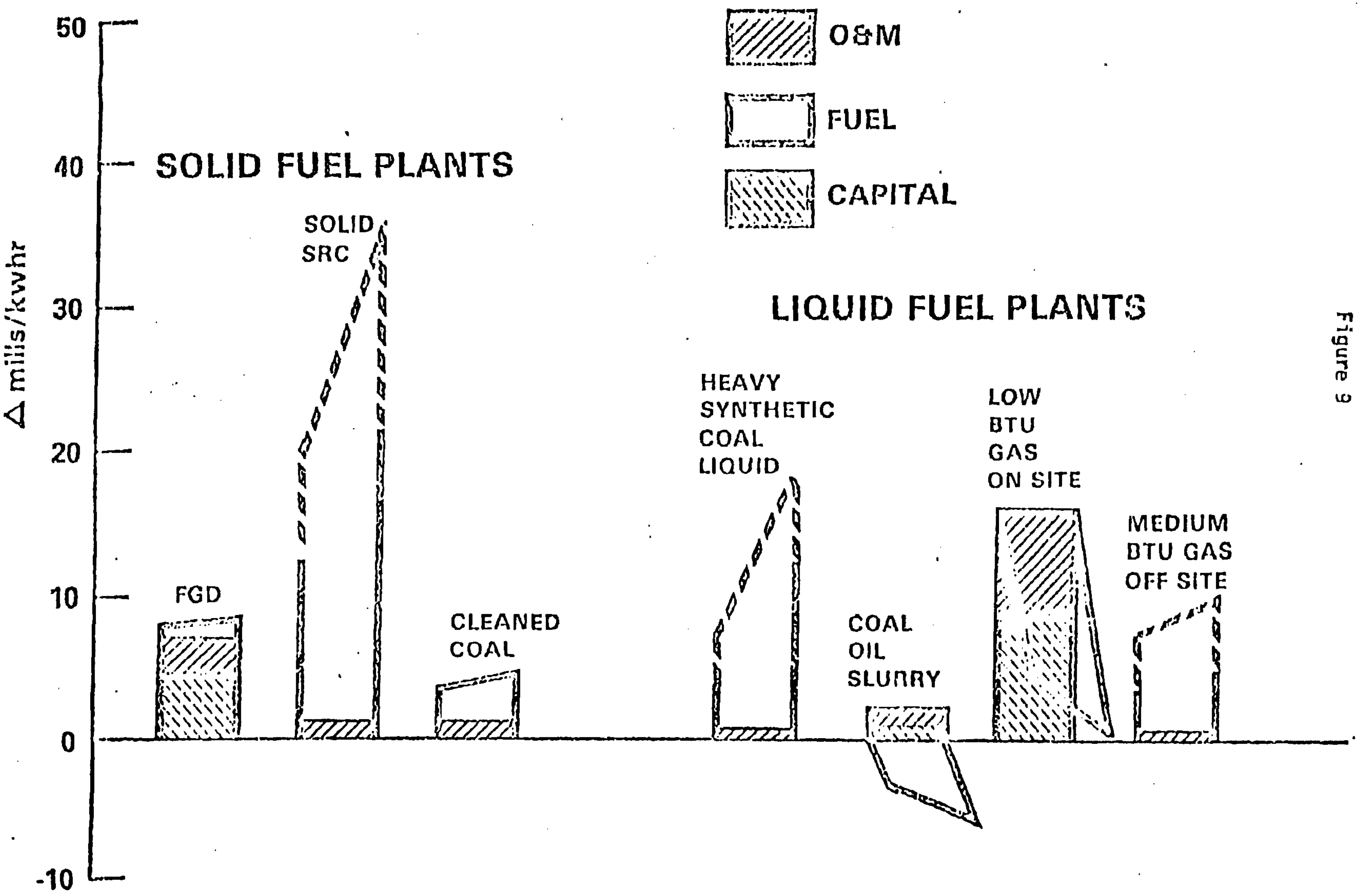


Figure 9

NEW INDUSTRIAL BOILERS

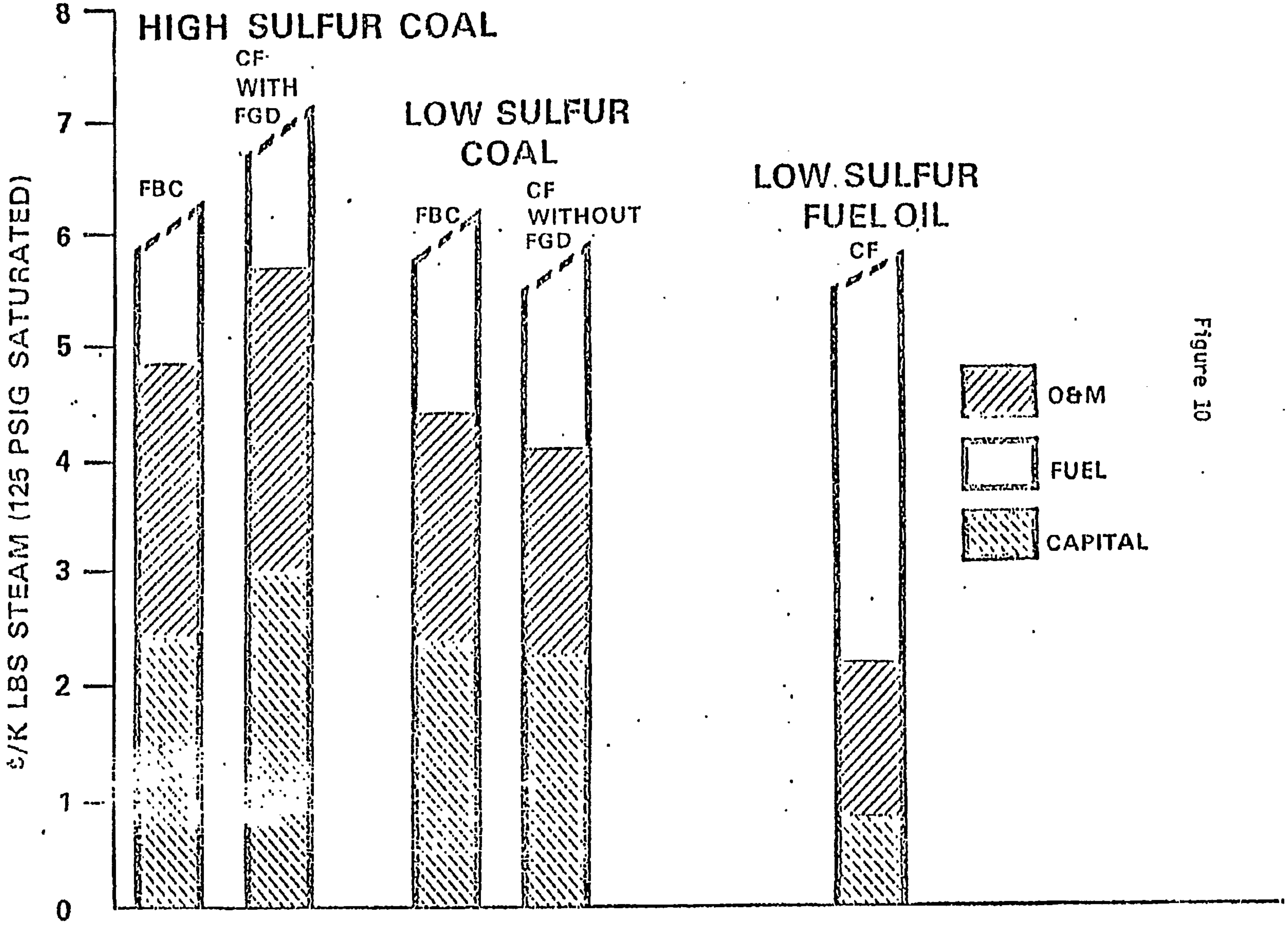


Figure 10

LECTURE - KNUDSEN

HOLIGHAUS: Thank you for your excellent paper giving a very good impression of how you work, your considerations and aims. The discussion is now open.

ZIEGLER: The first graph showed that when substituting any other fuel by a liquid product from coal, (particularly in the electricity sector) the costs were much higher than when other methods of supplying the power plants with fuel were used. Does your agency conclude from these results that one should not produce liquid products for the electricity sector?

KNUDSEN: No, but this matter is now under consideration. There are specific cases: Gulf has recently performed a marketing study which shows that the process can be very attractive in specific locations, which can project a market penetration for the next 20 years by the specific locations that make the process look quite attractive to them. The problem with these types of comparisons is that they are so broad and over-simplify the matter in that they indicate that on thing should surpass all others. In this spcific case, an estimate has been developed for the substitution of S.R.C. Gulf is interested in the so-called S.R.C.2 material; they present a very reasonable case for its introduction in specific situations to retrofit. One can easily be misled by the retrofitting slide into thinking that any plant might be retrofitted in a simple manner by some other solid or liquid - this is not always the case.
In the specifics, a healthy portion of the market sufficient to attract an interest has been defined, and we are in general agreement at the agency with that and are very interested in proceeding with the process. It is generally recognised by anyone experienced in this business that it is extremely difficult to build these enormous refineries and pump coal through them with all the attendant problems and to be able to be on a par in the general case with some basis that has a low - sulphur coal and fines it directly with a minimum of other handling.

HOLIGHAUS: All your considerations aplly only if one examines the base - load case. If one looks at the middle - load, the situation would change; the liquids from coal would have a better chance.

KNUDSEN: That is a very good comment. All these comparisons were for the base-load case and assume a 800 MW plant with 70 % capacity factor. All three comparisons were based on this very large plant on high-capacity factor, which is a high base-load rate for a coal plant.

HOLIGHAUS: This consideration concerns the contingency which indicates that one must expect much higher costs than those estimated today based on pilot or P.D.U. plants. This is rather pessimistic in so far as the results of the developments are concerned. Is there any case where the results of the development of future steps indicate a decrease in costs compared with earlier assessments?

KNUDSEN: Unfortunately, all available information showed a bias towards cost-estimating. Years ago, someone may have made a preliminary estimate of a coal liquefaction process and assumed that it would be a simple matter to mix it in an oil cresote slurry and pass it through a heated tube and raise its temperature. Slurry pre-heating tends to be a very difficult problem, running into considerable design sophistication and expense; therefore we do not advocate it strongly.
This is a statistical and general picture; there are a few processes, however, which might offer new leads and directions and be responsible for constant progress across a wide field of technology. The indications, on average, are positive.

WÜRFEL: The last slide showed electricity generation using several fuels; why should the use of a heavy fuel oil like HCOal be cheaper for electricity generation than SRCI?

KNUDSEN: The new plant was shown in the slide. We are dealing with a delta-change in cost. The question is: which additional costs are above or below the base situation, in this case a high-sulphur conventionally fired coal plant? But this is a low-sulphur fuel oil fired plant. What change in the mls. per kW hr. would result from these applications? This is so much higher than this one because we are substituting this for a very cheap fuel - the high - sulphur coal, which creates a differential. We obtain a credit for coal not used and then pay (75 cents - I dollar for the cheaper coal) in BTUs. We pay 3-5 dollars for this, and for the heavy synthetic coalliquid. This is "chained off" against a low-sulphur fuel oil, obtaining a credit of 2 dollars (in this case, 2 dollars 12 cents - 2 dollars 28 cents). Low numbers are subtracted from low numbers, and high numbers from high numbers: this is a credit and debit system.

STÖCKER: You showed that the prices were equal.

KNUDSEN: Yes, they are the same.

HILL: The reason Dr. Knudsen is looking at a refitting situation in a coal fired plant with SRCI is that new U.S. enviromental standards must be met. These state that some coal fired plants must use a cleaner fuel or employ stack gas scrubbers. Examination of the conversion of this solid-fired plant is undertaken because of the new regulations.

HOLIGHAUS: I would now like to close the discussion and ask Mr. Gaensslen from Lurgi to present his paper.

Thermal efficiency and production economics

of

chemical plants

Hans Gaensslen

September 1978

Introduction

During optimization calculations for the German Ministry for Research and Technology (BMFT) the investment costs of a large number of complex organic plant had to be determined. Doing this in the classic way by estimation proved extremely time-consuming. Therefore, it was desirable to find a calculation method that would yield quickly the investment cost of such a plant requiring at the same time a minimum of data.

For this purpose a number of chemical plants, all using organic substances as raw material, were subjected to a statistical analysis in order to find a suitable correlation between investment cost and certain characteristic parameters of the plant. After many failures it was found that investment cost of such chemical plants is related to energy transformation in the process, the result being independent of the internal structure of the plant. As a matter of fact, it is possible to estimate by the method described in this paper the investment cost of a chemical plant without even the slightest notion of how the plant is designed; it suffices to know the input and the output of said plant.

Investment cost estimation. First approximation

The relationship between investment cost and energy is very simple: in a large and complex plant investment cost forms a linear correlation with the energy generated by the system per unit of time. For practical purposes we may say that the investment cost of a plant is directly proportional to the dissipated energy per year. The dissipated energy is the difference between all yearly inputs and the products output, both expressed in consistent energy units.

This means that

$$I = \xi \dot{D} \qquad (1)$$

where I is the investment, ξ is a proportionality factor, and $\dot{D}$ is the dissipated energy per unit of time. The value of ξ will depend on the currency, the location. as well as on the energy unit used, and will have to be escalated in accordance with inflation. In West-Germany, in mid-1976, ξ had the following values

$$\xi = 100 \text{ DM/(Gcal/a)}$$

$$= 23.9 \text{ DM/(GJ/a)}$$

$$= 0.688 \text{ DM/W}$$

$$= 25.2 \text{ DM/(MM Btu/a)}.$$

In Tables 1, 2, and 3 the values of ξ for a number of plants based upon organic raw materials are found, as well as the input and output information that led to the results.

Dividing expression (1) by the yearly capacity of the plant we arrive at the specific investment, I_s:

$$I_s = \xi D_s = \xi E \left(\frac{1 - \eta}{\eta}\right) \qquad (2)$$

Where D_s represents the dissipated energy per ton of product, E is the LEV of the product or products, and η is the overall thermal efficiency, based upon LEV.

The successful application of eq. (1) or (2) depends on the following conditions:

- chemical processes based on organic raw materials and involving substantial energy transformations
- complex production process
- large capacity; scale-up exponent close to one
- average mix in the plant of equipment, materials and catalysts
- plant with today's degree of thermal optimization
- the bulk of the dissipated energy to leave the system through equipment walls.

If these caveats are considered the method yields investment costs of the process plant within battery limits with a maximum error of ± 20 %. The value obtained corresponds to the process-dependant investment and does not include the cost of tank farms, administrative buildings, workshops, fence and flag pole, and other items not directly relevant to production.

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$$I_s = \xi D_s = \xi H \left(\frac{1 - \eta}{\eta}\right) \qquad (2)$$

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For the practical application of eq.(1) or (2) the following values are used:

- input raw materials	LHV
- input fuel gases	LHV
- input steam	500,000 kcal/t
- input electricity	860 kcal/kWh
- output products, by-products and fuel gases	LHV
- output excess steam	not considered.

Example 1

The production of 2 million metric tons per year of a certain organic product requires the following input:

2,800,000 t/a coal, LEV = 7,500 kcal/kg

$16 \cdot 10^8$ kWh/a electrical energy

1,000,000 t/a steam.

If the LEV of the organic product is 8,000 kcal/kg, what is the investment cost of the corresponding plant?

Answer:

Input		
Coal, as energy	21,000,000	Gcal/a
Electrical energy, 16,000,000 kWh à 860 kcal/kWh	1,376,000	Gcal/a
Steam, 1,000,000 t/a à 500,000 kcal/t	500,000	Gcal/a
Total	22,876,000	Gcal/a
Output		
Product, 2,000,000 t/a à 8,000,000 kcal/t	16,000,000	Gcal/a
Dissipated energy per year	6,876,000	Gcal/a

Hence, investment cost, West-Germany, 1976, amounts to

$$I = 6,876,000 \times 100 = 687,6 \text{ million DM}$$

and the specific investment is

$$I_s = 687.6\text{x}10^6/2\text{x}10^6 = 343.8 \text{ DM/(t/a)}$$

As the utilities are imported from outside battery limits the investment costs do not include the equipment or installations for the production of these utilities.

Example 2

A liquid fuel with a LHV of 10 Gcal/t is produced from coal with an overall thermal efficiency of 42 %. This figure includes the generation of utilities required for production. What is the investment cost of a plant capable of producing 3 million t of fuel per annum?

Answer:

Specific energy input: 10/.42 = 23.81 Gcal/t of product

Annual total energy input, 3,000,000 x 23.81	71,430,000 Gcal/a
Annual output, $3\text{x}10^6\text{x}10$	30,000,000 Gcal/a
Dissipated energy per annum	41,430,000 Gcal/a

The investment cost (West-Germany, 1976) amounts to

I = 41,430,000x100 = 4.143 billion DM.

As the utilities required for production are generated inside battery limits the investment cost will include the equipment or installations for the production of these utilities.

Justification of the first approximation

Although a theoretical deduction of the value of ξ is not possible at present, a statistical justification can be given. The results mentioned at the beginning of chapter 1 indicate that chemical plants behave like large heat exchangers as far as investment is concerned. These huge heat exchange systems are needed to transport the energy from the system to the surroundings. Such a heat exchange system requires heat transfer surfaces which again determine the investment.

It would be sufficient to calculate the cost of a heat exchange surface capable of dissipating 1 Gcal/ a, because this would correspond to the value of ξ . The difficulty lies in the fact that the variables involved in the calculation are not fixed but can vary within a certain range. These variables are:

- the temperature gradient across one heat exchange surface
- the heat transfer coefficient
- the number of heat exchange surfaces the energy has to pass
- the specific cost of a heat exchanger
- ratio of carbon steel to (say) stainless steel in a plant
- the modified Lang factor (i.e. based only on the heat exchanger cost instead of on cost of all equipment)
- the overdesign factor.

The values of the variables and their corresponding range used in the calculations are shown in Table 4. In the computations a large number of values of ξ were calculated by the Monte-Carlo method using statistically randomized sets of variables within the range of variation. The average ξ found this way was 95 DM/(Gcal/a) which is in good agreement with the empirical value of 100 DM/(Gcal/a), both based on West-German conditions in 1976.

Production cost calculations based upon the first approximation

If the capital cost charge per annum is z, and H_u represents the the lower heating value of the product, the production costs K can be deducted from eq. (2), yielding

$$K = z\xi H_n (1 - \eta)/\eta + H_u \kappa/\eta \qquad (3)$$

where κ stands for the raw material cost per energy unit.

If two raw materials are available at different costs the condition for identical production cost will follow from eq. (3) leading to

$$\kappa_2 = z\xi(\eta_1/\eta_2 - 1) + \eta_2\kappa_1/\eta_1 \qquad (4)$$

where the subindices refer to the respective raw materials. Eq. (3) and (4) are intended for the actual process units only. In case of a complete integrated plant ("grass roots" basis) the first terms of the expressions would have to be multiplied by a off-sites factor, which normally ranges between 1.1 and 1.5.

The numerical results obtained from eq. (2), (3), and (4) are very satisfactory and compare well with actual figures worked out by the more cumbersome classical estimating methods. Time needed to obtain a result by the described method is approximately 1 % of the time required by the classical method.

Example 3

Production of gasoline by coal hydrogenation leads to the following thermal efficiencies:

For brown coal wtih LHV 1.6 Gcal/t $\eta_1 = 53$ %

For hard coal with LHV 7.5 Gcal/t $\eta_2 = 60$ %.

What will be the cost of hard coal (subindex 2) that will lead to identical production with brown coal (subindex 1)?

The following parameters are used:

$z = 0{,}2\ a^{-1}$

$\xi = 111$ DM/(Gcal/a)

$f = 1.3$ (off-sites factor).

Using eq. (4) we obtain

$$\kappa_2 = 0.2 \cdot 111 \cdot (60/53 - 1) + 60 \cdot \kappa_1 / 53$$
$$= 3.812 + 1.132\ \kappa_1 .$$

This expression can be represented graphically or in form of a table:

brown coal		hard coal (equivalent cost)	
κ_1 DM/Gcal	DM/t (§)	κ_2 DM/Gcal	DM/t (§§)
1	1.60	4.9	37
2	3.20	6.1	46
3	4.80	7.2	54
4	6.40	8.3	63
5	8.00	9.5	71
7	10.20	11.7	88
10	16.00	15.1	113

(§) at 1.6 Gcal/t (§§) at 7.5 Gcal/t

The figures in columns 3 and 4 represent the hard coal costs for which production costs are identical with the corresponding brown coal. If brown coal cost 5DM/Gcal or 8.- DM/t, hard coal

Second approximation

As indicated the empirical results lead to a practically constant proportionality factor ξ between anergy flow and investment cost. This amazingly simple relationship which yields good practical results is unsatisfactory in one respect: if a certain process is improved by better internal re-use of energy a constant ξ would lead to a reduced investment. This does not necessarily always happen, and it was concluded that this contradiction was related to the fact that all plants analyzed belonged to the same generation, having approximately the same degree of technical maturity. Therefore, it was thought necessary to refine the method in order to account for changes in the internal efficiency of processes.

For this the overall efficiency η was split into a theoretical efficiency η_t and a practical efficiency η_p such as

$$\eta = \eta_t \eta_p$$

The theoretical efficiency is given by the thermodynamics of the process under ideal conditions. The practical efficiency indicates to what extent this ideal is approached in practice.

If eq. (2) is expressed for 1 Gcal of end product we obtain

$$\frac{I_s}{\Xi} = I_{su} = \xi \frac{1 - \eta}{\eta}$$

and if we assume ξ to be a function of the practical efficieny but independent of the theoretical efficiency, then

$$I_{su}(\eta_p, \eta_t) = \xi(\eta_p) \frac{1 - \eta_t \eta_p}{\eta_t \eta_p}$$

The function $\xi(\eta_p)$ must fulfil the condition

$$\lim_{\eta_p \to 1} \xi \to \infty$$

because ideal conditions would require, inter alia, infinitely large heat transfer surfaces. In addition the condition

$$\lim_{\eta_p \to 1} \xi \to \text{const.} > 0$$

must be fulfilled because even an extremely poor design of a plant should not lead to zero investment.. Finally, $\xi(\eta_p)$ should be continuous in the interval $0 < \eta_p < 1$.

From the possible functions fulfilling these conditions the expression

$$\xi = a\left(\frac{\eta_p}{1 - \eta_p} + b\right)$$

was selected because of it similarity in structure with eq. (2). Combining eq. (3) and (4) the generalized investment cost function

$$I_{su} \doteq a\left(\frac{\eta_p}{1 - \eta_p} + b\right)\left(\frac{1 - \eta_t \eta_p}{\eta_t \eta_p}\right)$$

Investment cost information on a number of chemical plants for which η_t and η_p was known led to the following values of the parameters a and b giving the best fit (basis : West Germany, 1977):

$$a = 39 \text{ DM/(Gcal/a)}$$

$$b = 0.15$$

The value of a obviously will be affected by the currency used and by the historical time (inflation).

For production cost calculations it is convenient to refer everything to 1 Gcal of final product. Thus

$$K_u = F_u + V_u = za\left(\frac{\eta_p}{1 - \eta_p} + b\right)\left(\frac{1 - \eta_t\eta_p}{\eta_t\eta_p}\right) + \frac{\kappa}{\eta_t\eta_p}$$

where F_u represents the fixed costs per 1 Gcal end product, V_u the variable costs, z is the capital load factor, and κ the cost of the organic raw material per Gcal.

In Figure 1 the investment cost and the production cost corresponding to 1 Gcal of end product for a process having a theoretical thermal efficiency of 70 % have been plotted versus the practical efficiency. Investment and production costs show minima which do not coincide. The position of the production cost minimum changes with rising raw material costs towards higher practical efficiencies, requiring correspondingly higher investment cost.

In Figure 2 the η_p-values of the minima for investment and production costs have been plotted as a function of η_t. Parameters are the raw material costs (κ = 10 and 20 DM/Gcal respectively), whereby the investment cost curve corresponds to raw material costs zero.

It can be seen that at constant η_t the position of tthe minimum moves towards higher values of η_p. This effect is marked for low raw material costs, and diminishes once η_p-values of about 75% have been reached.

By the method presented we feel that a good description is possible of what economic effects an inprovement of the internal, practical efficiency, η_p, can have on a system. Heat recovery and efficiency in general have a clear limit, beyond which the costs increase again. As the position of the cost minimum changes only little with κ , once this situation is reached, either reduction of production costs is practically impossible or such a reduction can only be achieved by increasing η_t, which is equivalent to a change in the fundamentals of the system.

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Symbols

D	dissipated energy, Gcal
$\dot{D}$	dissipated energy per unit of time, Gcal/a
F	fixed costs
H	LHV of product, Gcal/t
K	costs of product
V	variable costs
a	parameter, currency/(Gcal/a)
b	parameter, adimensional
s	subindex; specific value
u	subindex; specific value related to energy
z	capital cost factor, a^{-1}
η	overall thermal efficiency, based on LHV
η_t	theoretical thermal efficiency
η_p	practical efficiency
ξ	investment factor, currency/(Gcal/a)
κ	cost of raw material; currency/Gcal.

Table 1 Determination of ξ Values for Ammonia and Methanol Plants

Product	Raw Material	Capacity t/a	Investment, MM DM (2) 1973	1976	Input, Gcal/t (1) (H_o)	(H_u)	Output Gcal/t (H_u)	Dissipated Energy Gcal/t (H_u)	10^6 Gcal/a	ξ DM/(Gcal/a)
NH_3	Natural Gas	333 333	80	99	8,500	7,633	4,453	3,180	1,060	93
NH_3	Heavy Oil	333 333	135	167	10,000	9,500	4,453	5,047	1,682	99
NH_3	Coal	333 333	175	216	11,600	11,194	4,453	6,741	2,247	96
CH_3OH	Natural Gas	333 333	70	87	8,000	7,200	4,700	2,430	0,810	107
CH_3OH	Heavy Oil	333 333	120	148	9,000	8,550	4,770	3,780	1,260	117
CH_3OH	Coal	333 333	160	198	10,500	10,133	4,770	5,363	1,788	111

(1) H_u/H_o, for natural gas: 0,90
H_u/H_o, for heavy oil: 0,95
H_u/H_o, for coal: 0,965

(2) Increase of investment till 1976 = 9 % + 7 % + 6 %; factor 1,236.

Literature: H. Hiller, "Erdöl und Kohle", Compendium 1974/1975, pp. 268-283.

Table 2 Determination of ε Values for SNG Plants

SNG-Capacity MMM m^3_n/a	HHV of SNG H_o, kcal/m^3_n	LHV of SNG H_u, kcal/m^3_n (1)	Output, 10^6 Gcal/a			Coal, waf (2)		Coal, as used		Coal Input MM t			Input 10^6 Gcal/a	Dissipated energy 10^6 Gcal/a	Investment MM DM (4)	ε
			SNG	Byproducts (3)	Total	HHV H_o, Gcal/t	LHV H_u, Gcal/t	Water + ash %	LHV H_u, Gcal/t	Gasification	Power Plant	Total				DM/(Gcal/a)
2,5	9,097	8,187	20,468	4,962	25,430	7,346	7,089	35,5	4,572	7,17	1,50	8,67	39,639	14,209	1391	98
2,4	9,123	8,211	19,706	5,680	25,386	7,507	7,244	38,0	4,491	7,35	1,50	8,85	39,749	14,363	1391	97
2,4	9,123	8,211	19,706	4,429	24,135	7,066	6,819	33,6	4,528	7,01	1,50	8,51	38,530	14,395	1391	97
2,4	9,123	8,211	19,706	3,516	23,222	6,700	6,466	33,7	4,287	7,25	1,50	8,75	37,508	14,286	1391	97
2,2	9,264	9,264	18,343	5,096	23,439	7,176	6,450	34,7	4,217	6,86	1,50	8,36	35,257	11,818	1391	118

(1) H_u/H_o, for SNG: 0,900

(2) H_u/H_o, for coal: 0,965

(3) LHV (H_u):

Tar and oil	9,0 Gcal/t
Gas naphtha:	10,0 Gcal/t
Crude phenol:	8,6 Gcal/t
NH_3:	4,5 Gcal/t

(4) Increase of investment till 1976:
9 % + 7 % + 6 %, factor: 1,236
Exchange rate: 1 $ = 2,50 DM

Literature: H. Hiller, "Erdöl und Kohle", Compendium 1974/1975, pp. 260-283.

Table 3 Determination of ξ Values for various Chemical Plants

Product	Raw Material	Capacity t/a	Investment D/L MM/DM	Specific Investment (DM/(t/a))	Specific Input Gcal/t	Specific Output Gcal/t	Specific Dissipated Energy Gcal/t	ξ DM/(Gcal/t)
Ethylene	C_3/C_4	680 000	377	554	25,668	20,352	5,316	104
Hydrogen cyanide	CH_4/NH_3	120 000	72,5	604	14,59	8,21	6,38	95
Acetic Acid	CO/CH_3OH	170 000	41	241	6,19	3,77	2,42	100
Maleic Acid	Benzene	60 000	58	967	13,260	3,302	9,958	97
Acetylene	Naphtha	100 000	216	2160	59,44	38,82	20,62	105
Acetylene	CH_4	100 000	216	2160	61,037	36,420	24,617	88
Syngas	Bituminous Coal	$10^{10} m_n^3/a$	1922	$192/1000 m_n^3$	4,766	2,744	2,022	95
Acetic Anhydride	Acetaldehyde	270 000	81,5	302	8,215	4,813	3,402	89
Cumene	Benzene/C_3H_6	250 000	38	152	11,813	10,402	1,411	108
Sasol II	Coal via Fischer-Tropsch	1 600 000 + Tail gas	3000	-	49,35 (1)	19,00 (1)	30,35 (1)	99
Fuels	Coal by hydrogenation	1 300 000	729 (3)	-	20,363 (2)	12,788 (2)	7,575 (2)	96
Electricity	Lignite	5 600 000 (MWh)	926	176	2,500	1,860	1,64	107

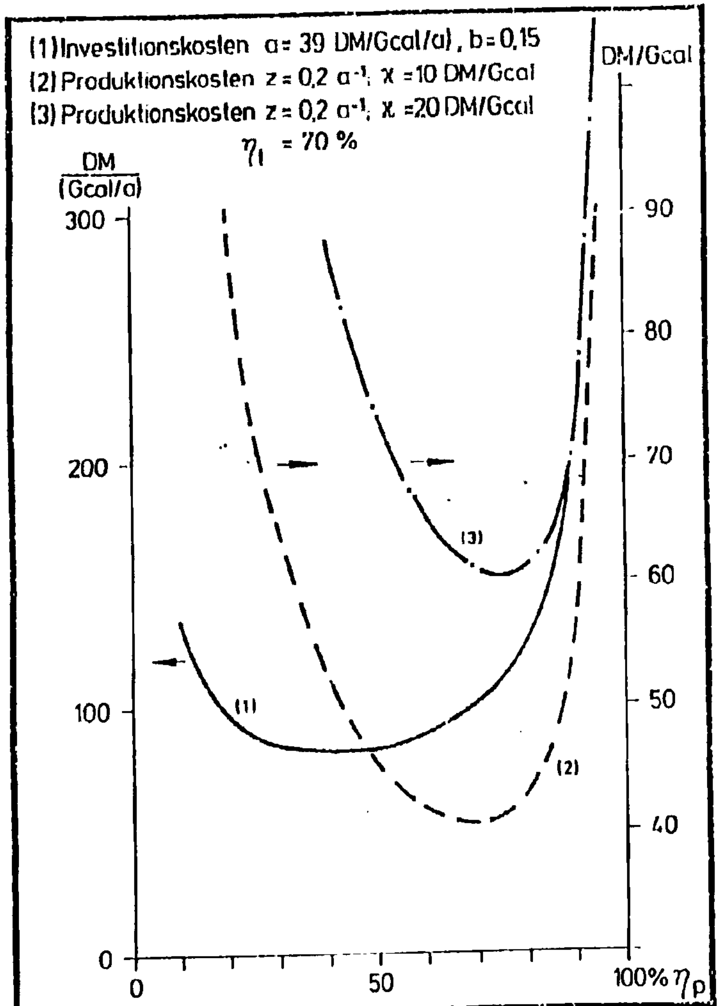
(1) Thermal efficiency of Fischer-Tropsch plant: 38,5 %. Investment cost as per Sasol publication. Uncertainty of investment costs due to varying rates of exchange.

(2) Thermal efficiency of coal hydrogenation: 62,8 %.

(3) According to a study prepared by IGT in 1975 for the Australian Government. The investment costs have been reduced by the value of the auxiliary units and the infrastructure and the balance increased by 23,8 % to cover escalation.

Table 4

Variable	Dimension	Minimum value	Maximum value
Temperature difference	^{o}C	20	50
Heat transfer coefficient	$kcal/m^{2}h\ ^{o}C$	120	800
Number of anergy passes	1	1.0	2.0
Specific cost of heat exchanger (carbon steel)	DM/m^{2}	250	350
Specific cost of heat exchanger (stainless steel)	DM/m^{2}	450	600
Fraction carbon steel	kg/kg	0.90	0.95
Modified Lang factor	1	10	22
Overdesign factor	1	1.05	1.20



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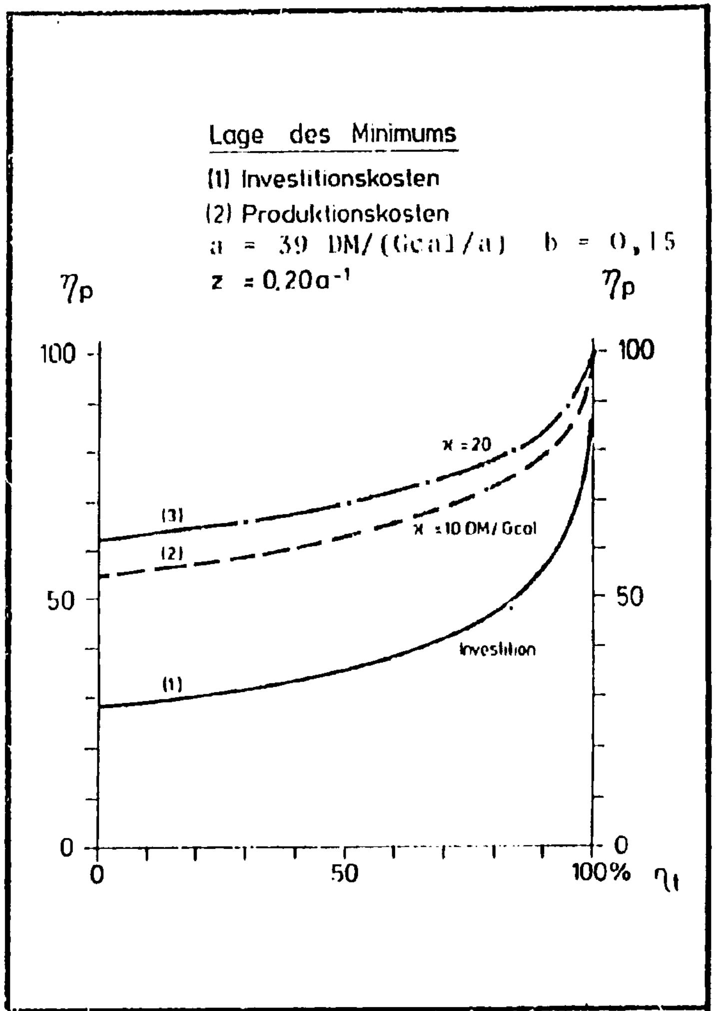

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LECTURE - GAENSSLEN

HOLIGHAUS: Comparing your presentation with that of 1 - 2 years ago, I see that you have again made many new contributions regarding your method. I would like to ask all participants to raise questions.

KNUDSEN: I would like to explore the point that this has been used with much success for chemical processes and the fact that you would like to look at energy processes with this fascinating method. I want to learn more about it, and we would like to begin taking advantage of it in the U.S. Waht would you say about its use for these energy plants? Would you make a distinction between chemicals and energy?

GAENSSLEN: There is actually no clear boundary between energy and chemicals; this is a floating limit. However, when we developed this method we used only chemical plants. Yet, we have used it several times in connection with electricity production, for example, and good results have been obtained within the limitations of the method. One could actually consider the production of electricity from fossil fuels as a kind of chemical process leading to the end product electricity.
At the foot of table 3 the production of electricity using lignite (which was also used as one of the statistical points of the study) is mentioned.

HOLIGHAUS: Did you include the Sasol plant in your study?

GAENSSLEN: Yes, we included this plant in our paper.

HOLIGHAUS: This context can be seen as a major reason for our meeting.

IMHAUSEN: We also used the method in coal liquefaction considerations.

SCHULZE: 1) Waht ist the relationship or distinction between the efficiencies in your calculation compared with the energy efficiency? Which is now considered to be very important in thermo-dynamic calculations? There is also a distinction between overall energy efficiency and energetic efficiency, which has an upper boundary. I think there is a close relationship between your practical efficiency and the energetic efficiency of a process.

2) Is it a complication of your procedure when you wish to apply it to processes in the early of their development, when you have to distinguish between the two efficiencies and when you need much energetic "backing up" of the process. (This may not be available in the early stages of development)

GAENSSLEN: 1) In my manuscript, I also introduced the concept of energy. "Energy" is not a printer's error; nowadays, one is not up to date if one does not include energy in a paper dealing with thermo-dynamics. When it comes to practical applications, not many people in engineering companies and elsewhere are accustomed to calculating with energy and anergy. The simple way of doing this is be saying that the dissipated energy, which is measured by the difference between input and output, is just the anergy. This is in fact true and, moreover, is easy to calculate. If theoretical methods were used, the approach you mentioned would be the right one; however, we are concerned with practical matters and deal with dissipated energy.

2) Complications obviously arise when one splitts the overall efficiency into the practical and the theoretical efficiency. In an early stage of development, it suffices to take the overall efficiency. The overall efficiency is only splitt when one wishes to make refinements. This raises a difficulty; what is the theoretical efficiency of a process?

This question has sometimes no simple answer. When making methanol from methane, the theoretical thermal efficiency is easily calculated from the formula:

Methane + Oxygen = Methanol.

Unfortunately, neither nature nor industry is as simple as that in many cases. It is sometimes very difficult to find out what the thermal theoretical efficiency is. So in some instances, this second approach will remain a kind of theoretical one. Most people will work, as before, with an overall efficiency.

However, when ohne is considering electricity production, for example, the theoretical efficiency can be calculated easily from the maximum and minimum temperatures of the system by the Carnot process. This is a clear-cut situation, in which the practical efficiency of the system is obtained via the theoretical efficiency.

KRÖNIG: I would like to raise a question regarding coal hydrogenation. There are two processes, both of which have the same thermal efficiency. One needs two and a half times as much space room as the other - your problem does not bring this into consideration.

GAENSSLEN: This is true, and we are aware of the difficulty. However, in the case of a large coal hydrogenation plant, the reactor volume is actually a comparatively small part of the total investment.
If the space velocity decreases by a factor of 2 1/2, the overall investment will increase only slightly. In a Fischer-Tropsch-plant for example, the heaviest investment is not in the reactors but in the gas or hydrogen production in the "work up" of the product after recovery. The reactor system itself is not a heavy burden on these systems; therefore I already raised the question 2 or 3 months ago as to whether one should spend so mucht time on designing reactors. Would not this time and effort be better spent in bettering the heavy expenses, which definitely do not lie in the field of reactors.
To refer to your question: the difference will lie within the 20 % error, which the system has in any case.

HOLIGHAUS: I would like to draw your attention to the session this afternoon, when there will be time for a general discussion about today's papers. During this time, we will be able to continue considering Mr. Gaensslen's paper.

KÖLLING: In coal hydrogenation, without considering your system, we do two things:

1) increase the thermal efficiency
2) to lower the conditions and investment costs, i.e. to lower pressure, temperature, etc.

Can you inform us what is primarily necessary, when using your system, for increasing the thermal efficiency or lowering the investment costs?

GAENSSLEN: Both are parallel. As soon as the thermal efficiency is increased, the investment costs always go down. This is the usual tendency, as long as one stays within the normally accepted limits.

For example, more oil can be made from coal if the thermal efficiency is increased. This means that less coal must be ground and treated and that less hydrogen must be produced. These factros tend to decrease the amount of investment needed.

It is always advantageous to increase the thermal efficiency, and when the practical efficiency is also increased, it is good within certain limits. The theoretical efficiency therefore has to be increased, and the practical efficiency must be increased until the cost becomes too high. In any case, improvement in the overall efficiency generally leads to lower investment.

HOLIGHAUS: This question was very well answered. I would now like to proceed to the next paper, and turn from the more basic methods of cost assessment and estimation to concrete estimations of special processes. Dr. Hubert of Schering will now present his paper.

Dr. Hans-Jürgen Hubert September 1978

INVESTIGATIONS ON ECONOMICS OF THE FISCHER-TROPSCH-PROCESS FOR THE SYNTHESIS OF CHEMICAL RAW-MATERIALS

At proper reaction conditions carbon monoxide and hydrogen are converted in the presence of a catalyst to hydrocarbons and oxygen-containing compounds. This reaction is called by the name of its discoverers Fischer-Tropsch-Synthesis. The composition of the resulting products is influenced by the type of catalyst and by the reaction conditions. The products may predominantly consist of hydrocarbons or oxygen-containing compounds. The hydrocarbons may be rich or poor in olefins. The molecules mostly have non-branched carbon chains, the double bands are predominantly in the end position. With the oxygen-containing compounds alcohols predominate and to a lesser extent ketones, aldehydes, fatty acids and esters are formed. The chain length of the molecules are subject to a statistical distribution, however the mean C-number may be varied in a wide range.

The Fischer-Tropsch-synthesis, known in short as the FT-synthesis, has, up to now, mostly been used for the production of petrol for internal-combustion engines. Many of the produced compounds are important chemical raw materials, which are at present processed from crude petroleum.

On behalf of the Bundesministerium für Forschung und Technologie (Federal Ministry for Research and Technology), Schering AG has made a study on the "Synthesis of Raw Materials for the Chemical Industry on the Basis of the Development of the Fischer-Tropsch-Process" (Synthese von Rohstoffen für die chemische Industrie mit Hilfe des weiterzuentwickelnden Fischer-Tropsch-Verfahrens). This study was completed in May 1977.

Within the scope of this study a literature and patent investigation was carried out, determining the current level of know-how. The information is stored in a central archive. By means of a visual punched card index, quick access to the original literature is possible. Our own know-how, which we obtained by running a FT-synthesis at the Bergkamen site of Schering AG up to April 1, 1962 has been filed. This know-how has been updated by information obtained from the SASOL-company in South Africa, in discussions with Professor Kölbel and by accompanying investigations in a continuously working FT-plant on laboratory scale with a feedgas throughput up to 5 Nm^3/h. Besides this the basic engineering for a demonstration pilot-plant with a throughput of 10.000 Nm^3/h synthesis gas has been done. This corresponds to a capacity of 12.000 tons per year of FT-products.

By means of a simulation model comparative analyses on economics for FT- and competitive petrochemical processes have been carried out.

Today I want to report on the structure of this simulation model, the input data and the acquired results.

The following requirements have been set on the simulation model:

1. The processes must be calculable for different running conditions in order to investigate as in how far these conditions influence the economics of the process.

2. The process steps must be interchangeable in order to determine the most favourable process on an economic basis.

3. The processes must be calculable for different plant location dependent coal prices. By this means the influence of the coal price on coal dependent energy and raw materials can be taken into account.

4. The calculation of the processes must be possible for different references years in order to estimate the economy which may be expected. These are only model investigations which presume certain future developments in price.

The basis of the simulation modell is a process structure in which process steps are connected to each other. Fig. 1 shows a simplified FT-synthesis-structure. In this structure process steps are indicated by triangulars, products entering and leaving the process are marked by rectangles and finally product distribution is marked by rhombuses.

The coal is converted in the gasification step to synthesis gas which is processed in the then following synthesis. Gasoline, oils and water are obtained as condensates. The residual gas contains the short-chain-olefins which are scrubbed out in the following scrubber.

A part of the residual gas is recycled, symbolized by the rhombus, to the FT-synthesis. The remainder serves as fuel-gas. The C_2/C_4-olefins, the gasoline and the oil are split up by fractionation into single compounds or fractions. Alcohols and ketones are separated from the reaction water by means of distillation. In this process structure single processes are interchangeable, which means e. g. that the Lurgi-pressure-gasification may be substituted by the Koppers-Totzek-gasification, the FT-synthol-synthesis by the liquid phase-synthesis, a low temperature scrubber by a Cosorb-scrubber. By means of the gas distribution rhombus, gas recycles may be varied as required.

The actual processes involved are much more complex. Without going into more detail one can see such a variable basis structure of a FT-process in Fig. 2. It involves 13 single processes - marked by triangulars - and 20 product distributions - marked by rhombuses. For the calculation of FT-processes four such basis structures have been developed.

Basis for the process calculation are the process steps. The size of each step may be fixed arbitrarily. It may include one or more process steps and may also include auxiliary and subsidiary units.
Each single process is denoted by a number. The data sheet of the process step which is stored in the electronic data processing, contains information on inputs and outputs proportional to the product capacity and on nonproportional values such as basic capacities, demand of capital investment, circulating capital and labour. The basic capacity corresponds to the plant capacity being investigated. When no data for the circulating capital was available, this was taken to be 10 % of the capital investment. The labour demand is understood to be the

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number of operators per shift. The data for most single processes are based on information from production plants, obtained from plant operators and engineering companies.

For example Fig. 3 shows the data sheet of process No. 201.
It is based on the Lurgi-pressure-gasification, which produces a synthesis gas from coal whith a CO/H_2-ratio of 0,35. The basic capacity amounts to 10^{10} Nm^3 hydrogen and carbon monoxide in the synthesis gas per year.

The total plant is of such a size - as can be seen by the example - that for capital investment and labour degression a power of 1 may be taken for calculation. This is valid for all FT-processes. For petrochemical processes, powers of 0,8 for demand of capital investment and Of 0,2 for labour have been used. If not mentionend otherwise in the data sheet, the demand of capital investment is based on "inside battery limits".

In order to evaluate the quantitatively specified inputs and outputs in the data sheet, prices for raw materials and energies have been determined, which are contained in price-lists. The calculation can be performed with different price-lists. When determining prices, which were valid in Western Europe in the middle of 1975, market prices were used where possible. If not, internal prices have been determined. Location and contract specific prices have not been taken into consideration when determining market prices. The following prices are quoted as important key-prices:

- 6 -

bituminuous coal	140,00 DM/t SKE
lignite	70,00 DM/t SKE
crude oil	220,00 DM/t

When the structures are defined and the process data and the prices are known, costs and turnover can be determined by help of the simulation model. The costs dependent on capital are based on fixed percentages per year of the capital investment. These percentages are:

depreciation	10 %
interest	6 %
maintenance % repair	4 %
tax and insurance	2,5 %

The labour costs have been fixed at 40.000 DM per man and year, including all overheads incurred. As the most important economic dimension we took the cost-recovery percentage, which shows how much per cent of the production costs are covered by turnover.

The first basis calculations performed were related to a Westeuropean site in 1975. Such basic calculations were performed for 59 FT-processes

The following parameters were varied:

1. The gasification raw material (bituminuous coal or lignite)
2. the gasification process (Lurgi-pressure-gasification or (Koppers-Totzek-gasification)
3. the CO/H_2-ratio in the synthesis gas
4. the FT-process (fixed-bed-, liquid-phase or Synthol reactor)

5. the C_2/C_4-olefin isolation (low temperature scrubber or Cosorb-scrubber)
6. the gas conversion rate
7. the recycle gas ratio
8. the product range (long-, mediumlong and short chains or as a hypothetical limiting case only ethylene production)
9. In some structures the FT-synthesis was followed by a steam-cracker in order to increase the yield of olefins. Efforts were made to increase either the yield of C_2/C_4 or of C_{10}/C_{18}-olefins, which are the most important raw materials for the production of anionic active tensides.

For a comparison of economics, competitive petrochemical processes were calculated. There are 6 processes for the production of short chain olefins and 5 processes for the production of mediumlong chain olefins.

It is not possible to report the results of all calculations. 4 FT- and 2 petrochemical processes were chosen, which are referred to in the following report.

They are shown in Fig. 4.

The processes chosen for the production of C_2/C_4-olefins are:

1. The FT-Synthol-processes, followed by a steam cracker, feedstock: lignite.
2. same as under 1, but feedstock: bituminuous coal

3. a FT-process producing - hypothetically - only ethylene. This process is certainly not realisable, but represents a threshold limit for the economics of FT-synthesis for the production of low olefins,
   feedstock: lignite

4. a Naphtha cracking process as a competitive petrochemical process.

As processes for the production of C_{10}/C_{18}-olefins were taken:

1. a FT-process giving long chain products, which are cracked to medium long chain olefins in a following steam cracker. The synthesis gas is made from lignite.

2. The Molex-Pacol-Olex-process, were n-alcanes are separated from kerosene, cataytically dehydrogenated to olefins and separated by means of molecular sieves from the reaction mixture.

The basic calculations show, as expected, that in 1975 in Western Europe no FT-process-variant could be run under full cost-recovery. Even if one is able to asume that the FT-synthesis produces ethylene only, the cost-recovery percentage is only a maximum of 75 %. The most favourable of todays possible FT-processes, with a cost-recovery percentage of 67 %, is the process 2.1 for the production of olefins with medium-long chains.
With this process after the steam cracker are 44 % of the primary formed products C_{10}/C_{18}-straight chain hydrocarbons with an olefin content of 90 %. 16 % are C_5/C_9-straight chain olefins, which are suitable as raw material for plasticising alcohols.

The Synthol-synthesis followed by a steam cracker has a cost-recovery-rate of 52 % with lignite as feedstock (process 1.1), which falls to 38 % when bituminuous coal is taken as feedstock. With these processes 62 % of the primary products formed after the steam cracker are C_2/C_4-olefins, which consist of 44 % ethylene, 36 % propylene and 20 % butenes. Without the steam cracker only 22 % of primarily formed products are C_2/C_4-olefins. At present one tries to produce primary products with a higher C_2/C_4-olefin content by using new catalysts.

The cost structures of FT- and petrochemical processes are very different as shown in Fig. 5. The types of costs are given as a percentage of the production costs. The sundry costs, which are essentially energy costs, are balanced against credits for by-products. This counter-balance leads in many cases to negative values. The figure shows that with the FT-process besides the costsof raw materials the capital dependent costs are of considerable importance. This is especially valid for the production of olefin with medium-long chains. With the petrochemical processes the raw material costs dominate. The particularly high credits for the Molex-Pacol-Olex-process can be traced back to n-alcane free kerosene, which is produced as a by-produ

After having finished the basis calculations it was investigated in which manner the economics of FT-processes are dependent on the locatio i. e. dependent on the coal price in a particular area. With it, not only the price of coal, but also the prices of coal price dependent energies and products were varied. Thus the simulation could be performed closer to reality by taking into account the described interdependency.

The results of the simulation for the chosen processes are shown in Fig. 6. Plotted in the figure are coal prices at which cost recovery and parity in costs with the competitive petrochemical processes will be reached.

The hypothetical ethylene process could be run at parity in costs with the naphtha steam cracker process at a coal price of 28,70 DM/t SKE. For the Synthol-synthesis this is not possible until the coal price is as low as 15,40 DM/t SKE, nor is a FT-synthesis for the production of olefins with medium-long chains able to work with full cost recovery even when the price of coal is zero.

This, at first surprising result - in the initial calculation this was the most cost advantageous readily realisable process - can be put down to the high proportion of capital dependent costs (50 %) and to the relatively low proportion of costs of raw materials (23 %). The investigation shows that already today in countries with cheap coal, such as South Africa, Australia and Canada, FT-processes for the production of low olefins are competitive to the naphtha steam cracker process.

The influence of 3 different price developments on the economics of FT-processes were investigated in simulated calculations for a location in Western Europe.

Prices, wages and capital investment were subject to inflationary increases, which are classified into 3 groups. The classification in groups and the different rates of inflation for these groups are shown in Fig. 7.

Simplifying the first group contains all coal price dependent energy and product prices, wages and capital investment. For this group in all cases a yearly increase of 6 % was assumed.

The second group contains all highly on crude oil price dependent energy and product prices. Here the yearly inflation rates are 8,9 or 10 %.

Finally, in the third group are all products, which are less strongly bound to the price of crude oil. The inflation rates are 7, 8 or 9 % and are 1 % lower than the rates in the second group.

The inflation rates for price of energy and products were simply equated either with those of coal or those of crude oil, whereby that the price influencing primary energy is which is able to produce these energies and products more cheaply.

The economic development was extrapolated for a period of 25 years up to the year 2000. The results of the three different simulation-calculations for the processes for the production of low olefins are shown in Fig. 8.

The number at the points of intersection of the curves indicate the year in which the FT-process will be at the break-even-point with the competitive petrochemical process in Western Europe. The solid lines show FT-processes, the dotted lines the naphtha-steam cracker-process. When the inflation rate for the price of crude oil with 8 % per year is only2 % higher than the inflation rate for coal the break-even-point is only reached in the year 1995 by the hypothetical ehtylene process.

With an increase in price for crude oil by 9 % per year the break-even-point with the hypothetical ehtylene process will be reached in 1988, with the technically already feasible Synthol-process with steam cracker in year 1993, with lignite as feedstock for gasification. If the price of crude oil increases by 10 % per year, the break-even-point of FT-synthesis with bituminuous coal as feedstock and naphtha steam cracker will be reached in 1995, whereas this occurs already in 1989 with lignite as feedstock.

The results of the processes for the production of olefins with medium-long chains are shown in Fig. 9.
In order to reach the break-even-point before the year 2000, the price of crude oil must rise at least 3 % faster than that of coal. With an increase of 9 % per year the break-even-point is to be expected in 1999 and at 10 % per year in 1993. A comparison with Fig. 8 shows that FT-processes for the production of low olefins are competitive earlier. As a result of these simulated calculations it may be expected, that FT-processes in Western Europe can be operated economically probably at the end of this century.

Finally one more word on the combination of FT- with coal hydrogenation processes. FT-processes are suitable for the production of low olefins. With the coal hydrogenation BTX-aromatics are formed. By combination of these two processes it is possible to produce olefins and aromatics from coal.

Fig. 10 shows in which manner a product range suitable to demand can be produced by employing both processes. Indicated are the tonnages of particular olefins and BTX-aromatics. The last column shows the consumption of these products in the Federal Republic of Germany in 1976, which amounts to 7,1 million tons. To produce the same tonnage in suitable composition from coal 4,6 million tons would have to be produced with the FT-process and 2,5 million tons by hydrogenation of coal, whereby for the former a Synthol-synthesis was used, followed by a steam cracker, and for the latter a coal hydrogenation process by which the primary product is made aromatic and the remaining alphatic hydrocarbons are converted in a steam cracker to C_2/C_4-olefins.

Fig. 10 shows that the produced and consumed particular chemical raw materials do not differ very much from each other.

I would like to conclude my lecture with the indication, that the product ranges of both processes complement themselves in an almost ideal manner for the supply of chemical raw materials.

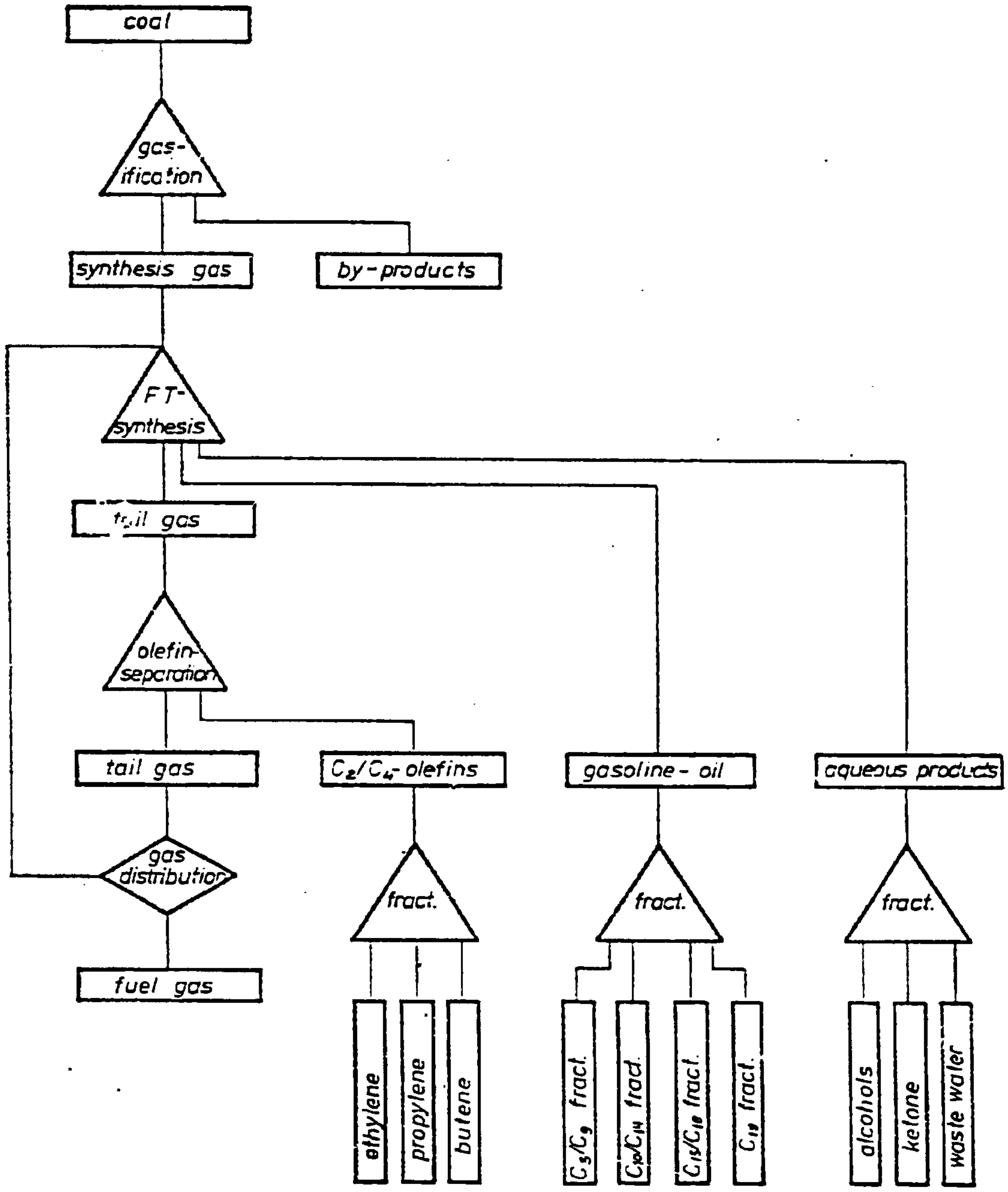


Fig. 1 Simplified structure of a FT-process

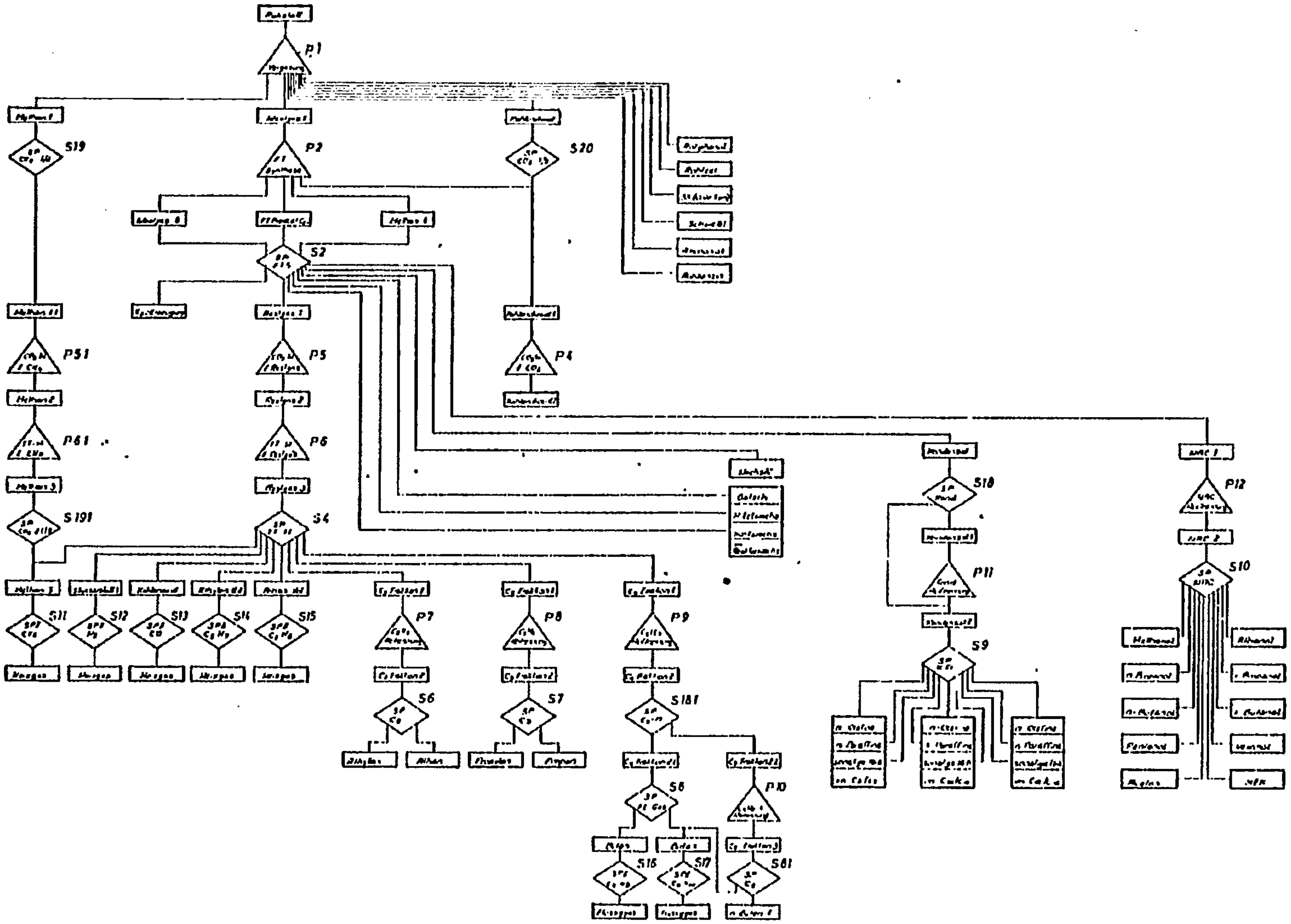
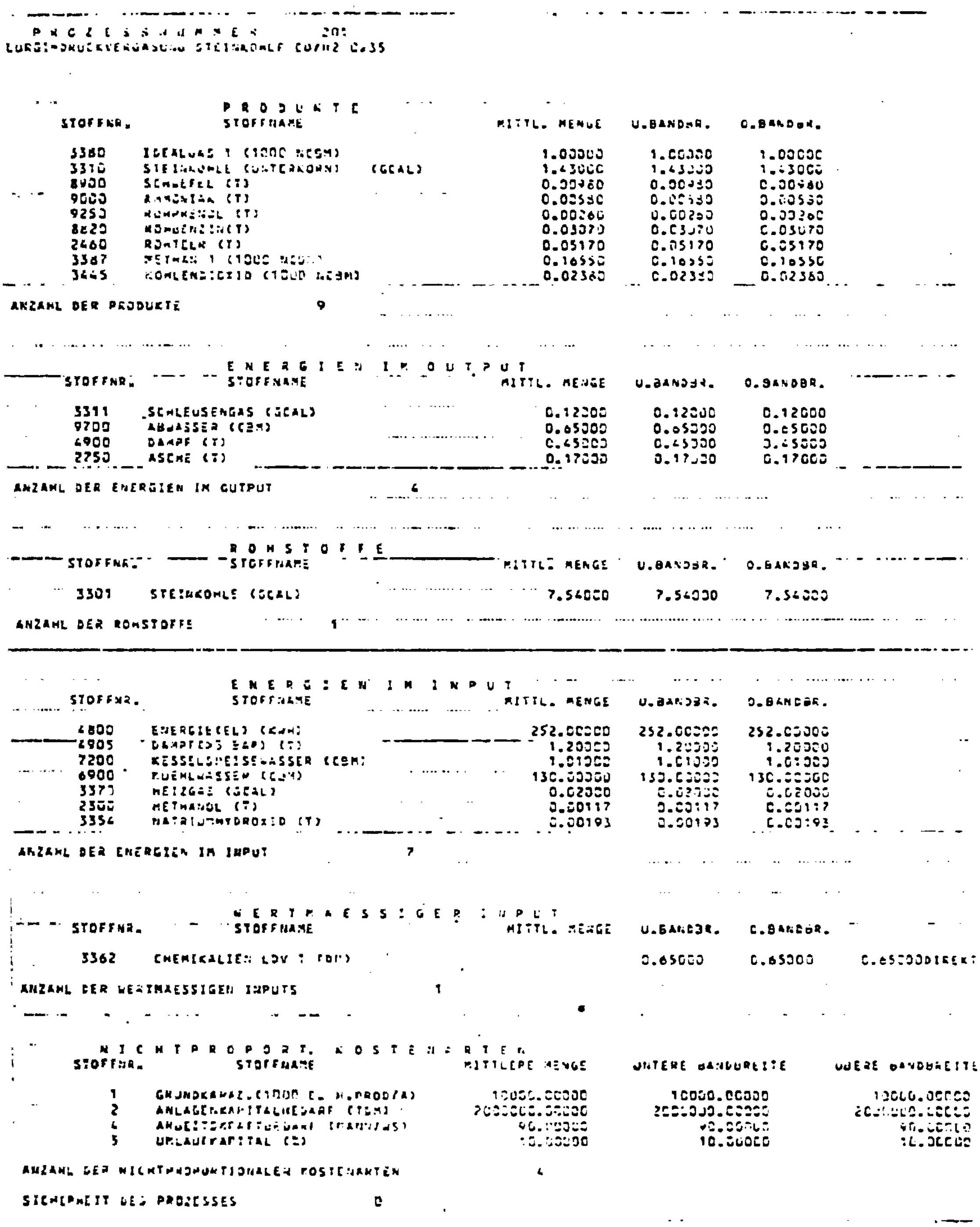


Fig. 2

Basis structure 1



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   P R O Z E S S N U M M E R        201
LURGI-DRUCKVERGASUNG STEINKOHLE CO/H2 0.35


                     P R O D U K T E
    STOFFNR.         STOFFNAME                          MITTL. MENGE   U.BANDBR.    O.BANDBR.

     3380    IDEALGAS 1 (1000 NCBM)                       1.00000      1.00000      1.00000
     3310    STEINKOHLE (UNTERKORN)     (GCAL)            1.43000      1.43000      1.43000
     8900    SCHWEFEL (T)                                 0.00980      0.00980      0.00980
     9000    AMMONIAK (T)                                 0.00530      0.00530      0.00530
     9250    ROHPHENOL (T)                                0.00260      0.00260      0.00260
     8820    ROHBENZIN(T)                                 0.03070      0.03070      0.03070
     2460    ROHTEER (T)                                  0.05170      0.05170      0.05170
     3387    METHAN 1 (1000 NCBM)                         0.16550      0.16550      0.16550
     3445    KOHLENDIOXID (1000 NCBM)                     0.02360      0.02360      0.02360

ANZAHL DER PRODUKTE              9


                     E N E R G I E N   I M   O U T P U T
    STOFFNR.         STOFFNAME                          MITTL. MENGE   U.BANDBR.    O.BANDBR.

     3311    SCHLEUSENGAS (GCAL)                          0.12000      0.12000      0.12000
     9700    ABWASSER (CBM)                               0.65000      0.65000      0.65000
     4900    DAMPF (T)                                    0.45000      0.45000      0.45000
     2750    ASCHE (T)                                    0.17000      0.17000      0.17000

ANZAHL DER ENERGIEN IM OUTPUT              4


                     R O H S T O F F E
    STOFFNR.         STOFFNAME                          MITTL. MENGE   U.BANDBR.    O.BANDBR.

     3301    STEINKOHLE (GCAL)                            7.54000      7.54000      7.54000

ANZAHL DER ROHSTOFFE             1


                     E N E R G I E N   I M   I N P U T
    STOFFNR.         STOFFNAME                          MITTL. MENGE   U.BANDBR.    O.BANDBR.

     4800    ENERGIE(EL) (KWH)                          252.00000    252.00000    252.00000
     4905    DAMPF(35 BAR) (T)                            1.20000      1.20000      1.20000
     7200    KESSELSPEISEWASSER (CBM)                     1.01000      1.01000      1.01000
     6900    KUEHLWASSER (CBM)                          130.00000    130.00000    130.00000
     3370    HEIZGAS (GCAL)                               0.02000      0.02000      0.02000
     2300    METHANOL (T)                                 0.00117      0.00117      0.00117
     3354    NATRIUMHYDROXID (T)                          0.00193      0.00193      0.00193

ANZAHL DER ENERGIEN IM INPUT               7


                     W E R T M A E S S I G E R   I N P U T
    STOFFNR.         STOFFNAME                          MITTL. MENGE   U.BANDBR.    O.BANDBR.

     3362    CHEMIKALIEN LDV 1 (DM)                                    0.65000      0.65000      0.65000DIREKT

ANZAHL DER WERTMAESSIGEN INPUTS              1


           N I C H T P R O P O R T.   K O S T E N A R T E N
    STOFFNR.         STOFFNAME                   MITTLERE MENGE      UNTERE BANDBREITE      OBERE BANDBREITE

        1    GRUNDKAPAZ.(1000 E. H.PROD/A)          10000.00000          10000.00000           10000.00000
        2    ANLAGENKAPITALBEDARF (TDM)           2000000.00000        2000000.00000         2000000.00000
        4    ARBEITSKRAFTBEDARF (MANNJ/WS)             90.00000             90.00000              90.00000
        5    UMLAUFKAPITAL (%)                         10.00000             10.00000              10.00000

ANZAHL DER NICHTPROPORTIONALEN KOSTENARTEN              4

SICHERHEIT DES PROZESSES               0
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Fig. 3 Data sheet of process 201

Lurgi-pressure-gasification

Processes for the production of C_2/C_4-Olefins

1.1 FT-Synthol-process with steam cracker
Raw material: lignite

1.2 The same process but raw material for gasification:
bituminous coal

1.3 Hypothetical FT-ethylene-process.
Raw material: lignite

1.4 Naphtha-steam-cracker

Processes for the production of C_{10}/C_{18}-Olefins

2.1 FT-Synthesis with long chains product range and following steam cracker. Raw material: lignite

2.2 Molex-Pacol-Olex-Process

Fig. 4 Selected Processes for Olefin Production

Product	C_2/C_4-Olefins				C_{10}/C_{18}-Olefins	
Process	Synthol Steam-cracker Lignite	Synthol Steam-cracker Bituminous coal	Hypothetical Ethylene Lignite	Naphtha Steam-cracker	long chains Steam-cracker Lignite	Molex Pacol Olex
Costs of raw materials	54 %	64 %	47 %	95 %	23 %	124 %
Capital depending costs	44 %	33 %	42 %	11 %	50 %	25 %
Labour costs	5 %	4 %	4 %	1 %	8 %	11 %
Sundry costs minus credits	-3 %	-1 %	7 %	-7 %	19 %	-60 %

Cost data as percentage of production costs

Fig. 5 Cost Structures of Processes for Olefin Production

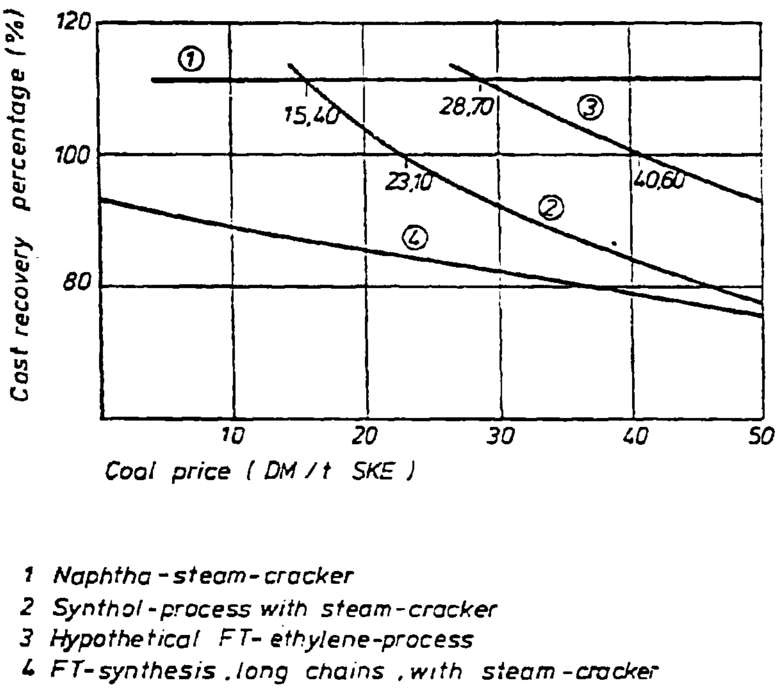
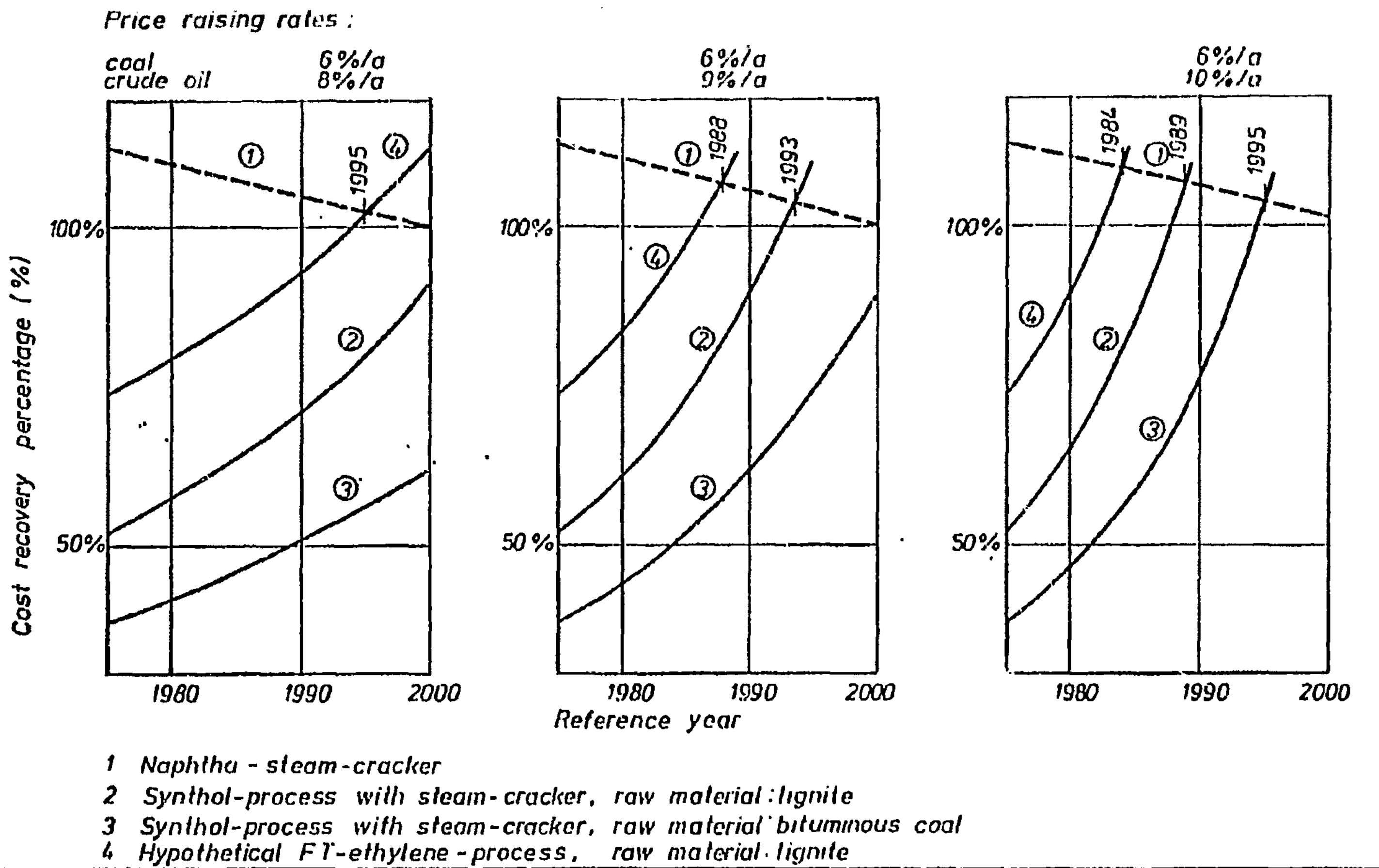


Fig : 6 Dependency of cost recovery percentage on coal price

Price development	1	2	3
Group	Price raising rates (% per year)		
Energies and products dependent on coal price, wages and capital investment	6	6	6
Energies and products highly dependent on crude oil price	8	9	10
Products less highly dependent on crude oil price	7	8	9

Fig. 7 Price and cost raising rates



1 Naphtha - steam-cracker
2 Synthol-process with steam-cracker, raw material : lignite
3 Synthol-process with steam-cracker, raw material bituminous coal
4 Hypothetical FT-ethylene-process, raw material lignite

Fig: 8 Economics on processes for the production of low olefins in western europe from year 1975 to 2000

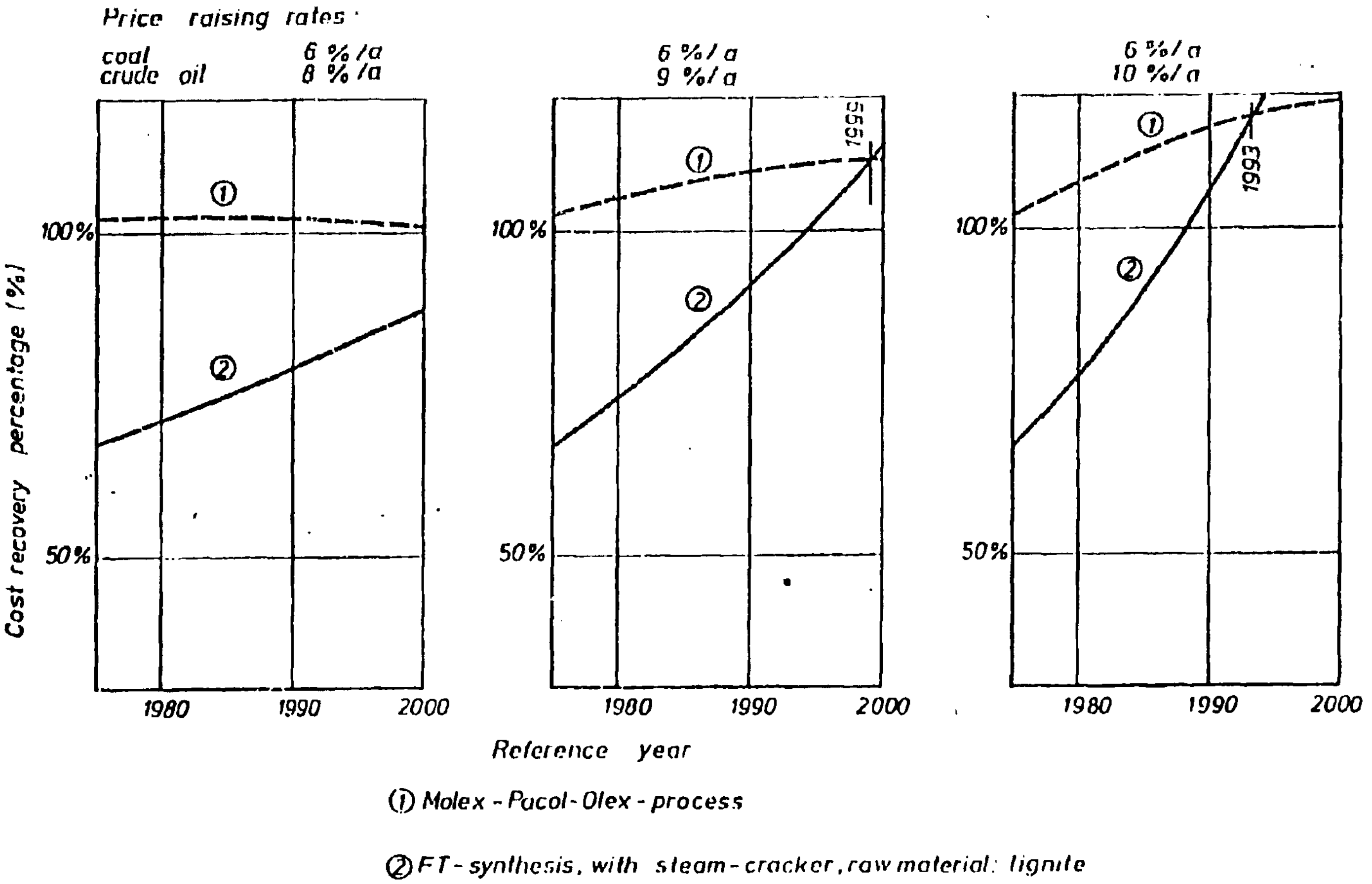


Fig: 9 Economics on processes for the production of C_{10}/C_{18}-olefins in western europe from year 1975 to 2000

Product	FT-Process	Coal hydrogenation	Production	Consumption
Ethylene	$1,9.10^6$ t	$0,6.10^6$ t	$2,5.10^6$ t	$2,8.10^6$ t
Propylene	$1,6.10^6$ t	$0,2.10^6$ t	$1,8.10^6$ t	$1,9.10^6$ t
Butene, Butadiene	$0,9.10^6$ t	$0,1.10^6$ t	$1,0.10^6$ t	$0,5.10^6$ t
BTX-Aromates	$0,2.10^6$ t	$1,6.10^6$ t	$1,8.10^6$ t	$1,8.10^6$ t
Sum	$4,6.10^6$ t	$2,5.10^6$ t	$7,1.10^6$ t	$7,1.10^6$ t

Fig. 10 · Product range of coal-chemical processes adjusted to the consumption of the Federal Republic of Germany in year 1976

LECTURE - HUBERT

HOLIGHAUS: I would like to thank Dr. Hubert and ask participants to raise questions.

ZIEGLER: In figure 3, the bottom line showed the expression "security of the process". What does the figure 0 indicated mean?

HUBERT: The figure is not significant within the context of our present discussion.

GAENSSLEN: I am going to read an extract from a journal and then ask you whether you agree with the opinions expressed.

HUBERT: One should bear in mind that the exchange of oil by coal will start at the thermal end and calorie will be substituted by calorie. If coal is transformed by chemical means, which will result in higher prices on the market, then just the fuel should be chosen. Obviously, ethylene has a higher value than S.N.G. fuels.

IMHAUSEN: One must also take into consideration the fact that the prices of olefins in Europe and the U.S. are very different. They cost considerably less in the U.S. than in the Federal Republic of Germany.

KÜHN: The basic assumption that one has coal and oil in parallel available amounts exists. In Europe, we must take into account the fact that we do not have sufficientoil. We have it at the moment, but no-one can be certain that supplies will last for ever. We are talking about research; a solution for our problems must be found within the next 10 - 20 years. It is not a matter of economics of today, but for tomorrow.

SCHNUR: In Europe, naphta - the main feedstock used to produce lower olefins - is not available in unlimited quantities. Forecasts indicate that there will soon be a shortage in Europe. Thus, higher boiling mineral oil fractions (Dg - Diesel oil range or even higher) will have to be utilized for olefine production.
On employing these feedstocks, the ethylene and propylene yields will be lower causing increase in the production and therefore

SCHULZE: Doubts are often raised as to whether substitution will take place in favour of the chemical industry. What is the constraint limiting the thermal usage of oil? Maybe this is the price boundary or planning measure laid down by the government (saying that mineral oil products must be used only in the chemical industry sector and be saved in the thermal sector)? Because we have, supposedly, a free economy, the price probably will be the regulating indicator of all the different products' uses. In many fields of mineral oil products used today, I doubt whether there is an upper price limit. One need only think of the case of gasoline in the private car sector; people will be prepared to pay 3 - 10 fold increases, resulting in the using up of all available gasoline. Long-term forecasting of automobilisation in Western Europe shows rapid increases; this means that in the future, there will be a very wide usage of gasoline (i.E. naphta), also in the medium time range.
We must get prepared according to these predictions, and consider the situation in communist countries, where people readily pay high prices for gasoline. What is the competition of the chemical industry against this private sector of gasoline usage? We have not investigated this yet, but I doubt whether there is an upper limit. This would mean that the substitution in favour of coal consumption by the chemical industry is not limited by this thermal energy market conditions.

IMHAUSEN: I agree with all your points, except for your remark on the situation in communist countries. There, matters are directed fully by government policy - not by the willingness of people to pay.

SCHULZE: Freedom to choose: whether to pay DM 3 per litre of gasoline or not.

IMHAUSEN: Very few private cars are owned in communist countries.

KÜHN: The basis of comparison for all these coal liquefaction processes is today's naphtha - based petro - chemical industry. Whatever we calculate for on or the other of the coal liquefaction processes, the products resulting from such a process are double

the prices shown by the present situation. Until we arrive at a situation where naphtha doubles in rice, so that, for example, Fischer-Tropsch-products would pay, we should use, for instance, middle destillates instead of naphtha.
Production costs for olefins from middle destillates rather than from naphtha are a little more higher, probability by 10 or 20 %. The next step would be converstion, e.g. hydro-cracking. This would result in an addition of about DM 150. Even this procedure would not double the raw material price, as it certainly would if the Fischer-Tropsch-synthesis were used.

HOLIGHAUS: I would like to introduce the next paper, which deals with the same topic; the discussion can be continued afterwards. I would like to introduce Dr. Schnur, who influenced to a very large extent the R.N.D. energy programme in this field in the Federal Republic of Germany as a member of the Advisory Committee of the Ministry of Research and Technology.

Workshop Kohleverflüssigung
Jülich, 18.-19.09.1978

Dr. F. Schnur

Estimation of economics of a modified Fischer-Tropsch synthesis

This report covers some aspects of economic evaluations as part of a study sponsored by the German Minister for Research and Technology and prepared by Ruhrchemie AG in 1974-1975. The considerations were directed towards the elaboration of concrete aims for the development of a modified Fischer-Tropsch synthesis on basis of price and cost level in 1974.

If the Fischer-Tropsch synthesis based on syngas from coal is considered an alternative to the production of hydrocarbons from mineral oil the question arises to which extent the primary products have to consist of valuable short chain olefins to make the economics of synthesis feasible.

At present time the required organic primary chemicals are almost quantitatively generated from mineral oil products. About 75 % of these primary chemicals consist of non-aromatics, which may in principle be produced via Fischer-Tropsch synthesis. 80 % of these aliphatic hydrocarbons are ethylene and propylene, only small amounts of which are present in the primary products of any version of conventional Fischer-Tropsch synthesis. In the German Federal Republic the annual consumption of C_2 to C_4 olefins amounts to 5 mio t compared to only 300 000 to 400 000 t per year of long chain hydrocarbons by the chemical industry. Generally short chained olefins can be produced by pyrolizing long chain synthesis products provided that extremely cheap coal - not available in the Federal Republic - is fed into the gasification. Even German brown coal at the 1974's costs of 5 DM per Gcal does not fit the premises. Diagram 1 renders in the lower part a survey of market and prices at the time when the study was accomplished.

The upper part shows some general routes of syngas production, of synthesis and of product work up. Though the source of the syngas is in principle unimportant for the synthesis, the combining of gasification and synthesis process has a markable influence on the economics and if there is much methane in the syngas, the question arises whether it is senseful to combine Fischer-Tropsch synthesis and SNG production.

Calculations of economics have been dealing with the following topics:

- the estimated costs of production of primary products of a modified synthesis coupled to a conventional syngas production and related to the costs of coal;
- the estimation of expected revenues of primary products depending on the selectivity of the modified synthesis which is highly influenced by the development of improved catalysts. Mainly ethylene and propylene as primary products are covered by this aspect;
- derived from preceding considerations the selectivity required to equal the costs of production was calculated.

When estimating the costs of production the optimum reaction conditions for possibly highly selctive catalysts had to be taken as unknown. Consequently the costs of production of three well established synthesis processes were calculated in parallel assuming acceptable precision in a range which is given by the different catalysts and synthesis technics used by these processes. It was taken as granted that a modified synthesis would not differ from these processes to an extent which markedly influences the costs of production. However, for reasons of security an extra charge of 1o % was added to capitel costs.

The costs of production were calculated for the three processes: fixed bed synthesis, entrained catalyst synthesis and liquid phase synthesis. Sasol at South Africa operates the first two processes, the latter has proven applicable by Rheinpreußen in Germany.

The study was based on existing data and no engineering detail work was performed. The data available were

- several projects by the working association of Lurgi and Ruhrchemie of 1952 to 1959 directed to the production of motor fuels by fixed bed synthesis with production capacities from 50 000 to 500 000 t per year: these projects were based on the effective costs of units, delivered by Lurgi for the plant of Sasol;

- the study of Pichler of 197o for the Minister of Education and Science covering the production of motor fuels by either fixed bed or by entrained catalyst synthesis with production capacities from 2 to 6 mio t per year;

- the report of Schulze of 1974 in Chemie-Eng.-Techn. dealing with the liquid phase synthesis with a production capacity of 1,5 mio t primary products per year.

All amounts of investment were transferred to the cost level of late 1973 by using the cost indices for chemical plants published regularly by Kölbel and Schulze in the paper "Chemische Industrie". As far as possible investment costs and requirements of energy and labour have been attributed to the corresponding process steps.

Diagramm 2 illustrates the capital requirement versus capacity for a fixed bed synthesis plant derived from preceding estimations. The lines referring to gas production and to synthesis show a steep ascent which is caused by the assumption that an increase in production capacity is accomplished simply by multiplication of units for gasification and synthesis, each having a constant capacity given by the development reached between 195o and 196o. Based on the technical improvements from to day's point of view capital requirements will be lower. On the other hand, the line for auxiliary investment may be too low in view of today because higher investments with respect to pollution regulation and higher social requirements are necessary.

...

Diagramm 3 illustrates the capital requirements for three different synthesis processes based on the sources. The upper line beginning on the left, which is dotted in the range of capacities from o,5 to 2 mio t of motor fuels per year shows the capital needs for the fixed bed synthesis process which have been estimated by using previous project studies from the year 1952-1959. It exceeds by 1o % the values calculated in 197o by Pichler for the production of motor fuels. This fact leads to the conclusion that in the 197o calculations gasification units of enlarged capacity have been assumed.

Following the Piohler calculation the capital requirement for producing motor fuels via the entrained catalyst process is by 2o % higher than for the fixed bed process. Obviously this fact may be attributed to the increased methane production by this process, as methane has to be reformed to syngas subsequently. With respect to the primary product pattern the difference should be lower than the 2o % mentioned.

Below the lines of investments to produce motor fuels there are 3 lines, showing the investment costs for the primary products of the fixed bed process derived from diagramm 2, and a break down of these costs into the costs for syngas production and synthesis, respectively. For a production capacity of 1,5 mio t primary products per year the investment costs are about 1,176 billion marks, where as the investment costs for the liquid phase process are slightly higher as calculated by Schulze.

Diagram 4 shows the main figures emerging from cost estimations based on a conventional terminology for the primary products (C_3+) as well as on the total product range, that is including methane and C_2 hydrocarbons. Following this chart the "working costs" that is understood as costs of production excluding costs of coal, are as an average for the 3 processes estimated to 400 DM per t C_{1+} ± 1o %. Including the costs of coal taken as 5 DM per Gcal the average costs of production are estimated to 545 DM per t C_{1+} ± 7 %.

Diagram 5 explains the importance of the assumptions made for costs depending on capital requirement. These annual costs are ranging from 14,7 % of the capital investment costs (case 1) to 26,4 % (case 6) assuming in interest rate of 8 % p.a. Depreciation as in case 1 and case 2 has been calculated for production units which are

> adapted to a given technology without basic modifications
> generating products of constant quality for a given market
> without change in quality requirements.

These production units may be compared to facilities producing steam, electricity, ammonia, methanol or motor fuels.

The said calculations for costs of production for primary products are based on estimations following case 6. They are comparable with the base taken by Schulze for the liquid phase synthesis (case 5) increased by an extra charge of 1o % for unforseeable cost increases emerging from a modified synthesis. The augmented risk of producing chemical feedstocks from a broad product pattern for a moving market is covered by the formally shorter depreciation time that means an augmented return on investment.

The average costs of production would be reduced by about 12o DM per t C_{1+} assuming a calculation following case 2 but with an extra charge of 1o % for covering the risk.

To find out the estimated revenues on primary products the product pattern depending on varying selectivity of synthesis had to be assumed. Additionally the part of products, which does not consist of ethylene or propylene, had to be classified for usage as chemical feedstock. The remaining products were valued as a pyrolysis feedstock. Maximum and minimum values were determined for the returns as shown in table 6.

The resulting gross revenues were reduced by 6 % for sales costs and possible royalties thus leading to the net revenues. The result of the calculations is shown in diagram 7 with respect to the fixed bed synthesis. The bold line represents the costs of production for primary products by credits for by-products of gasification and depending on costs of coal. For this case the Lurgi pressure gasification was taken. The hatched section below the line represents the possible reduction in cost of production by using units of extended dimensions. This section corresponds to a decrease in the order of 6o DM per t, that is half of the difference mentioned which arises from a calculation following case 2.

The lowest line in diagram 7, which parallels the line of costs of production, represents the costs of coal. Starting with the costs of German hard coal, which amount to DM 18 per Gcal and which are not outlined in the diagram, the primary products had to be charged by DM 325 per t, resulting in costs of production of DM 89o per t of primary product.

The hatched horizontal bands represent the estimated returns, maximum and minimum respectively. They depend on the selectivity of the reaction.

Summing up it may be concluded, that - for costs of coal of DM 5 per Gcal and for a return on investment of 7,5 years - the break even point can be reached if the primary products contain 55 % of ethylene and proyplene at minimum, whereby ethylene as estimated must be favoured by 1,8 in relation to proyplene. As the desired olefin distribution in favour of ethylene is difficult to archieve larger gasification and synthesis units are needed to lower the costs, so that either a reduced selectivity of 4o % becomes sufficient or equal amounts of ethylene and propylene with a selectivity of 5o - 55 % can be tolerated.

The aim for the required improvement of catalyst selectivity would be markedly restricted if cheaper syngas could be made available. As shown in diagram 3 roughly 66 % of the costs of investment for products C_{3+} are covered by syngas production, and roughly 98 % of the costs of coal as the heat of reaction nearly corresponds to the needs of energy for synthesis and product recovery. This relationship is only slightly modified for a selective synthesis with a high yield of ethylene because of the increase in energy need for product recovery.

...

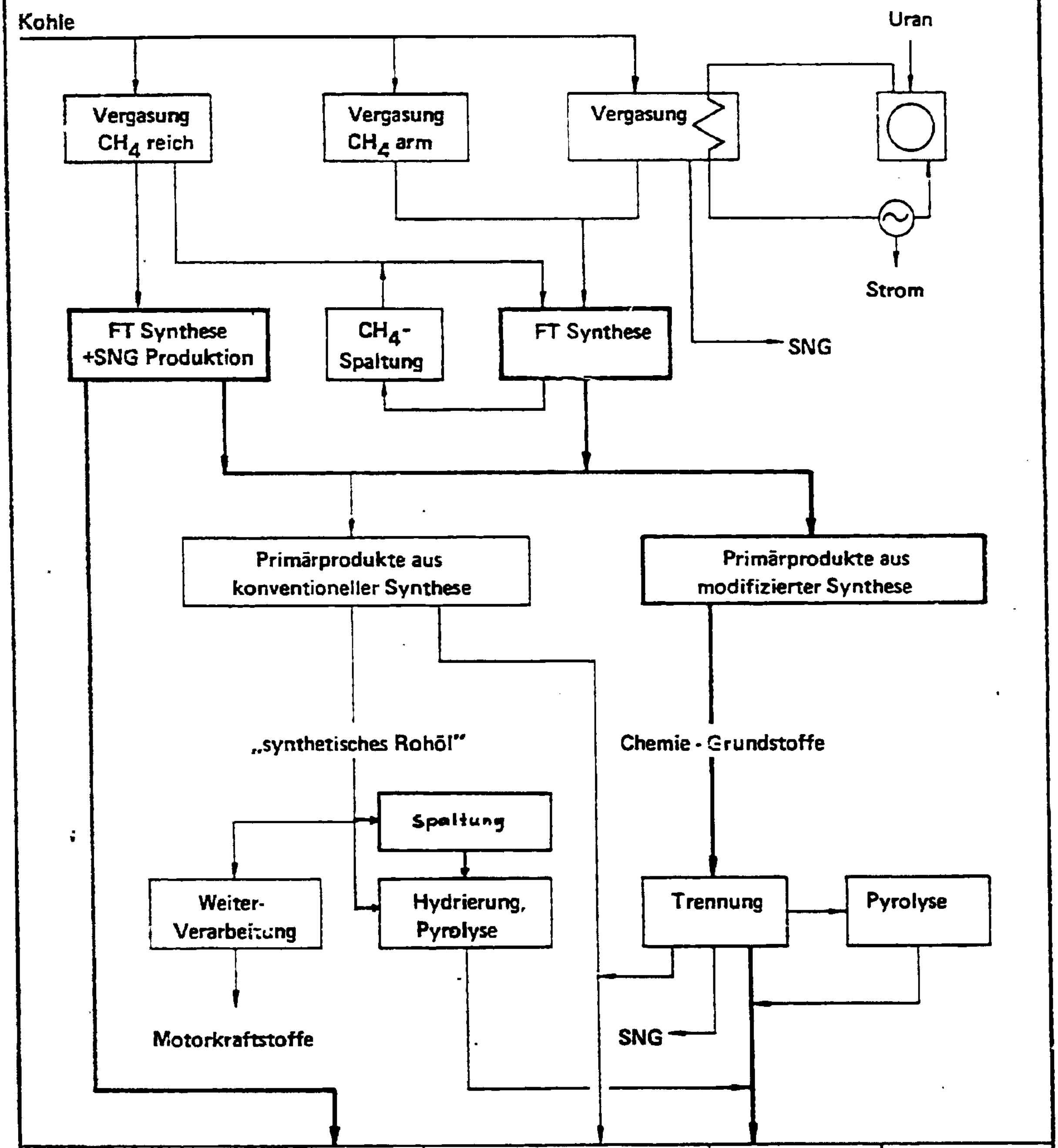
The line marked with an asterisk shows the change in costs of production under said assumptions when syngas produced by nuclear heat is used on the lowest cost level of 1974 forecasts. With cost of coal of DM 5 per Gcal a selectivity of 3o % would be sufficient to cover the costs of production by the estimated returns. However, a selectivity of 55 % would correspond to a coal price of DM 13 per Gcal, thus nearing the range of costs for low quality hard coal in 1974.

A concluding remark may be added with respect to the diversification of the product program of a modified synthesis. If the gasification process renders syngas rich in methane it is apparently advantageous to sell the methane as SNG instead of reforming it including the losses of heat which is needed. From the matter of total costs of production it is estimated that about DM 5o per t of primary product may be saved in case that methane can be sold as SNG to costs of production.

Otherwise a concept may be of advantage which comprises the production of short chain olefins in comparatively small reactors run in parallel if cheap syngas from a huge gasification unit can be made available. Besides special chemicals of high specific value - for instance high boiling paraffins - may be produced at low costs so that the economics of the whole process are favoured.

Fig. 1

Fischer - Tropsch - Synthese als Lieferant von Chemiegrundstoffen



	SNG	aliphatische Chemie-Grundstoffe ∝ Olefine C_{5+} flüssige, gesätigte KW, Paraffine, Alkohole	$C_2 - C_4$ Olefine	Erdöl-naphtha für Chemie-einsatz
Bedarf BRD 1973 (10^3 t)	~ 25 000	~ 300 – 400	~ 5 000	13 000
Richtpreise 1974 (DM/Gcal)	30,–	60 — 90	50 — 70	~ 25

Fig. 2: Schätzungsweiser Anlagekapitalbedarf für die Haupt-Anlagegruppen eines FT-Treibstoffwerkes bei Anwendung des FT-Festbettverfahrens

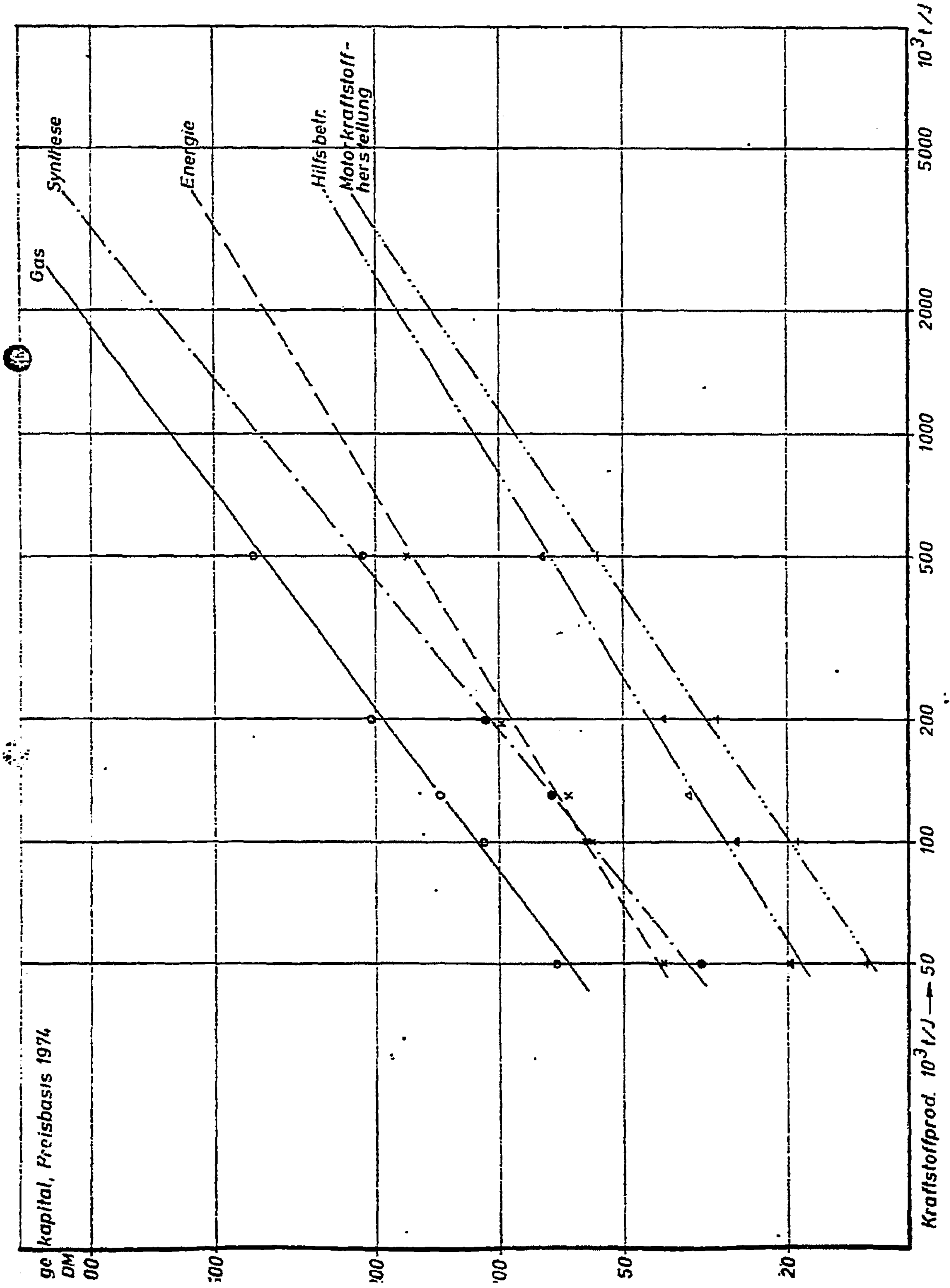
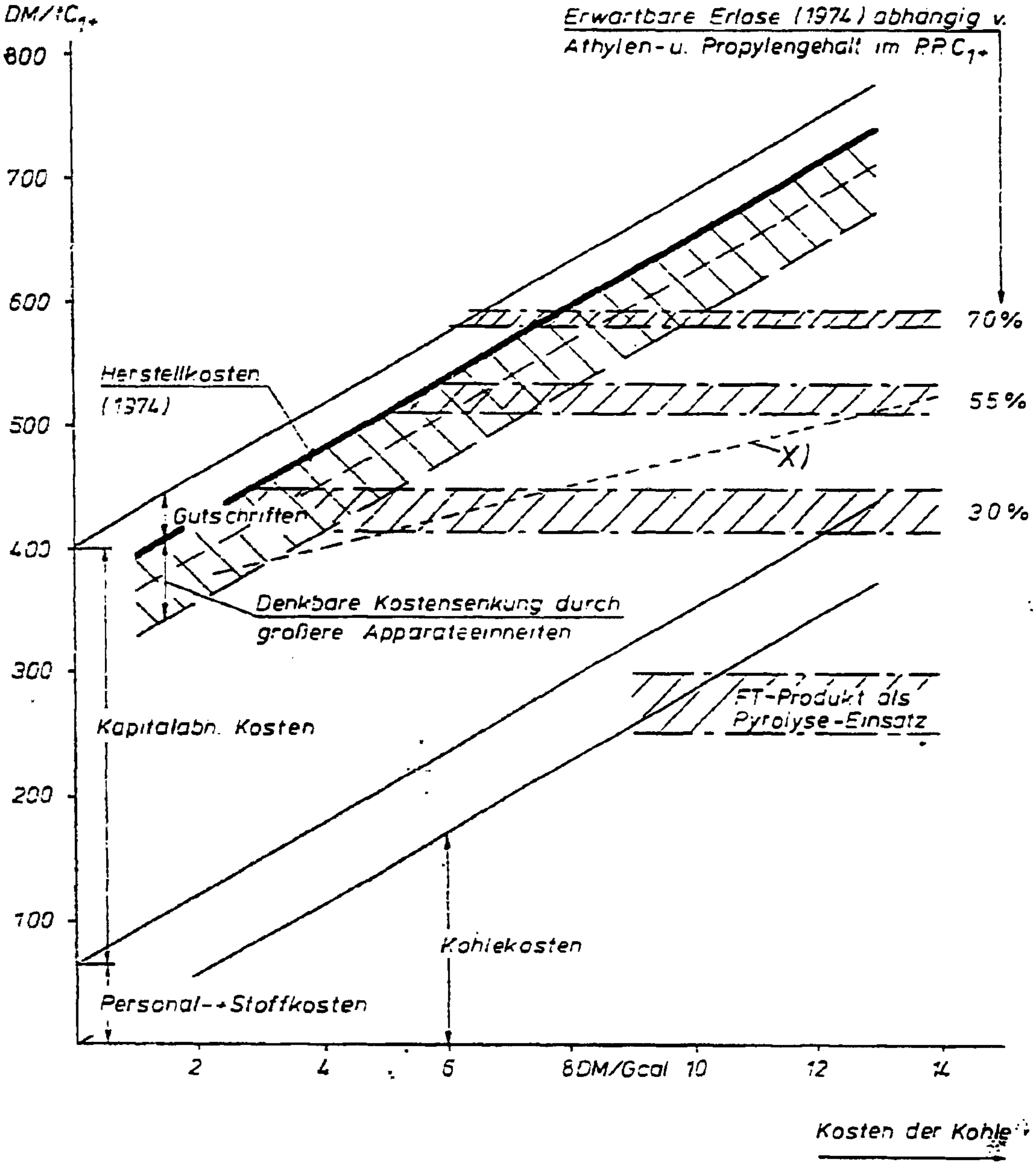


Fig. 7:

Erwartbare Herstellkosten und Erlöse bei einer modifizierten Fischer-Tropsch-Synthese für 1,5 Mio jato C_{1+}

(Konventionelle Kohlevergasung, Festbett-Synthese)



X) = Herstellkosten bei günstigsten Angaben über Synthesegas aus Kohlevergasung im Verbund mit Kernenergie

AFTERNOON SESSION

HOLIGHAUS: I would like to proceed to this afternoon's first paper, and introduce Dr. Kühn, who is a member of the board at U.K. Wesseling, in the Federal Republic of Germany, and to ask him to commence with his presentation.

Wesseling, 29.08.1978

"Chances for Production of Basic or Intermediate Chemical Products via Fischer-Tropsch Process"

With support from the Federal Ministry for Research and Technology my company, the Union Rheinische Braunkohlen Kraftstoff, prepared a study to determine under what conditions the Fischer-Tropsch synthesis could be suitable, not only for the production of motor fuel, as has been done frequently in the past, but also for the production of raw- and basic materials, particularly low-molecular olefins. This study examined all available data previously prepared by engineering colleagues, industrial companies and special institutions.

The technical level of the various processes reviewed varied widely among one another. Frequently the necessary data for an economic consideration were not available. In such case, it was necessary to project analogies from information obtained from the Sasol Corp. based on its experience with two special Fischer-Tropsch processes, Kellog and Arge.

The syntheses processes can be grouped together according to the process technology:

1. Processes with a fixed bed catalyst in the gas phase
2. Processes in the gas phase with moving catalyst
3. Processes carried out in the liquid phase

Atlow temperature, moderate methane formation and high yields of heavier hydrocarbons are the common feature of the processes with a fixed bed catalyst in the gas phase. The content of oxygen-containing compounds in the synthesis effluent varies.

The syntheses with moving catalyst require higher temperatures thus promoting methane formation. The yield of heavier hydrocarbons is lower in favor of hydrocarbons with medium boiling range. The hydrocarbon fractions are richer in olefins. The proportion of oxygen-containing compounds is relatively high.

The lowest methane formation is obtained with the processes in liquid phase although the temperature needs to be higher than in gas phase reaction with fixed bed catalyst. The products obtained are always of higher molecular composition than in the gas phase synthesis with a moving catalyst, but of lower molecular composition than in the gas phase with a fixed bed catalyst.

More than 2o synthesis processes were compared. Processes not yet proven in large scale applied techniques were considered as long as a chance could be seen to develop the technology to a larger technical maturity within a period of 5 to 8 years. Slide 1 and 2 show the processes considered. Dia 1 Dia 2

The comparison led to a continued consideration of a selection of 12 processes.

The direct synthesis of ethylene was not selected, although since 1945 it has been claimed in several patents that ethylene can be produced with high selectivity directly from carbon monoxide and hydrogen. But the prospect for developing an industrial process on the basis of these patents are looking rather poor.

The selected processes and the working conditions relating to them

Dia 3 are shown in slide 3. All processes are carried out with iron catalysts at medium pressure and temperatures. The yield of C_2+ hydrocarbons varies between 125 and 197 g/Nm^3 of processed fresh gas.

For carrying out a comparison of profitability the following assumptions were made for all processes:

1. Production of 2.5 million tons-a-year of C_2+ -products, including their oxygen-containing compounds.
2. Fresh gas contains oxygen and carbon monoxide in an optimal ratio for the process determined.
3. The residual gas is fed back to the synthesis. If necessary, the methane content of the residual gas will be reduced to an acceptable concentration by steam reforming.

The Fischer-Tropsch synthesis is often only considered to be the reaction of carbon monoxide and hydrogen. The recovery of salable pure products from the synthesis effluent mixture is at least equally complicated and costly as the synthesis and is generally underestimated.

Depending on the type of synthesis, the derived products very greatly in quantity and composition.

Dia 4 Slide 4 shows the production of C_2+ -hydrocarbons and oxygen-containing compounds. The selectivity of the individual processes can be clearly recognized.

The Synthol L process and the Hot-Gas-Recycle Process 2o A show the largest portion of $C_2 - C_4$ hydrocarbons.

The fluidized bed synthesis with nitrided iron catalyst produces a spectrum with the greatest preference for oxygen-containing compounds and relatively high $C_2 - C_4$ hydrocarbons, while the Arge process produces mainly C_5+ -hydrocarbons. The distribution of products of the other synthesis processes lies within the limits of these borderline cases.

Dia 5 Slide 5 shows a more detailed breakdown of the $C_2 - C_4$ fraction. The $C_2 - C_4$ hydrocarbons of the Rheinpreussen Slurry Phase S process contain an olefin/paraffin ratio of 2.3 : 1. Further processes with a high olefin to paraffin ratio are Synthol N, Rheinpreussen Slurry Phase L and Hydrocol with 2.2 : 1. All other processes lie under these values. The highest quantity of olefins is produced by the Hydrocol and the Synthol N processes. I would like to point particularly to the $C_2 - C_4$ fraction if the Hot Gas-Recycle process 2b A, which is mainly paraffinic. This process produces, above all, low-molecular paraffinic hydrocarbons.

The composition of the oil phase of the selected processes is shown

Dia 6 in slide 6. One may see that the Rheinpreussen-Koppers Slurry processes S and L produce the greatest quantity of naphtha. Next come the Synthol N and Hydrocol processes.

Dia 7 Slide 7 shows the distribution of the alcohols. As already stated, the fluidized bed process with nitrided iron catalyst produces the greatest portion of alcohols, particularly ethanole.

The working parameter exercise a certain influence on the selectivity of the known processes. However, it should not be overestimated.

For the recovery of the primary products, it was further assumed that

- all final products leave the plant in a quality that is marketable.
- only those products which could be absorbed by the market can be considered as end products. A 50 per cent share of the current EG market was taken as the potential market basis.

The cost analysis of this study is based on the following calculation model:

Price of coal:	5o cts/million BTU = DM/Gcal
energy:	on the basis of the same coal price
depreciation:	15 years
capital:	7o % borrowed 3o % equity
interest:	8 % on borrowed capital 12 % on equity capital

price basis 1975 without inflation rate.

177 - 150 !

New prices for power and inflationary influences obviously cause higher gas prices, but on the other hand there may come new developments that reduce costs, such as better gasifying processes and the introduction of nuclear energy, which were not taken into account.

Dia 6 Slide 6 contains an investment breakdown for the different processes, given for a 2.5 million ton-a-year plant, splitted into coal gasification, synthesis, product recovery and power plant. Total investment is between o,8 and 1,2 billion $ per million tons-a-year capacity. In general, one can say that the total investment distributes to one third each on gasification, power station and synthesis + recovery.

Dia 9 Slide 9 shows a breakdown of the production costs, splitted into proportional shares for coal, power, capital service, labor and some general costs. The biggest share falls to capital service.

Dia 10 Slide 10 gives the share of different product groups in relation to the total production and today's total return. From comparison of the columns "Per cent of Productions" and "Per cent of Revenue" shows whether the revenue from a given product group stands above or below the average revenue from all products. On the basis of current revenue, it can be deducted that the aim should be a yield maximization towards detergent-ra. material and alcohols rather than towards naphtha and low-molecular olefins.

Dia 11 Slide 11 shows the ethylene yield obtainable with the selected processes. The first column shows the original ethylene quantities contained in the synthesis raw product. The figures in the next columns demonstrate, how much ethylene can be obtained by further pyrolysis of suitable Fischer-Tropsch byproducts.

In total the highest ethylene yield of 1.1 million tons-a-year can be expected from the Hot gas-Recycle process 2c A, that is 43 % of the plant capacity.

Dia 12 Slide 12 finally shows the operating costs of a 2.5 million tons-a-year plant, as well as the minimum return per ton of synthesis product and on the other hand the obtainable return on the basis of today's market prices. The column "Revenue required" shows that eight processes require a revenue of $ 242 to $ 288 per ton. More than $ 300 per ton minimum return requirement was calculated for such processes that need high gas circulation rates, because of lower conversion per pass and with expensive CO_2 washing systems within the circuit.

However, the average minimum returns mentioned will change when additional costs for further development are charged to those processes, which are still underdeveloped.

There is no easy answer as to which process may be considered the most economic one at the time. There is one criterion given by the quotient: Revenue required over sales revenue, which is reproduced in the last column. It is clear that the oil circulation process with steel turnings with a quotient of 1.25, followed by the Lurgi Multistage Reactor with a quotient of 1.47, seem to be the most favorable. The average market price of the products must increase by 25 % to 47 %, with operating costs remaining constant, if the processes are to come into a profitable zone.

The chances of several processes are substantially limited because of the small market for some of their products. To consider a combination of two processes with different product pattern in order to obtain optimal revenue is therefore quite natural.

To summarize:

1. Due to the relationship between costs and product revenue, the Fischer-Tropsch synthesis is considered to be uneconomic for Europe today. However, since 1970, this relationship has decreased from 4:1 to 3:1 and even lower. This of course raises the question: how expensive must the petrochemical raw-material naphtha become to make Fischer-Tropsch material competitive?

In a first approximation, this question is answered by the above mentioned relationship between "Revenue Required" and "Sales Revenue",

Within our to-day naphtha prices, capital costs are practically negligible. On the other hand, they come to about 75 % in that naphtha which is derived from the Fischer Tropsch route. In so far such that naphtha is less dependent on raw material than the petrochemical one. Favorable as this may sound, it must be borne in mind that today's prices for crude products are not cost-bearing. So that even the prices of petrochemical naphtha and Fischer-Tropsch naphtha became balanced, the unrationally high capital dependance of a Fischer-Tropsch plant would always bear a high risk, since a competitive product derived from oil can be priced lower at any time as long as oil is available.

As an alternative to coal, natural gas could of course be used for synthesis gas production. The same results with coal at 50 cts/million BTU are obtained with natural gas at 110 cts/ million BTU. So, inspite of the considerable lower investment costs, natural gas cannot be considered as a competitive raw material. DM/Gcal

2. As far as we heve found in 1975,
the prospect seems minimal for directing the synthesis for chemical raw materials in a considerably more selective manner, since the influence of the reaction parameters on the product selection appears to be very limited in all processes.

3. The syntheses employed in South Africa for the production of motor fuel and solid paraffin are relatively favorable by comparison. The seemingly better processes, such as the oil circuit with iron chips or the Lurgi Multistage Reactor are still technically underdeveloped, but their further development should be encouraged.

4. By the end of 1975 no substantially improved catalysts had come to light.

5. A characteristic common to all processes is that they produce greater or less quantities of undesired by-products. In the more extended use of the Fischer-Tropsch synthesis a definite limit of marketability is soon encountered.

6. New tasks are given to research activities, not only with the synthesis, but also with the different stages of product recovery. Two questions that immediately come to mind are: At what degree of yield can Fischer-Tropsch naphtha be pyrolysed to ethylene and how high can the boiling point of the paraffinic Fischer-Tropsch naphtha be raised without impairing the capacity of a steam cracker?

7. In order to improve the profitability of the Fischer-Tropsch process, new developments in coal gasification processes, particularly gasification with nuclear power, might be usefully employed. We cannot expect to see a lowering of the cost of the Fischer-Tropsch process itself or of the product recovery.

FISCHER-TROPSCH-TECHNOLOGIES

1. PROCESSES COMMERCIALLY EXPERIENCED

- SYNTHESIS ON COBALT CATALYST WITHOUT OR UNDER MEDIUM PRESSURE	(RUHRCHEMIE)
- ARGE PROCESS	(RUHRCHEMIE/LURGI)
- SYNTHOL PROCESS N	(KELLOGG, SASOL)
- SYNTHOL PROCESS L	(SASOL)
- HYDROCOL PROCESS	(HYDROCARBON RESEARCH INC.) -

2. PROCESSES TESTED IN DEMONSTATION PLANTS

- SLURRY PHASE PROCESS B	(RHEINPREUSSEN-KOPPERS)
- SLURRY PHASE PROCESS L	(RHEINPREUSSEN-KOPPERS)
- OIL CIRCULATION PROCESS USING NITRIDED FUSED-IRON CATALYST	(BUREAU OF MINES)
- MICHAEL-PROCESS	(BASF)

Fischer-Tropsch-Technologies

(continued)

3. Syntheses tested in Pilot Plants or Laboratory

- Lurgi Multistage Reactor	(Lurgi)
- Hot-Gas-Recycle-Process 2o A	(Bureau of Mines)
- Hot-Gas-Recycle-Process 7 C	(Bureau of Mines)
- Synthesis with a fluidized, nitrided catalyst	(Bureau of Mines)
- Foam Process	(BASF)
- Oil Circulation Process on steel turnings	(Bureau of Mines)
- Synol Synthesis	(Ammoniakwerk Merseburg 1940)
- Iso Synthesis	(Kaiser-Wilhelm-Institut 1948)
- Slurry Synthesis	(Bureau of Mines)
- Ethylene Synthesis from carbon monoxide and hydrogen	(various patents)

Operating Conditions of Selected Processes

rocess	Type	Catalyst	Pressure PSIG	Temp °C	Fresh Feed H_2:CO	Recycle Ratio	Yield of C_2+-Prod g/CBM Fresh Feed
Arge	Fixed Bed	Precipitated Iron	284	220	2:1	2:1	197
Synthol H	Entrained Fluidized Bed	Fused Iron	284	320	2:1	2:1	172
Synthol L	Entrained Fluidized Bed	Fused Iron	284	320	2:1	2:1	125
Hydrocol	Fluidized Bed	Fused Iron	355	320	2:1	3,5:1	182
Rheinpreussen-Koppers B	Slurry Phase	Precipitated Iron	171	268	0,7:1	no recycle	172
Rheinpreussen-Koppers L	Slurry Phase	Precipitated Iron	171	268	0,7:1	no recycle	?
Bureau of Mines	Oil-Circulation	Fused Iron (nitrided)	384	290	0,7:1	2,5:1	178
Lurgi-Multistage Reactor	Fixed Bed	Precipitated Iron	284	274	2:1	20:1	154
Hot-Gas-Recycle 20A	Fixed Bed	Steel Turnings	384	278	1:1	51:1	143
Hot-Gas-Recycle 7C	Fixed Bed	Steel Turnings	384	280	1:1	21:1	129
Bureau of Mines	Fluidized Bed	Fused Iron (nitrided)	299	251	1:1	9:1	185
Bureau of Mines	Oil-Circulation	Steel Turnings	299	270	1:1	1:1	138

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Fig. 3

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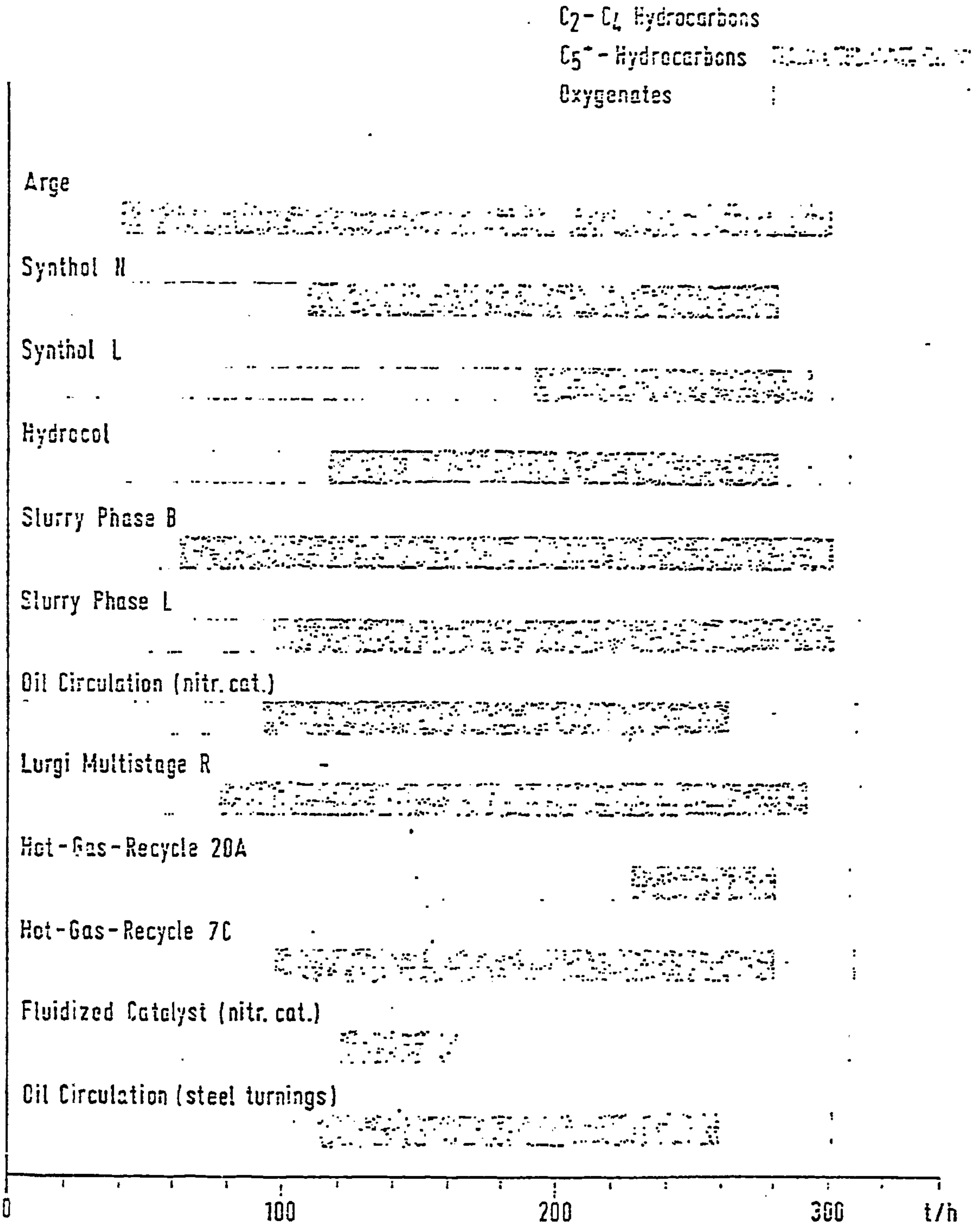
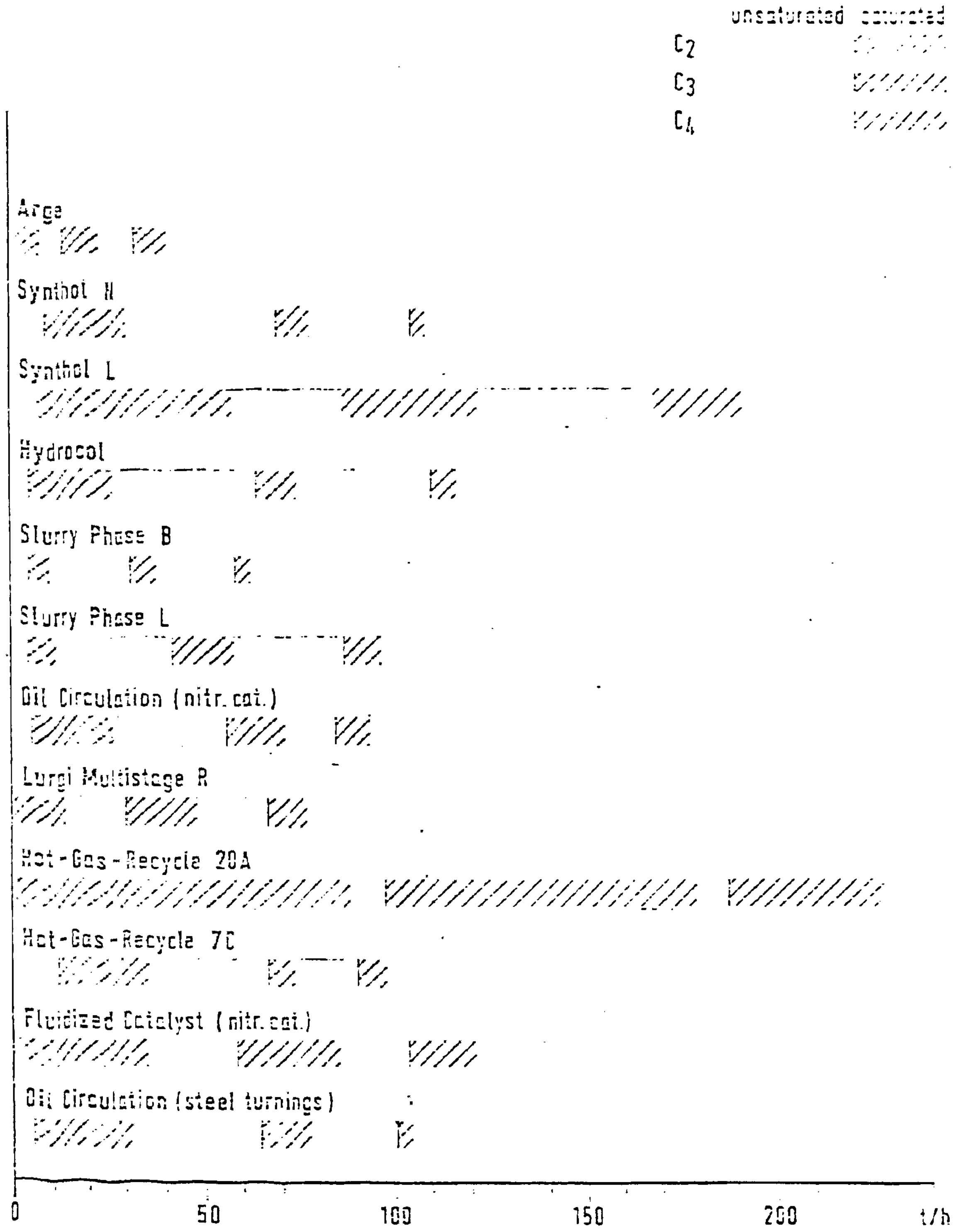
C_2^+-Hydrocarbons and Oxygenates Production



Union Kraftstoff Fig. 4 April 1976

Light Hydrocarbons Distribution



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Oily Phase Distribution

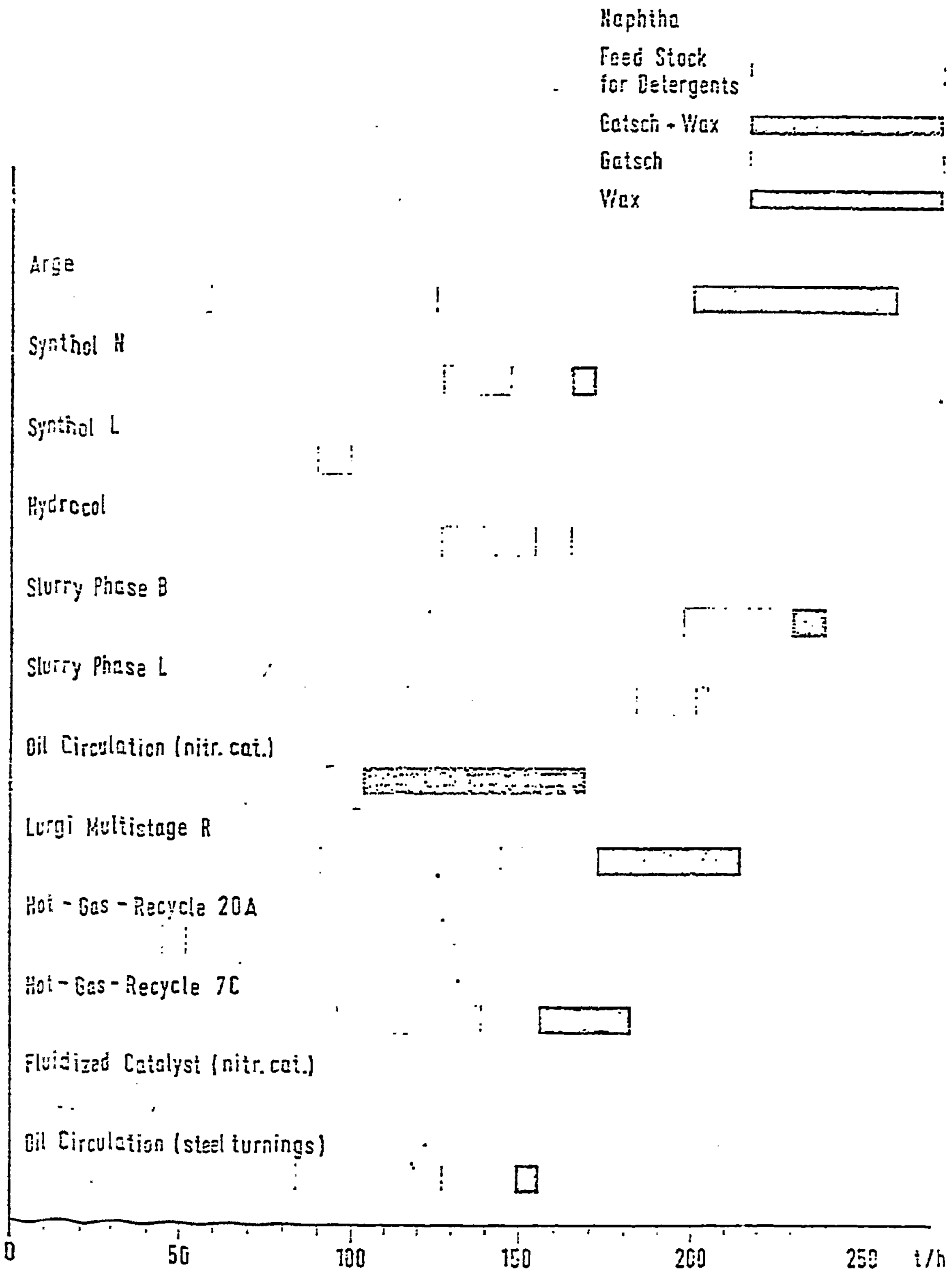

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Fig. 6

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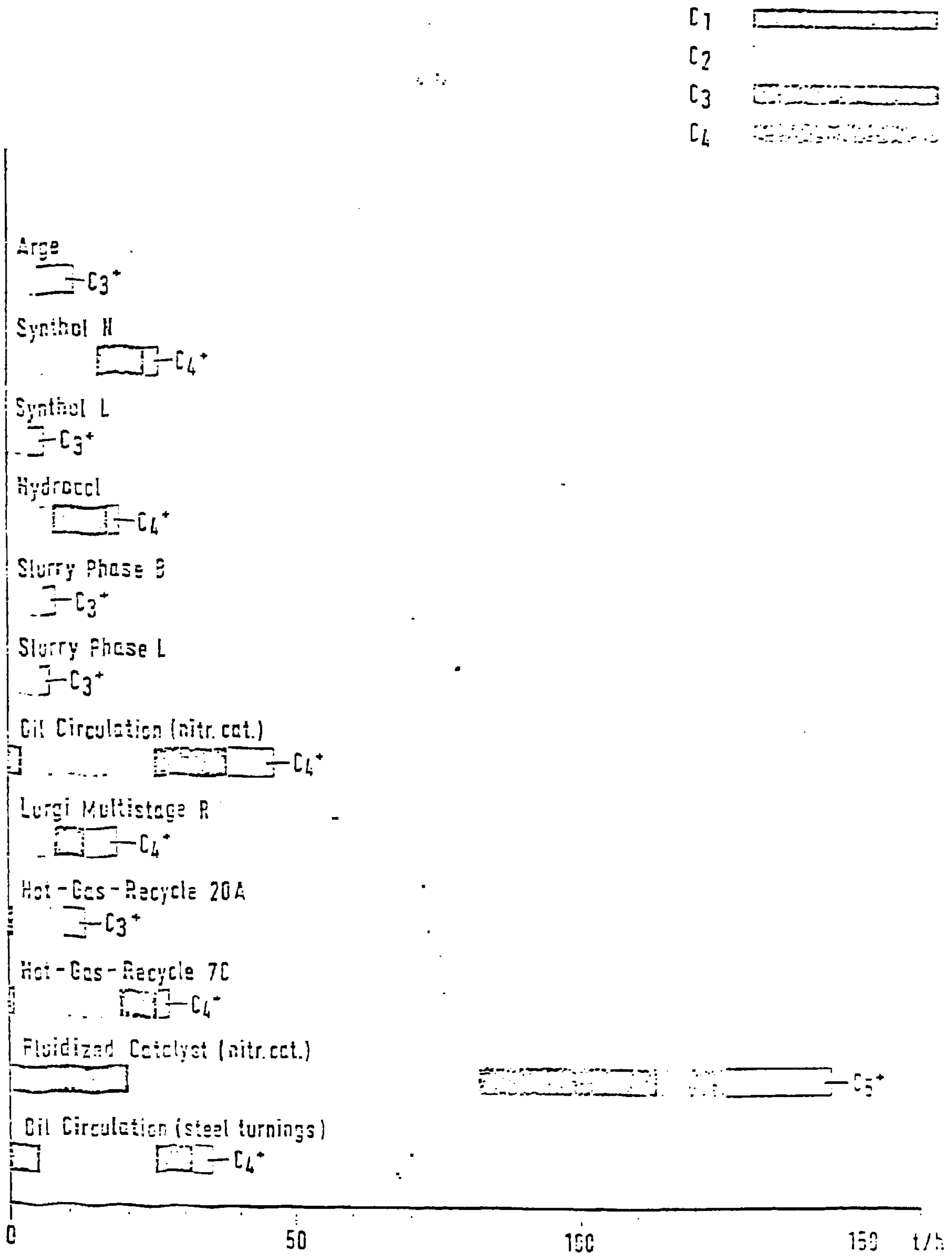
Alcohol Distribution



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INVESTMENT BREAKDOWN SPLITTED INTO GASIFICATION, SYNTHESIS, PRODUCT RECOVERY AND POWER PLANT

PROCESSES	TOTAL INVESTMENT BILLION $	GASIFICATION		SYNTHESIS		PRODUCT RECOVERY		POWER PLANT	
		BILL.$	%	BILL.$	%	BILL.$	%	BILL.$	%
ARGE	2,21	o,93	42,1	o,57	25,7	o,23	1o,2	o,48	22,o
SYNTHOL N	2,o8	o,87	41,6	o,51	24,5	o,17	8,1	o,53	25,8
SYNTHOL L	2,17	o,8o	37,1	o,63	28,9	o,27	12,4	o,47	21,6
HYDROCOL	2,12	o,91	42,7	o,47	22,1	o,15	7,2	o,59	28,o
SLURRY PHASE B	2,o6	o,87	42,2	o,49	24,o	o,2o	9,7	o,5o	24,1
SLURRY PHASE L	2,o5	o,86	42,o	o,49$_5$	24,2	o,2o	9,6	o,49$_5$	24,2
OIL CIRCULATION (NITRIDED CATALYST)	2,81	o,96	34,2	o,92	32,8	o,25	8,8	o,68	24,2
LURGI MULTISTAGE REACTOR	2,16	o,93	43,1	o,48	22,2	o,23	1o,6	o,52	24,1
HOT-GAS-RECYCLE 2o A	2,95	1,o6	35,9	o,62	21,1	o,53	17,7	o,74	25,1
HOT-GAS-RECYCLE 7 C	2,75	o,99	36,1	o,82	29,7	o,29	1o,5	o,65	23,7
FLUIDIZED CATALYST (NITRIDED CATALYST)	2,34	o,78	33,2	o,54	23,o	o,34	14,7	o,68	29,1
OIL CIRCULATION (STEEL TURNINGS)	1,97	o,85	43,2	o,4o	2o,6	o,25	12,5	o,47	23,7

FIG. 8

Production Costs

Processes	Coal %	Power %	Capital Service %	Labour %	General Costs *) %
Arge	14,2	9,1	41,4	1,9	33,4
Synthol N	14,2	9,1	41,0	1,7	34,0
Synthol L	12,9	10,8	41,5	1,6	33,2
Hydrocol	14,5	9,7	40,1	1,6	34,1
Slurry Phase B	14,3	9,6	40,8	2,0	33,3
Slurry Phase L	14,1	9,2	40,4	2,0	34,3
Oil Circulation (nitrided catalyst)	12,0	10,4	41,4	1,8	34,4
Lurgi Multistage Reactor	14,6	9,9	40,2	1,8	33,5
Hot-Gas-Recycle 20A	11,8	18,8	38,2	1,3	29,9
Hot-Gas-Recycle 7C	12,6	11,0	41,4	1,5	33,5
Fluidized Catalyst (nitrided catalyst)	11,4	12,3	41,3	1,5	33,5
Oil Circulation (steel turnings)	14,5	10,2	40,3	2,0	33,0

* MAINTENANCE T
INSURANCE OVE
HEAD AUXILIAR

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Fig. 9

April 1976

SHARE OF PRODUCT GROUPS WITH ESSENTIAL INFLUENCE ON YIELD AND REVENUE

PROCESSES	C_2-C_4 HYDROCARBONS		NAPHTHA 375 °F		RAW MATERIAL FOR DETERGENTS		WAX		ALCOHOLS	
	% OF PRODUCTION	% OF REVENUE	% OF PRODUCTION	% OF REVENUE	% OF PRODUCTION	% OF REVENUE	% OF PRODUCTION	% OF REVENUE	% OF PRODUCTION	% OF REVENUE
ARGE	13,3	11,7	26,9 *)	2o,8	8,2	17,4	43,o	33,o	3,8	6,6
SYNTHOL N	35,7	37,4	41,5	3o,o	3,3	8,o	7,9	3,7	8,5	14,6
SYNTHOL L	64,o	66,3	3o,1	24,o	1,7	4,3	o,7	o,4	2,2	4,1
HYDROCOL	39,1	4o,4	42,5	32,8	3,3	8,4	3,4	9,7	6,4	12,1
SLURRY PHASE B	2o,4	21,o	63,9	47,5	8,1	17,4	2,9	3,1	2,8	4,5
SLURRY PHASE L	31,4	32,o	59,5	46,4	4,9	1o,8	-	-	2,4	4,2
OIL CIRCULATION (NITRIDED CATALYST)	3o,5	27,9	3o,2	19,8	2,7	4,9	21,o	19,3	15,o	21,5
LURGI MULTISTAGE REACTOR	25,2	2o,7	34,9 *)	24,o	1o,3	19,3	22,7	18,3	6,1	9,3
HOT-GAS-RECYCLE 7C	32,6	34,4	33,9 *)	23,9	5,7	1o,o	14,2	12,5	9,2	14,3
HOT-GAS-RECYCLE 2oA	77,o	75,6	18,6	16,o	-	-	-	-	4,4	8,4
FLUIDIZED CATALYST (NITRIDED CATALYST)	39,2	33,3	12,5	8,6	1,1	2,2	-	-	46,9	55,7
OIL CIRCULATION (STEEL TURNINGS)	35,7	31,8	32,4 *)	2o,8	9,2	17,o	9,5	8,3	11,9	16,6

*) NAPHTHA 43o °F

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FIG. 1o

APRIL 1976

ETHYLENE YIELD FROM FISCHER-TROPSCH PROCESSES

10^3 TONS-A-YEAR

PROCESSES	ORIGINAL ETHYLENE	ETHYLENE FROM ETHANE	ETHYLENE FROM PROPANE	ETHYLENE FROM BUTANE	ETHYLENE FROM NAPHTHA 375 °F	TOTAL ETHYLENE YIELD
ARGE	3,6	46,7	30,4	54,6	159,4 *)	294,7
SYNTHOL N	75,1	134,7	29,4	97,5	344,4	681,1
SYNTHOL L **)	49,6	359,8	99,8	222,2	244,2	975,6
HYDROCOL	42,2	131,2	34,1	134,6	346,1	688,2
SLURRY PHASE B	40,4	30,3	21,4	77,8	536,3	706,2
SLURRY PHASE L	38,3	42,4	49,2	125,4	498,8	754,1
OIL CIRCULATION (NITRIDED CATALYST)	45,3	127,9	48,0	73,6	252,3	547,1
LURGI MULTISTAGE REACTOR	8,2	80,6	58,1	92,5	245,3 *)	484,7
HOT-GAS-RECYCLE 20 A	-	590,7	260,9	153,9	123,0	1.128,5
HOT-GAS-RECYCLE 7 C	94,0	145,9	22,5	79,0	261,6 *)	603,0
FLUIDIZED CATALYST (NITRIDED CATALYST)	9,5	209,2	86,6	114,2	101,3	521,0
OIL CIRCULATION (STEEL TURNINGS)	44,3	164,0	41,2	85,9	227,3 *)	562,7

*) NAPHTHA 430 °F **) ESTIMATED VALUES

UNION KRAFTSTOFF FIG. 11 APRIL 1976

Share of Product Groups with Essential Influence on Yield and Revenue

Processes	C_2-C_4 Hydrocarbons		Naphtha 375 °F		Raw Material for Detergents		Wax		Alcohols	
	% of Production	% of Revenue	% of Production	% of Revenue	% of Production	% of Revenue	% of Production	% of Revenue	% of Production	% of Revenue
Arge	13,3	11,7	26,9 *)	20,8	8,2	17,4	43,0	33,0	3,8	6,6
Synthol N	35,7	37,4	41,5	30,0	3,3	8,0	7,9	3,7	8,5	14,6
Synthol L	64,0	66,3	30,1	24,0	1,7	4,3	0,7	0,4	2,2	4,1
Hydrocol	39,1	40,4	42,5	32,8	3,3	8,4	3,4	9,7	6,4	12,1
Slurry Phase B	20,4	21,0	63,9	47,5	8,1	17,4	2,9	3,1	2,8	4,5
Slurry Phase L	31,4	32,0	59,5	46,4	4,9	10,8	-	-	2,4	4,2
Oil Circulation (nitrided Catalyst)	30,5	27,9	30,2	19,8	2,7	4,9	21,0	19,3	15,0	21,5
Lurgi Multistage Reactor	25,2	20,7	34,9 *)	24,0	10,3	19,3	22,7	18,3	6,1	9,3
Hot-Gas-Recycle 7C	32,6	34,4	33,9 *)	23,9	5,7	10,0	14,2	12,5	9,2	14,3
Hot-Gas-Recycle 20A	77,0	75,6	18,6	16,0	-	-	-	-	4,4	8,4
Fluidized Catalyst (nitrided catalyst)	39,2	33,3	12,5	8,6	1,1	2,2	-	-	46,9	55,7
Oil Circulation (steel turnings)	35,7	31,8	32,4 *)	20,8	9,2	17,0	9,5	8,3	11,9	16,6

*) Naphtha 430 °F

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Fig. 10

April 1976

ETHYLENE YIELD FROM FISCHER-TROPSCH PROCESSES

10^3 TONS-A-YEAR

PROCESSES	ORIGINAL ETHYLENE	ETHYLENE FROM ETHANE	ETHYLENE FROM PROPANE	ETHYLENE FROM BUTANE	ETHYLENE FROM NAPHTHA 375 °F	TOTAL ETHYLENE YIELD
ARGE	3,6	46,7	30,4	54,6	159,4 *)	294,7
SYNTHOL N	75,1	134,7	29,4	97,5	344,4	681,1
SYNTHOL L **)	49,6	359,8	99,8	222,2	244,2	975,6
HYDROCOL	42,2	131,2	34,1	134,6	346,1	688,2
SLURRY PHASE-B	40,4	30,3	21,4	77,8	536,3	706,2
SLURRY PHASE L	38,3	42,4	49,2	125,4	498,8	754,1
OIL CIRCULATION (NITRIDED CATALYST)	45,3	127,9	48,0	73,6	252,3	547,1
LURGI MULTISTAGE REACTOR	8,2	80,6	58,1	92,5	245,3 *)	484,7
HOT-GAS-RECYCLE 20 A	-	590,7	260,9	153,9	123,0	1.128,5
HOT-GAS-RECYCLE 7 C	94,0	145,9	22,5	79,0	261,6 *)	603,0
FLUIDIZED CATALYST (NITRIDED CATALYST)	9,5	209,2	86,6	114,2	101,3	521,0
OIL CIRCULATION (STEEL TURNINGS)	44,3	164,0	41,2	85,9	227,3 *)	562,7

*) NAPHTHA 430 °F **) ESTIMATED VALUES

UNION KRAFTSTOFF FIG. 11 APRIL 1976

Cost and Revenue Comparison

for a capacity of 2,5 mill. tons-a-year

Process	Operating Costs Billion $/year	Revenue Required $ / ton	Actual Sales Revenue (Europe) $ / ton	Revenue Required / Sales Revenue
Oil Circulation (steel turnings)	0,57	238,8	186,9	1,28
Lurgi Multistage Reactor	0,63	245,6	173,5	1,47
Synthol N.	0,60	243,8	164,6	1,48
Slurry Phase B	0,60	241,9	160,4	1,51
Flzidized Catalyst (nitrided catalyst)	0,66	267,7	172,3	1,55
Slurry Phase L	0,60	242,3	152,7	1,59
Hycrocol	0,61	256,5	154,6	1,66
Arge	0,64	259,2	153,8	1,68
Synthol L	0,62	260,4	149,2	1,74
Oil Circulation (nitrided catalyst)	0,79	323,8	181,5	1,78
Hot-Gas-Recycle 7C	0,78	325,4	168,5	1,93
Hot-Gas-Recycle 20A	0,91	401,9	142,3	2,82

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Fig.12

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LECTURE - KÜHN

HOLIGHAUS: Thank you for your excellent contribution. There should be many questions, as the opinions you expressed differed in some ways from those voiced this morning. The discussion is now open.

GAENSSLEN: You mentioned that the product would have to increase in price by 40 - 50 % in order to arrive at a break-even situation. In order for this to happen, the price of oil would have to increase - meaning that the cost of everthing would rise, including coal and capital goods.
Have you refined your calculations to take into account this sliding effect? Because an economy where all prices except those of oil are static, cannot possible exist. A so-called "cascade effect" is produced.

KÜHN: We have not refined the calculations within the study mentioned in my paper. However, we have done this within our "parent" company, and our philosophy is as follows:

1) We do not agree with you in that costs rise in a parallel way. We believe that raw material prices (especially here in Europe, where we do not have raw materials in large quantities) will rise more. This is normally calculated taking into account a 2 % higher inflation rate rather than, for instance, capital. Every plant having a high capital charge will ultimately have an advantage in the course of time with the 2 % rise in material costs.

2) I do not agree that all costs will automatically rise parallel from the standpoint of "coming up" costs. I am convinced that raw material will increase in price because everyone calcualtes that the time of surplus is over.
The cost increasing factors of brown coal differ from those of hard coal, which has a high percentage dependent upon personnel costs. We calculate that we are arriving at a situation where the Rhineland brown coal will have a considerable advantge in cost over other competitive raw materials.

ZIEGLER: Is it difficult to establish these coupling factors: oil prices increase by a certain percentage - what influence does this exert on other products? If one assznes that the price ratio between mineral oil prices and other goods needed for such a plant, for example, should change by a factor of 30 % - 40 % (1.3 - 1.4), then one would reach a break-even point - but only in this case.

GAENSSLEN: When one examines something like the Fischer-Tropsch synthesis, one is dealing with a material containing oxygen, as opposed to oil, which contains no oxygen. So everything preserving that skeleton with oxygen has an economic advantage. The problem which arises is that we now have a chemical economy based on oil, favouring hydro-carbon structured materials. As soon as one examines synthesis gas, one must consider things which favour something having an oxygen structure. The best type of Fischer-Tropsch-synthesis is the methanol synthesis, which preserves everything.
An actual profitable example of a Fischer-Tropsch-plant exists; at least, if one studies the balance sheets, the Sasol plant is making a profit. For several years, it has had a very good combination of making the best use of the chemicals (admitedly, for a limited market). An extensive F/T plant would cover the entire oil needs of a large country, and produce many surplus chemicals. These could not be disposed of easily.
In the situation mentioned, an excellent optimal usage of chemicals (to improve the economic prospects for the fuels and automotive spirits) exists. This example shows that a F/T plant is able to "live" when properly run.

KÜHN: We must be careful when comparing the South African situation witz ours. The South Africans do not have an open market and compete with their technology and their product price level in total with the world-wide free energy markets.

GAENSSLEN: But South African petrol prices are lower than ours.

KÜHN: I would like to add more figures concerning the attractiveness of F/T: we have a U.S. paper showing that F/T has a liquid conversion efficiency of somewhat more than 30 %, whereas competing processes for upgrading coal, e.g. the coal processes or the combination of methanol production with the processshow efficiences of 40 - 50 % or more including gaseons byproducts and therefore, I cannot see a chance for F/T in a time when we calculate that raw materials in particular, being the most important factor for efficiency, are increasing in price.

KNUDSEN: I would like to explain what type of organisation ESCOE is: it is an engineering group. Our comparisons were made on the basis of useful liquid products and did not, by definition, account for the gas made as well. The question of efficiency has been an important one for a long time, and as it is pointed out as SASOL I, the process has become much more efficient as industries have become located around the plant, and used directly the methane produced by Lurgi generators instead of trying to recycle to extiniction.
A well-organised plant which can use methane in this way without reforming it has an efficiency of 55 % - high 50 %; it may begin as an isolated plant with below 45 %.
Jan Hygendorn of SASOL 2 confirmed that it had an efficiency of low 40 % and could perhaps look forward for some increase, although there is the question as to the number of chemical and satellite industries which can locate around the new plant. The industry located around the first plant is saturating its need for some products. It is not just a question of locating industry around this new isolated location; this was chosen solely for coal and is 100 miles from the nearest towns and villages.
How efficient SASOL 2 will become is unclear as yet, but is has potential. In order to achieve consistency, one needed to make some assumptions. In Barmouth, U.S. where there is a large demand for gas, it would make sense to choose the assumption of not recycling to extinction, but of selling it. A marginal competition exists, but the comparisons between different numbers are to vague now, that one must be cautious. We are becoming more interested in direct hydrogenation.

BAKER: More will be mentioned about this comparison during my talk tomorrow. The economics, particularly the question as to whether one should sell the gas at some price, or not, will be considered.

HOLIGHAUS: The thermal efficiency of such a F/T plant was extremely high; that was a combination of F/T with try-gas.

GAENSSLEN: Up to 67 % - 69 %.

KÜHN: It is absolutely impossible to equalise all these processes and not to take advantage of one process. The advantage of on process, for e.g. ist that it makes gases which can be sold. Yet in my opinion, this is illogical, because there is a better process for gas production: a very selective methanation step is made behind a gasifier. One subsidises by gas-selling the costly part of a F/T synthesis to make it more profitable than a F/T complete enclosed process with recycling.

HOLIGHAUS: S.N.P. makes the main product cheaper.

KÜHN: It is right that F/T processes are more econimically viable than others which produce gas as a by-product and sell it. I believe that this selling creates an advantage in that it subsidises the rest of the F/T synthesis. There is an even better and more selective process to make gas: methanation. F/T is not needed in order to sell gas.

SCHNUR:When producing 1 kg methane from syngas, more syngas is needed than for 1 kg of the higher grade liquid F.T.-products. If both processes are combined then the methane is left in the gas as SNG instead of being cracked to syngas. The liquid products are then manufactured from CO/H_2 via the F.T..Synthesis. Both products are favoured in this way.

KNUDSEN: That is an excellent point. We are under the pressure to find a way of making liquids from the various alternatives, all of which appear to be expensive. This one did seem more expensive until we realised that it could be subsidised with the gas which naturally comes off. (It is a natural phenomenon that a certain amount of methane is produced - and it is cheaper to use it than to try to regulate it to extinction just as a matter of principle. It is a question of being practical in this real situation.)
There are better ways of producing gas, but the first purpose was to try to eliminate the input problem; since this was the original goal, gas which results is a by-product. It is important that the by-product can be sold and used; it substitutes directly for many uses, particularly in industry, e.g. in a glass plant,

or a cement plant, which would perhaps have to use a fuel oil. We find that we can substitute; for example, cement plants can use coal. Enviromental restrictions prevented this, so they started using gas. Other plants, such as power plants, which have used coal, have had to switch to oil. Now some type of gas supplant that oil. One could treat all kinds of things, resulting in a reduction of the liquid problem; so we do not mind making the best use of methane rather than recycling, although when one makes comparisons it appears as if recycling should take place.

GAENSSLEN: When it comes to methane production in the F/T plant, one must distinguish between two sources of methane (I am thinking specifically of the Lurgi pressure gasification):

1) The actual F/T synthesis, where some amount of methane is produced

2) the main source of methane: the gasification step.
Economic analysis shows that these two methanes differ in cost. The metahne stemming from the gasification step is much cheaper, because it is only produced by pyrolysis. It does not have to be subjected to a difficult treatment in order for production to take place.
The methane produced in the F/T synthesis is in fact an undesirable by-product. So a good process would make a methane needed in the gasification steps and no methane whatsoever in the F/T synthesis itself.
When liquids and gas are produced simultaneously, the thermal efficiency of the process goes up from about 40 % to about 60 %. Lurgi is now conducting an optimisation study, which confirms this. Is it a blessing or not? It depends firstly on the relation between methane produced and liquids produced, and, last but not least, on the price obtainable for gas. If this price werde roughly 2-3 times the price of energy in the coal (i.e. the cost per G. calorie), the outlook would be quite pleasing, and an acceptable price would be gained for the liquid products. If, however, the methane only competes with the coal in heating value, a hopeless situation arises. So there is no clear-cut answer; one must be very specific.

I would like to comment on the thermal efficiences. The thermal efficiency of the F/T synthesis when aimed at liquid fuels is between 38 % - 40 % (depending on how it is run). A figure of above 30 % was mentioned earlier; no doubt this was owing to the fact that seperation of chemicals was very expensive in this case.

KÜHN: The figures were obtained from "The Comparison of Coal Liquefaction Processes" and include all costs from coal preparation down to produced liquids cleaning.

KNUDSEN: I would like you to expand upon your comments about looking for new catalysts. You mentioned that you tried unsuccessfully to obtain a more selective catalyst, and that there was littel hope for a CO selective hydrogen synthesis.

KÜHN: We are not catalyst specialists. I merely made the comment you referred to in order to explain that we attempted to clarify the picture regarding decisions about further research into coal upgrading. At that time (1975) our impression was that the chances of finding a really improved catalyst, which would make more than the available catalysts at SASOL, were remote. I would like Ruhr-Chemie to comment on this problem.

FROHNING: I disagree with your concluding remark; maybe some of the figures up to 1976 have been collected in 1975 or even earlier. Development from then on has taken a more favourable turn; it is possivle nowadays to have a special selectivity considering short chain olefins of above about 50 %. This is a high figure when compared with those processes shown in Dr. Kühn's diagrams, and indicates a real success. However, many questions remain unsolved and are perhaps insoluble. Maybe successful technical realisation, e.g. selectivity coupled to reaction conditions, economics and process conditions etc., cannot be achieved. The situation has altered to some extent in several research establishments; perhaps the methods you used in attempting to overcome difficulties have been followed up by other researchers tending with catalysts containing the iron in a modified form. Future improvements can be expected.

ZIEGLER: There are technical catalysts with selectivities of 50 % +; but their lifetime under operation conditions is very short.

FROHNING: But the lifetime of industrial catalysts in the past was not long either; for example, the lifetime of the SASOL catalyst in the entrained or fluidised bed was about 6 weeks.

HOLIGHAUS: The U.S. are probably more interested in fuel production; is there also a chance of improving selectivity in this direction to suppress methane formation, in order to obtain a higher yield on liquid products, e.g. gasoline?

KÜHN: Mr. Gaensslen and Mr. Knudsen pointed out that F/T might have an advantage when it produced methane in a saleable and desirable form. Mr. Gaensslen made an excellent distinction between methane in the original raw gas and methane coming from the synthesis. We found that these processes producing the desired products, i.e. low-molecular olefins, produce the high amount of methane (from the synthesis, not the gasification) in an undesired form.

HOLIGHAUS: Is there any chance of not only improving the selectivity for very special products, as we did, but also of improving the broad range which can be used as fuels?

FROHNING: This chance generally exists. But the problems related to the high-selectivity catalyst for producing shortchain olefins will be the same for this process. They can be solved to a certain degree - lower than 100 %. 100 % selectivity cannot be reached by such a process.

HOLIGHAUS: We should now proceed to the next paper. There will be an opportunity for discussing the problems already mentioned during the general discussion later. The next paper, by Dr. McNeese, deals with the comparison of coal liquefaction processes.

LECTURE - MC.NEESE

HOLIGHAUS: Thank you very much; please ask any questions regarding Dr. Mc. Neese's paper.

KRÖNIG: I agree with Dr. Mc. Neese in that the yields are good; but how can this process be carried out on a large scale, particularly with regard to the reactor? I am thining of a tubular reactor having tubes surrounded by molten metal, for instance sodium or potassium/sodium at a temperature of 950°C. It seems very doubtful, whether the tubes could stand the attack of the molten alcali metals from outside at the very high temperature of 950°C. A pure gas velocity greater than 4 meters per second exists in the tubes under the intended conditions. Coal, ash and turbulence could easily cause erosion inside the tubes.

The reactor described has 36 tubes with a space below; hydrogen with a temperatrue of 3.000°F is added by means of nozzles, around which coal is introduced into the tubed. This process could work with one tube; but it must be impossible to distribute the coal entering from outside equally through 36 tubes. I would be grateful to Dr. Mc.Neese if he could describe how these difficulties could be overcome by this process.

MC.NEESE: I agree with your comments concerning flash hydropyrolysis processes; inherent in the processes in question is a rapid heat-up and an equally rapid quench. The conditions under which the heating and quench must be carried out certainly present a challenge; on the other hand, if one examines the data base available, one can see the potential held by this class of processes and conclude that efforts toward solving these problems are worthwhile. This class of processes (rather than the conventional approach) may lead to a somwhat more attractive overall product, or particular, chemical feed stocks. There are no proven methods, however, for overcoming the problems at this point; work on the processes described should therefore be continued. We have performed several studies with this primarily in mind, and have indicated in a relative manner which possibilities present themselves.

KRÖNIG: A proposed pilot plant bult from one tube and one nozzle was mentioned in the paper; this would not be a suitable pilot plant for finding out whether the concept described would work. One tube and one nozzle are possible; but 36 tubes and nozzles

HOLIGHAUS: There are many significant and unsolved problems, e.g. those concerning materials and their selection; the uncertainty of the estimation in question has a very broad range.

MC.NEESE: Characteristically, as in the case of many systems in an extremely early stage of development, there are many questions unsolved as yet. What one is trying to answer in scoping studies is essentially the question: if these problems can be solved, what is the potential of the process?

HOLIGHAUS: You took these problems into consideration when making your estimation; I believe you cannot use conventional materials?

MC. NEESE: Yes. These factors were taken into account when the assessment was made. The range of uncertainty in estimated costs is relatively large.

ZIEGLER: You showed a number of figures concerning motor gasoline production starting with in situ coal gasification. Did you obtain figures comparing just the prices per Mio BTU for the syngas entering synthesis and the syngas produced by a Lurgi or Texaco

MC.NEESE: This comparison was shown on the last slide and relates to one specific case, namely syn gas produced by whatever means one wants to consider at $ 2,60 per 1,000 000 BTU. This syn gas cost would result in the gasoline prices which we have considered. Here, the syn gas could have been produced by any gasification system.

ZIEGLER: But it is doubtful whether one can relly calculate the costs of in situ gasification nowadays.

MC.NEESE: Calculating to the best of our ability, we arrive at a cost of $ 2,60 per million BTU. That value is consistent with the final gasoline cost which I have shown. We have not examined the ways of producing syn gas by other routes, and the costs involved, and translated these into gasoline costs.

HOLIGHAUS: I noticed that your figures indicated no influence exerted on the product price by drilling and logging and casing. This indicates that the drilling and related costs have no influence on the product costs; but the drilling, etc. are a great influence on the cost of syngas, as was shown by previous comments. Perhaps we could examine this conflict.

MC.NEESE: The drilling and casing costs can be significant portion of the final product cost. This cost element depends upon a number of factors: well-spacing, thickness of the coal seam, etc. For the conditions which we chose, little sensitivity existed; had we chosen a thinner seam and closer well-spacing, this cost element would have become more important.

GAENSSLEN: You mentioned the Schroeder process with around 76 % thermal efficiency; please could you explain this system?

MC.NEESE: This was a flash hydropyrolysis type process, which we have discussed earlier. In the case of the process described, and other flash hydropyrolysis processes, enough data exists to create interest. The data are not definitive, however, in a real engineering design sense.

HOLIGHAUS: I would like to thank Dr. Mc. Neese and introduce the next speaker, Dr. Hill, who will read Dr. Rogers' paper.

TECHNICAL AND ECONOMIC COMPARISON
OF COAL LIQUEFACTION PROCESSES

(Presented at the IEA Coal Liquefaction Workshop
September 18-19, 1978, Julich, West Germany)

Earlier this year, ESCOE conducted a study of the leading coal liquefaction processes. The purpose of that study was to appraise the commercial viability of each process. A ranking of the processes was desired to determine which would yield the best results from further development expenditures. We also have a broader task underway to appraise a variety of coal conversion processes, including some to produce gas and electric power. We will report here on partial results from both studies, confining this discussion to liquefaction processes.

We will discuss three direct liquefaction processes - H-Coal, Solvent Refined Coal, and Exxon Donor Solvent - and two indirect processes - Fisher-Tropsch and M-Gasoline via methanol. Both the technical and economic aspects of each process will be examined.

A standard measure for any process is the thermodynamic efficiency. However, a more useful measure for coal conversion processes is the product's efficiency which is the ratio of the heating value of all fuel products to the energy inputs. Thermodynamic efficiency is a theoretical limit for products efficiency after recovery of all waste heat. The product's efficiency is a good indicator for the amount of resource required to make a specified amount of product.

Most of the processes studied are in the development stage. A Confidence Index system is used to show the degrees of uncertainty for both process and economic data. This system is shown on Table 1. It consists of two parts, a letter to show the stage of process development and a number to show generally the degrees of confidence which may be associated with the technical development and cost estimates for each process.

TABLE 1: CONFIDENCE INDEX

Process Development	Economic Reliability
D - Exploratory stage, not beyond simple bench tests	4 - Screening estimate, very approximate
C - Development stage - operated on small integrated scale only	3 - Incomplete definition for estimates used
B - Pre-commercial, successful pilot plant operation	2 - Firm basis for values developed
A - Complete, process demonstrated sufficiently to insure commercial success	1 - Values considered to be satisfactory for commercial venture

The next comparison recognizes that some processes involve complex or difficult features not present in others. Operations which are common to many processes or are considered standard industrial technology are not included. A list of complicating features is used to arrive at a Complexity Index value as shown in Table 2. The features which make up Complexity Index are not of equal weight so no inferences should be drawn from the numbers except as a non-linear indicator. For example, a plant with a Complexity Index of 4 is certainly not four times more complicated than a plant with an Index of 1. There are no proportional relationships for these values. No attempt is made to use any sort of weighting factors because these would be too subjective at this time to be of real value. The Complexity Index alone cannot be used as a basis to choose among processes.

TABLE 2: COMPLEXITY INDEX

Item	SRC-I	SRC-II	EDS	H-Coal	F-T	M-Gas
1. Highest Pressure of Group				X		
2. Highest Temperature of Group						
3. Fluidized Bed						
4. Multi-Fluidized Beds						
5. Ash Fusion Problems						
6. High Head Recycle Pump or Compressor				X		X
7. Solid-Fluid Separation Problem	X			X		
8. Significant Solids Handling	X					
9. Power Turbine Critical to Cycle						
10. Oxygen Used in Process						
11. Tar Products					X	
12. Chemical Reaction Complexity			X		X	X
13. Fired Preheater Used	X	X	X	X		
Complexity Index Value	3	1	2	4	2	2

A summary of these comparisons for the processes studied is shown in Table 3. The only conclusion possible at this point is that all are technically feasible and the real differences relating to commercial viability must be determined by cost studies.

TABLE 3: PROCESS COMPARISON

Process	Complexity	Confidence	Efficiency Thermal	Efficiency Product
Fischer-Tropsch	2	B-2	56	48
M-Gasoline	2	C-3	52	47
H-Coal-Syncrude	4	C-3	-	71
H-Coal-Fuel Oil	4	C-3	-	74
Exxon Donor Solvent	2	C-3	66	64
SRC-I	3	C-3	71	70
SRC-II	1	C-4	70	70

It is the custom to show product costs on a $/BTU basis for coal conversion plants. The comprehensive study by C. F. Braun Co.(1) in 1976 on coal gasification plants followed this precedent. That study showed costs based on both utility and private sector financing. Their financial guidelines remained valid for present purposes and were used for our study. Costs were determined in 1978 $ for capital and all manufacturing expenses required to estimate total product costs. These costs, on both a utility and private financing basis, are shown in Table 4. This Table also shows the cost fractions associated with the fuel, capital, and remaining operating costs.

The differences in the calculated unit energy product costs are not so great considering the uncertainty of the input data. Of particular interest is the relative costs due to fuel capital and operating charges. Because of

TABLE 4: PRODUCT COST-ENERGY BASIS

PROCESS	UTILITY FINANCING		PRIVATE FINANCING	
	$\$/10^6$Btu	f / c / m	$\$/10^6$Btu	f / c / m
FISCHER-TROPSCH	5.49	.50/.27/.23	7.08	.39/.44/.17
M-GASOLINE	5.54	.47/.30/.23	7.29	.36/.47/.17
H-COAL SYNCRUDE	3.78	.50/.28/.22	4.88	.39/.44/.17
H-COAL FUEL OIL	3.34	.54/.25/.21	4.22	.43/.41/.16
EXXON DONOR SOLVENT	4.38	.47/.29/.24	5.74	.36/.46/.18
SRC - I	4.18	.48/.28/.24	5.43	.37/.45/.18
SRC - II	4.02	.48/.29/.23	5.26	.36/.46/.18

$$\text{PRODUCT COST} = \frac{\text{FUEL + OPERATIONS + k x CAPITAL}}{\text{MILLIONS OF BTU PRODUCED}} = \$/10^6\text{BTU}$$

$$\left.\begin{array}{l} k = .12 \text{ UTILITY} \\ k = .247 \text{ PRIVATE} \end{array}\right\} \text{FINANCING}$$

f = FRACTION OF COST FOR FUEL
c = FRACTION OF COST FOR CAPITAL CHARGES
m = FRACTION OF COST DUE TO OPERATING & MAINTENANCE COST

Note: Costs based on a conversion rate of 25,000 TONS/DAY of coal with a HHV of 11,200 BTU/LB-DRY with a delivered coal prise of 30 \$/TON (2,000 LBS = 1 TON) and a plant operating factor of 0.9.

the financing advantages for utilities, the fraction required for capital is higher with private financing. A change in fuel cost would, of course, cause all fractions to change.

The allocation of costs to the different products in a multi-product plant is, at best, an arbitrary process. It is necessary, however, to use a meaningful and consistent technique to apportion total costs among the various products. Further, it is highly desirable to calculate a single "average" product cost that recognizes the cost differential among products with different refined quality.

In our first report(2), we developed a technique to account for different product values. It uses weighting "value factors" for each product and computes a Cost Index. Such a technique recognizes and properly credits the higher value for premium gasoline from the M-Gasoline process as compared to heavy fuel oil products that dominate the product output from many of the other processes. The input values used here for the Cost Index approximate a 1978 market value.

The definition of Cost Index follows:

$$\text{Cost Index in \$/BBL} = \frac{F + M + k\,C}{\Sigma f_i B_i}$$

where: F = Annual fuel cost
M = Annual operating and maintenance costs
C = Total capital investment
k = Capital recovery factor
B_i = Annual production of each product in barrels
f_i = Value factor for each product

Using this Cost Index, the cost for individual products and the average product cost referenced both to premium gasoline and to #6 boiler fuel, are shown in Table 5. The value factors for each product, referenced to premium gasoline, are also shown. Individual product costs are determined by the product of Cost Index and value factor. Costs calculated for individual products will be the same regardless of which reference product is chosen.

TABLE 5: PRODUCTS COST COMPARISON

Process	Cost Index $/BBL Reference		Product	Daily Production		$f_i(1)$	Cost	
	Premium Gasoline	Boiler Fuel						
Fischer-Tropsch	41.24	28.88	Gasoline	18,200	BBL	.90	37.12	$/BBL
			LPG	18,800	BBL	.69	28.46	$/BBL
			No.2 Oil	1,200	BBL	.85	35.06	$/BBL
			No.6 Oil	2,000	BBL	.70	28.88	$/BBL
			Med. Btu Gas	4.2×10^{10}	Btu	.462	3.81	$\$/10^{6}\text{Btu}$
			SNG	4.5×10^{10}	Btu	1.216	10.03	$\$/10^{6}\text{Btu}$
M-Gasoline	36.20	28.03	Gasoline	52,700	BBL	1.00	36.20	$/BBL
			LPG	7,300	BBL	.69	24.98	$/BBL
H-Coal Syncrude	29.73	20.81	Naphtha	27,300	BBL	.88	26.16	$/BBL
			No.6 Fuel Oil	20,800	BBL	.70	20.81	$/BBL
			SNG	1.2×10^{11}	Btu	1.11	6.60	$\$/10^{6}\text{Btu}$
H-Coal Fuel Oil	30.78	21.55	Naphtha	17,000	BBL	.88	27.09	$/BBL
			No.6 Fuel Oil	37,200	BBL	.70	21.55	$/BBL
			SNG	7.3×10^{10}	Btu	1.10	6.77	$\$/10^{6}\text{Btu}$
Exxon Donor Solvent	36.72	25.70	Propane	3,270	BBL	.69	25.34	$/BBL
			Butane	3,500	BBL	.75	27.54	$/BBL
			Naphtha	19,900	BBL	.87	31.95	$/BBL
			No.6 Fuel Oil	28,700	BBL	.70	25.70	$/BBL
			SNG	5.45×10^{10}	Btu	1.3	9.55	$\$/10^{6}\text{Btu}$
SRC – I	53.37	37.36	SRC-I	52,230	BBL	.63	33.62	$/eQ.BBL
			Fuel Oil	7,100	BBL	.70	37.36	$/BBL
SRC – II	39.32	27.52	LPG	5,500	BBL	.69	27.13	$/BBL
			Naphtha	10,700	BBL	.85	33.42	$/BBL
			No.6 Fuel Oil	45,300	BBL	.70	27.52	$/BBL
			SNG	3.0×10^{10}	Btu	1.30	10.22	$\$/10^{6}\text{Btu}$

Notes:
1) Based on Premium Gasoline Reference Product
2) SNG = Substitute Natural Gas ≥ 900 Btu/SCF, (5 x 106 Btu/Equivalent BBL)
3) Costs are on Same Basis as Table 4
4) Costs were revised from original presentation due to data correction

11/10/78

Our earlier report, because it was specifically limited to liquid products, gave no credit for gas. This is unrealisitc. However, the value of fuel gas in the U. S. market is difficult to define because of the history of severe price regulation for natural gas. Value factors assigned to fuel gas are proportional to heating value. For computing purposes, the amount of gas produced is put on an energy equivalent to liquid products basis.

We believe the analytical tools described here are sound and can be applied to other types of coal conversion processes. The quality of the results must always reflect the reliability of the input values. Ultimate decisions will include additional considerations. In addition to assurance of adequate returns and best use of capital, actual decisions for plants are expected to be site specific. Only then will fully accurate allowances for all costs and materials be possible.

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LECTURE - BY ROGERS, READ BY HILL

HOLIGHAUS: I would like to thank Dr. Rogers and Dr. Hill for this excellent paper. We will discuss this paper and then proceed to a general discussion.

KÜHN: You showed (in the last slide but one) figures for the average cost in dollars per barrel of products obtained from the various processes and remarked that this average cost figure was unrelated to the value of the product in terms of today's market. But even bearing this in mind, I feel that your figures are rather misleading, because an average cost figure of 40 dollars per barrel of qualified motor gasoline cannot be compared with a figure of 30 or 38 dollars per barrel of mixture of products of heavy fuel oil up to gas.
I believe that even given a higher figure, a process which produces the higher value product may be profitable to a greater extent.

HILL: In today's market one cannot sell any of the products from liquefaction of coal. Here, we tried to derive a single cost number which was the least ridiculous one for a process providing a multitude of products in different quantities.
If a coal liquefaction industry is developed in the U.S., the market will bring in different technologies depending upon the coal available in various regions, the market which the company in question is attempting to satisfy, etc. We are trying to provide the Department of Energy with a ranking of these processes in terms of a single number. This number must be seen in the context of how it was developed. We feel that our comparisons will help decision-making regarding which process should receive additional research support. Our method provides a better insight for government decision-makers responsible for technology.

ZIEGLER: I support your single number concept as far as decision making is concerned; but if one can add another column showing, for example, the 1978 revenue on the mixture of products obtained by a specific process, and then takes the ratio of the costs per million BTU in relation to the mean number obtained when products are sold, one needs the real number afterwards as a decision-maker.

GAENSSLEN: I would like to raise an objection to one point: if one has an unlimited market, the best thing to do in a Fischer-Tropsch-plant would be to make only wax. But the question is: can 3.000.000 tons of wax be sold? The answer is negative. Clearly, one must add other constraints.

BAKER: I would like to ask Dr. Hill to amplify a little on the value factor, in particular with regard to table 5; is it a confusion of units that SNG appears to be worth more than liquids by the value factur? Or does a unit scaling occurt?

HILL: Derivation of the value factor of methane is a problem in the U.S., where methane is controlled at a price well below its value. The value factor for the liquids is a relatively simple concept; it is simply a matter of taking the present selling price for these liquids in the U.S. and assigning a value of one to the most expensive liquid.
In the case of gas, it was assumed in the U.E. that methane had the same value as premium gasoline on a BTU basis. Then values were assigned for other gases on the basis of the ratios of heating values.

BAKER: I will discuss adjusting the SNG relationship tomorrow.

GAENSSLEN: I would like to put a question regarding the table; the average cost is 41.04 per barrel. When this is split up into SNG and gas, one of these is valued at lesse. Then the other should increase in cost. The average should be obtained from weighing up the other costs - things are clearer than in the case of SRCI, where the average cost is 46 and the products all cost lesse. We did a similar calculation, giving the products the value I. Eventually, the weighted average would equal the average cost.

BAKER: Maybe the cost described in the paper was actually weighted by the value factor. Perhaps Dr. Hill was really saying that if that factor is included the 40.04 is in fact adjusted so that the LPG need only be sold at 27.63.

HILL: That number and these relative numbers are independent.

BAKER: You have taken that into account by the value factor. If you have a product which is low in value inherently, you have represented it in the value factor and taken that into account through the equation in table 6 in the cost index.

GAENSSLEN: If that were the case, a loss on every product would be made; only the average cost is shown - the remainder of the costs are not indicated.

KÖLLING: I would like to express a personal opinion here: in line 7 ("the fluid separation problem") it is feared that SRC I and H-coal have problems, whereas SRCII and other are problem free. I believe that they all create problems.

HILL: We agree. The question is where one should draw the line to obtain an complexity measure. One would certainly argue that in order to do this more throughly, one should start trying to assign continual values.

This study assumed that we had examined all the other technologies and what had been said about them. We had tried to obtain a relative comparison of those and find out whether some were more commercially viable than others. We also asked whether the technology would be ready, easy to construct, reliable, and in keeping with enviromental regulations, etc. Maybe one of these processes would be so much farther advanced that it could be put into practice before all the others.

The result of all this was that all processes were more or less commercially viable; if there were a market, one could introduce any, or all of them, reasonably quickly. We focused mostly on the question of economics.

HOLIGHAUS: I would like you to explain the differences in price (shown in table 4) for H-coal syncrude and the SRCII process. The amount spent on fuel is even lower in the case of SRCII and the efficiency on a product base is much higher. Why do you have a higher price for SRCII? Line 3 can be compared with the bottom line.

HILL: This ratio is multiplied by 4.10 to obtain the feed price. So the comparison which you would like to make is 4.10 x 45 versus 3,65 x 48.

HOLIGHAUS: If one takes the overall costs, then 45 % of these are fuel costs. As there is a similar efficiency with SRCII, then at least the price should be within the same range.

GAENSSLEN: Which has the highest efficiency - SRCI or SRCII?

HOLIGHAUS: SRCII has a 70 % efficiency on a fuel basis. HR-coal syn crude is compared with SRCII.

HILL: The fuel cost here is 48 %. This would work out at about $ 1.80. Another factor: these processes are for the U.S. market, and the different prices of coal also enter into the question. We are looking at different coals and the most likely process for a given coal.

GAENSSLEN: I believe, however, that it is always the same coal.

HOLIGHAUS: This topic can be discussed at greater length tomorrow; we could now consider other points.

KÜHN: I would like to make a very general observation; I feel that we are continually discussing the same problems during this meeting today: comparisons between various technologies.
We compared the efficiency of one or the other of the Fischer-Tropsch-processes; we are now debating the value of the products resulting from the respective processes. We can never compare our figures until we have defined beforehand the nature of the topics discussed. It is very clear that the more the product is upgraded, the more the thermal efficiency of a process or process chain decreases. The product value per BTU or per barrel increases considerably as one upgrades in a product chain. We are always considering different things; Mobil gasoline is different from the SRCII fuel process.

HILL: This is a very valuable point; the reason for starting our type of study was to try to produce for the U.S. government some standards against which all these different technologies could be compared. We put much effort into developing uniform assessment methods in order to make comparisons. We aim to obtain some kind of agreement as to how we are going to perform the analysis.

In the U.S. we are now arguing about numbers whose differences are so much greater than in the cases we are examining here. For example, one is forever saying that SNG costs 6 dollars, whereas imported LNG costs 4 1/2 dollars. 4 1/2 dollars relates to the expected price of the project planned for 1982; the former price relates to an entirely different estimate, based on different processes, times, dollars, assumptions, etc. Yet some are prepared to make decisions after considering these kinds of figures. Much of our work consists of examining data very carefully, in order that the meaning of the numbers, rather than the approach, can be understood.

KÖLLING: Discussion of the thermal efficiency is difficult; but you now have figures for the costs, and I would like to repeat Dr. Ziegler's earlier question: which process might be economic if the coal price were zero?

HOLIGHAUS: An answer to this question would call for a general discussion.

KÖLLING: I am considering table 5, which contains many figures for the costs.

HILL: The paper I will present tomorrow deals with the difficulty of coal liquid products entering the U.S. market, the varrying price of oil in the U.S. as opposed to coal price in the U.S. in the 1970s, and inflation. The projected cost of liquefaction is rising very steeply.

HOLIGHAUS: I would like to refer to your calculations in table 4, where you have integrated your fuel price.

HILL: The numbers shown here were not derived by us; they were prepared by the efforts of several researchers. We have been obtaining construction costs from various engineering firms and research groups and put them on the same basis, i.e. 1978 dollars.

Now we are performing other work with the DOE: raising questions about the validity of the presently available data we are using. We are trying to put this data on a consistent basis; then we must deal with its quality.

BAKER: If one subtracts the coal costs from the figures which most of us obtain when we calculate with these numbers, one has a result of about 3 dollars per million BTU with free coal. This translates into roughly 18 - 20 dollars per barrel - this is comfortably more than the oil price at the moment. One must expect a substantial increase in the price of oil, unrelated to coal price, for these prices are competitive. This constitutes a worry for us.

HILL: In the U.S. the average price of coal has increased at a greater rate than the price of oil. On the U.S. market, oil is not sold, at world prices, because of our government regulations. But the average price of coal has increased slightly more than the price of oil; a very close correlation exists between the two.

HOLIGHAUS: I would welcome any general statements or comments regarding this rather difficult matter.

KNUDSEN: I would like to instigate a discussion about catalysis. Most work in Germany in this field deals with catalysing selectively with valuable long-life chemicals. I believe there is a definite disagreement regarding the prospects for improvement - what can research do to test out new areas? Which developments have taken place in the Federal Republic of Germany since the 1930s - 1940s?

FROHNING: A programme was launched in 1974 base on previous knowledge. It aims directly to produce more selectively specific chemicals by reaction between carbon monoxide and hydrogen. So far, considerable improvements compared to the technical knowledge of 1974 have been made; however, these are not great enough to serve as a base for calculations for technical units. In principle, it appears to be possible to increase selectivity of known catalysts to a degree which makes them attractive for technical processes. Processes which have been mentioned, defined and compared during