

(1)

SCANNING OF REEL #13
U. S. GOVERNMENT TECHNICAL OIL MISSION

Bag 3500 - Target No. 30/4.05 - Bohlen

Alkylation (Grabag)

The first section of the reel is apparently a continuation of Alkylation work from the preceding film as it starts with Item Index #71. The frames in this section are not numbered, but at the beginning of each item, there is a notation representing the bag, target and item number, such as 3500-71 - 30/4.05.

Item Index #71 - Utility Requirements for 50,000 T/Yr. Alkylation Plants - July 22, 1941. This is a one page summary of the steam, cooling water, refrigeration, electricity, etc. required by an alkylation plant of the above capacity.

Item Index #72 - Alkylation Operating Schedule, 1941, 1942, and 1945. Bohlen - September 17, 1942. The values given (one frame) include operating hours per year, average throughputs, and average yields.

Item Index #73 - Leuna Alkylation Operating Conditions and Relation of Octane Number to Isobutane Concentration - 11/3/42. The single page of this item includes a flow sheet with analysis of the hydrocarbon streams and two curves showing the variation of alkylate octane number with isobutane concentration when using isobutylene and normal butylene. The Leuna operation was with 40-45% isobutane and 18-20% alkylate concentrations. The inlet gas was 56.2% i-C₄+, 33.0% n-C₄+ 4.4% n-C₄-, and 4.6% C₃'s.

Item Index #74 - Production of High Antiknock Gasoline Through Alkylation of Isobutane with Butene-1 in a Semi-Technical Pilot Plant - January 20, 1942. There are seventeen frames including text, flow-sheet, and product distillation curve of the alkylation results on a semi-technical scale from March-July, 1941 after the original development of the process in the laboratory. Discussions are presented of catalyst, rate relations, formation of emulsions, temperature, pressure, and throughputs. The information presented does not seem to be of particular significance.

Item Index #75 - Comparison of the Production of Alkylate with Butylene Production by Catalytic Dehydrogenation and Chlorine Dehydrogenation by the Bahr Method. This discussion is contained on nineteen frames and consists of text, detailed cost comparisons in tabular form, and flowsheets containing stream analyses for the various schemes. The Bahr method consists of thermal chlorination, separation of the monochloride followed by thermal dechlorination. The construction costs were higher for the process using chlorine dehydrogenation, but the production costs were about the same.

Item Index #76 - Cost Study for Production of Alkylate Blending Agent and ET 100 by Various Methods - October 15, 1942. The various methods differ essentially in the dehydrogenation process employed. The processes are catalytic dehydrogenation, Bahr thermal chlorine dehydrogenation, and the Oppau liquid phase chlorination under ultraviolet light. The twenty-two frames contain detailed cost comparisons and process flow in the form of text, summary tables, and flowsheets.

Item Index #77-#82 (See Index). These items totaling twelve pages are discussions on the requirements, procurement, and regeneration of sulfuric acid for the various alkylation plants. Included is a table giving the production capacity as well as acid requirements of twelve alkylation plants. The regeneration method employed was dilution to 50% with water, removing the contaminants, evaporating to 96% acid and fortifying.

Item Index #83 - Mixture Rule - 8/18/41. Of no importance.

Item Index #84 - Alkylation: Utilization of Fischer Butane - 6/30/43. This item covering three frames gives a table summarizing the amounts of isobutane, n-butane, and olefins which various plants supplied to Leuna for alkylation. No process data are given but descriptions are presented of various equipment, materials, and energy requirements for a projected plant.

Item Index #85-#107 (See Index). Various maps, ground plants, and drawings of plant areas, plants, buildings, location of equipment, etc.

Item Index #108 - Construction Costs for Alkylation and Dehydrogenation Plants at Bohlen.

Bag 3043 - Target No. 30/4.02 - Leuna

The frames in this section are not numbered consecutively, but the pages composing the individual items are so numbered.

Item Index #1 - Organization Chart of Merseburg Main Laboratory - Oct. 1, 1944. These five frames are tables listing the men heading the various research groups and sections of the laboratory and the respective problems. In general, no unusual development work seemed to be going on. Some of the general problems being investigated were: detergents, sulfochlorination, sulfo-oxidation, sulfoamination of hydrocarbons, cutting oil development, nitration of aliphatic hydrocarbon, organic peroxides, plastics, catalysis on synthetic resins, inorganic catalysis reactions, kinetics, toluene and xylene production, oxidation of hydrocarbons, motor fuel production by hydrogenation and by catalytic cracking, HF catalysis, and production of isoprene from isobutylene.

Item Index #2 - Minutes of Technical Meetings of the I. G. Farben. This section totaling 193 frames, numbered consecutively, contains the minutes of seven such meetings dealing principally with development work at Leuna and Oppau on light hydrocarbon processes. The results and data presented appear to be fairly complete but without complete reporting of the discussion. The meetings were held at:

1. Oppau - July 24-25, 1940
2. Merseburg - January 29-30, 1941
3. Berlin - April 22, 1941
4. Ludwigshafen - March 16-17, 1942
5. Berlin - July 16, 1942
6. Berlin - November 11, 1942
7. Berlin - January 20, 1943

The technical subjects under discussion were:

Alkylation - H_2SO_4 , and $HF-BF_3$

Dehydrogenation - catalytic (Cr_2O_3) and with chlorine

Isomerization - $AlCl_3$, tungsten sulfide under pressure

Polymerization - phosphoric acid and magnesium phosphate

Olefin concentration - use of $AgNO_3$ solutions

Aromatization - Cr_2O_3

Michael's liquid phase hydrocarbon synthesis

Paraffin (wax) synthesis

Olefin synthesis and utilization

Raman, ultraviolet, x-ray and electron dispersion analyses

Metallic mirrors and non-reflecting glass by high vacuum evaporation.

Methanol from CO and H_2 with subsequent reaction with benzene to produce toluene

Use of residue from coal carbonization and from mild hydrogenation of coal for binding material to briquette non-coking coals to allow coking

Besides the main topics, there is given a topical summary of the research work at Leuna. Work was being carried out (Nov. 1942) on coal hydrogenation, CO and H_2 reactions; development of moving catalyst for C_2H_4 -chlorination, powdered catalyst cracking, ketone production, products from adipic acid, oxo-process, iron catalyst hydrocarbon synthesis, sulfide catalyst for olefin hydrogenation, and enrichment of heavy water.

In general, the technical aspects are not worthy of translation, though some points may be mentioned.

The experiments with tungsten sulfide isomerization in the presence of hydrogen gave good results at 200 atm., 400-415°C. Results were also favorable at 50-100 atm.

The development of bead catalysts for dehydrogenation having high activity and high mechanical strength.

Chlorine dehydrogenation at 450°C. with 1:0.6-0.7 n-C₄H₉Cl gave higher olefin concentrations than could be obtained catalytically.

The best polymerization catalyst developed on a lab. scale was acidic primary magnesium phosphate stabilized with silica gel or kieselguhr. This catalyst as well as copper pyrophosphate could be improved by the addition of a few per cent of carbon such as by adding an oil and carbonizing.

The activity of AlCl₃ for isomerization can be improved by the addition of a small amount of silicon tetrachloride.

Item Index #3 - Synol (straight chain alcohols from water gas) and hydrocarbon synthesis - 1941 and later. This section consists of 56 items (See Index) totaling 306 frames, numbered consecutively, dealing principally with the development of iron catalyst synthesis processes. Details of yields, product distribution, and quality are given. A number of frames are illegible and much of the subject matter is repetitious. A great deal of the material is of no particular value, but some is of technical interest. It is the opinion of the reviewer that no article is worthy of complete translation though partial translations have been made.

Items 1 and 2 contain very little technical information.

Item 3 - Results are given of the testing of three precipitated iron catalysts of similar composition.

Source of Catalyst	Leuna	Oppau	Ludwigshafen
Fe	100.0	100.0	100.0
Cu	20.8	21.5	15.4
MgO	38.1	38.7	38.0
SiO ₂	74.5	75.5	74.5
K	9.3	7.0	9.3

These catalysts were reduced for 20-42 hours at 180-230°C. and were tested under pressure at 200-220°C. These catalysts produced highly alcoholic material. A tabular comparison is given between the testing results of the Ludwigshafen catalyst and the standard fused catalyst (KY-17, analysis not given). The conclusion was drawn that the only advantage the precipitated catalyst showed was in the lower reduction temperature 200°C. vs. 500°C.

Items #4-7 contain no particularly useful information. The statement is made in Item 6 that at that time (March, 1943) no precipitated catalyst had been found superior to the fused preparation KW-17.

Item #8 mentions results of the development work at Oppau on the production of olefins with a cobalt catalyst and work with iron catalysts for paraffin production. The iron catalyst used was Fe:Cu = 4:1 on Kieselguhr and contained about 60 Kg. Cu/Ton. It was prepared by precipitating hot (90-100°C.), with 120 kg. of K_2CO_3 in 800 liters of water, the metallic carbonates from the solution of 80 kg. of copper - iron nitrates mixed with Kieselguhr in 800 liters of water. The supernatant liquid was decanted and the alkali content of the remainder adjusted by dilution to 10-12 gm/liter. Then the precipitate was filtered and dried under vacuum at 200°C. A high linear velocity during reduction was considered important to prevent local hot spots.

Items #9-11 contain material of no particular technical interest.

Item #12 is a fairly complete discussion of the iron catalyst Michael hot gas recycling process as developed up to September 1938, including a general description of equipment, catalyst (no analysis), product distribution, properties and use of products, patent situation, and relation to the Fischer process. The material contained herein has been rather generally reported in other reviews, and in the opinion of the reviewer is not of sufficient interest to report fully.

Items #13-15 contain further discussion of the Michael gas recycle process including the following comparison of this process with the standard Fischer-Tropsch synthesis.

COMPOSITION OF THE PRODUCTS - WT. %

Process	Fischer-n-press.	Fischer, 10 atm.	Michael with tail	Michael without	
			gas workup. Poly	workup. Poly. gaso.	from
			gaso. from C_2H_4, C_3, C_4	C_2, C_3, C_4	C_3, C_4
Products					
Primary Gasoline (Alcohols incl.)	36.8	14.2	57.3	57.4	64.0
Polymer Gasoline	6.5	4.7	23.0	23.0	15.9
Diesel Oil	45.5	42.2	14.4	14.3	16.0
Diesel from Poly.	1.0	1.2	3.7	3.7	2.3
Paraffin 320°C.	10.2	37.7	1.6	1.6	1.8
			Yield in Gm. per 1000 $CO + H_2$ (100%)		
Liquid + Solid	120	145	83	83	83
Polymer Gasoline from Gasol.	9.7	8.8	30	30	18.5
Total Output	129.7	153.8	113	113	101.5

The quality of the refined (probably clay-treated) 200°C. E.P. gasoline is reported as 84-86 octane number (I.G. Research).

Detailed cost estimates (based on gas from coke) are given including overhead, interest and licensing. The cost of gas was higher for the Michael process. If coke oven gas were used ($H_2:CO$ 5:1) the cost was reduced but the additional equipment needed to adjust the $H_2:CO$ ratio decreased the favorableness.

Item #16--11-24-39 contains a tabular comparison of the Winkler-Duftscheidt, Klemm, Wietzel, and Michael hydrocarbon syntheses. The results had been obtained on laboratory or semi-technical scale. The main differences in the processes are seen below:

<u>Process</u>	<u>Winkler-Duftscheidt</u>	<u>Klemm</u>	<u>Wietzel-Schäuermann</u>	<u>Michael</u>
$H_2:CO$	45.55	2.5:1	2:1	1:1
Catalyst	Iron (NH_3 Cat.)	Iron	Cobalt	Iron
Pressure - atm.	20	100	12	20
Temperature - °C.	350	290	190	300-310
Operation	Liquid phase with gas recycle	Pure Gas Phase	Gas Phase	Gas Phase

Various supply problems are discussed and remedies suggested such as blending heavy naphtha with 5% motor oil for use as a diesel fuel.

Item #17, 6-11-40, compares the Michael and the Wenzel-Winzer-Wirth processes. The essential difference is that the Michael process uses 100:1 gas recycle for heat removal and the W-W-W process uses a tubular reactor with diphenyl cooling.

Items #18 and 19 discuss results with iron catalysts in tubular reactors and with hot gas recirculation. The information given does not bear complete translation but the following points are of interest. The low operating temperatures which had been experienced at Leuna over those at Oppau could be traced back to lower catalyst reduction temperature. Precipitated catalysts gave a predominance of even chained paraffins. The products from the Michael process contained 0.5% acetaldehyde.

Item #20, 7-17-40, is of little interest except for the description of projected wax refining procedure. The lower melting point material was to be removed by vacuum distillation and the bottoms dissolved in xylene. The hard paraffin is centrifuged from the solution and the xylene removed by evaporation. The xylenes are removed from the middle fraction by vacuum evaporation.

Item #21, 8-7-40, contains a complete result of a life test of 153 days on a cobalt catalyst. It is stated that when Al_2O_3 is used as a support it should be glowd at over 1000°C. and that it is important that the oxygen content should be in the ratio of $CoO:CO_3O_4$.

Item #22, 9-19-40, contains the statement that fused iron catalysts produce a more highly branched product than do precipitated catalysts.

Item #23, 11-30-40, is a table comparing the various hydrocarbon synthesis processes. This table is attached.

83) Chart Comparing Various Catalysts for Hydrocarbon Synthesis From Water Gas (11-30-40)

No.	Kind of Synthesis	Pressure Atm.	Catalyst	CO:H ₂	Maximum Sulfur Content Mg/M ³ Gas	Synthesis Temperature °C.	Process	Heat Removed By:	Main Reaction With Heat Per M ³ of H ₂ + CO	Results of the Synthesis	
										Commercial Maximum Catalyst Chamber Size	Capacity Per Day - Tons Production
1	Gasoline Synthesis Fischer-Tropsch	0	Co/Kieselguhr and Oxides	1:2	4. 3 mg	175-215	Once-through 2- stages	Water Evapora- ting from Cham- ber Wall	$CO+2H_2 \rightarrow CH_4+H_2O +718$ K cal.	10 m ³	3.5
2	Pressure Gasoline Synthesis- Holten	7	Co/Kieselguhr and Oxides	1:2	3 mg	180-210	"	"	"	10 m ³	3.5
3	Gasoline Synthesis of Michael	20	Sintered Iron	1:1	3 mg	>300	Gas Recircula- tion 2-stages	Gas Recircula- tion	$2CO+H_2 \rightarrow CH_4+CO_2$ $CO+2H_2 \rightarrow CH_4+H_2O$ <718 K cal.	4 m ³	3.5
4	Gasoline Synthesis of Dufresne	20	Fused Iron	1:1	0.2-0.3 mg	300	Liquid Phase Liquid & Gas Recirculation	Liquid Phase With Oil Recycle	" "	200 liters	0.84
5	Olefin Synthesis - Leuna	19	Fused Iron	1:1	20 mg	210-230	Once-through 2- stages	Water Evapora- ting from Wall of Chamber	" "	100 liters	0.14
6	Olefins + 40% Alcohol - Leuna	19	Precipitated Iron	1:0.8	20 mg	200-220	"	"	$2CO+H_2 \rightarrow CH_4+CO_2$ $CO+2H_2 \rightarrow CH_4+H_2O$ $7CO+14H_2 \rightarrow C_2H_5OH$ $+6H_2O+638$ K cal.	3 liters	0.003
7	Hard Paraffin Synthesis - Leuna	12	Co-Al ₂ O ₃ - ZnO	1:2	3 mg	180-210	"	"	$CO+2H_2 \rightarrow CH_4+H_2O$ $+718$ K cal.	10 liters	0.003
8	C ₄ -Alcohol Process - Leuna (Isobutyl Alcohol)	250	ZnO-Cr ₂ O ₃	1:2.3		400-450	Gas + Product Recycle		$4CO+8H_2 \rightarrow C_4H_9OH+3H_2O$		
9	Methanol Synthesis Process- Leuna	250	ZnO-Cr ₂ O ₃	1:2.3		360-400	Gas Recycle Without Product		$CO+2H_2 \rightarrow CH_3OH$		

ALCOHOL SYNTHESIS FOR COMPARISON

Translated from T.A.C. Microfilm Reel #13 (11/30/40)

COMPARISON OF THE BASIC RESULTS OF THE HYDROCARBON SYNTHESIS FROM CO + H₂*
 ALL VALUES REFER TO NORMAL CUBIC METER

Main Reaction With Heat Per M ³ of H ₂ + CO	Results of the Synthesis		Yield in Kg/1000 Nm ³ of Reaction Gas						Hard Paraffin M.P.100°C.	Catalyst Capacity Ton Product/ M ³ Cat./Day	Necessary After- Treatment of Product	Economics - Used		
	Maximum Catalyst Chamber Size	Capacity Per Day - Tons	Technically Possible Yield	Practically Obtained Yield	C ₃ + C ₄ Gasoline	200°C. Gasoline	200-300°C. Diesel Oil	300°C. Paraffin				Per Ton Primary Product	Theoretical Coke Con- sumption-Tons	Theor. Coke Needed For Oxy. Removed From Gas
CO+2H ₂ → CH ₂ +H ₂ O +718 K cal.	10 m ³	3.5	160	147	18.1% Olef. 50%	50.9% 35%	18.1% 18%	12.9% 18%	3%	0.353	77.5% Cracking and Distillation; 22.5% Polymer.	6800	3.92	2.71
	10 m ³	3.5	160	147	6.5% Olef. 40%	39.3% 18%	23.4% 10%	30.8% 10%	7%	0.353		70% Cracked, 6.5% Poly. Distilla- tion.	6800	3.92
	<u>Experimental Plant Yield</u>													
2CO+H ₂ → CH ₂ +CO ₂ CO+2H ₂ → CH ₂ +H ₂ O <718 K cal.	4 m ³	3.5	160	149	31% Olef. 85% Alcohol	54% ----- 7.5%	13.5% ----- 50%	1.5%		0.882	48% Dehydrated 31% Poly. Dis- tillation.	6710	3.87	2.66
" "	200 liters	0.84	160	150 (130)	15% Olef. 75% Alcohol	45% ----- 10%	20% ----- 40%	20%		4.24	40% Dehydrated 20% Cracked Poly. Distilla- tion.	6670	3.85	2.64
<u>Semitechnical Plant Yield</u>														
" "	100 liters	0.14	170	155	18.1% Olef. Alcohol	45.8% 48% 35%	19.0% 37% 26%	17.1% 20% 5%		1.41	Distillation Extraction	6440	3.715	2.505
<u>Experimental</u>														
2CO+H ₂ → CH ₂ +CO ₂ CO+2H ₂ → CH ₂ +H ₂ O 7CO+14H ₂ → C ₂ H ₅ OH +6H ₂ O+638 K cal.	3 liters	0.003	187	170	13.8% Alcohol	55.6% 47.5%	14.7% 34.0%	15.7% 10-20%		1.06	Distillation Extraction	5880	3.39	2.18
CO+2H ₂ → CH ₂ +H ₂ O +718 K cal.	10 liters	0.003	170	150	7.1% Olef.	11.4% ----- 22%	16.2% ----- -----	65.4%	65.4%	0.353	Distillation 11% Cracked 10.6% Poly.	6670	3.85	2.64
<u>ALCOHOL SYNTHESIS FOR COMPARISON</u>														
4CO+6H ₂ → C ₄ H ₉ OH+3H ₂ O			250	Methanol	C ₄ -Alcohol 50%	C ₃ +C ₄ - C ₁₄ 40%	Alcohol	Dimethyl Ether			Distillation Refining.	3670	2.23	1.02
CO+2H ₂ → CH ₃ OH			452					1.5%			Distillation	2100	1.21	0.0

23) Chart Comparing Various Catalysts for Hydrocarbon Synthesis From Water Gas (11-30-40)

No.	Kind of Synthesis	Pressure Atm.	Catalyst	CO:H ₂
1	Gasoline Synthesis Fischer-Tropsch	0	Co/Kieselguhr and Oxides	1:2
2	Pressure Gasoline Synthesis- Holten	7	Co/Kieselguhr and Oxides	1:2
3	Gasoline Synthesis of Michael	20	Sintered Iron	1:1
4	Gasoline Synthesis of Durtschmid	20	Fused Iron	1:1
5	Olefin Synthesis - Leuna	19	Fused Iron	1:1
6	Olefins + 40% Alcohol - Leuna	19	Precipitated Iron	1:0.8
7	Hard Paraffin Synthesis - Leuna	12	Co-Al ₂ O ₃ - ZnO	1:2
8	C ₄ -Alcohol Process - Leuna (Isobutyl Alcohol)	250	ZnO-Cr ₂ O ₃	1:2.3
9	Methanol Synthesis Process- Leuna	250	ZnO-Cr ₂ O ₃	1:2.3

Translated from T.A.C. Microfilm Reel #13 (11/30/40)

FILE NO. D-1681
WPB:1 10/1/45

<u>CO:H₂</u>	<u>Maximum Sulfur Content mg/M³ Gas</u>	<u>Synthesis Temperature °C.</u>	<u>Process</u>	<u>Heat Removed By:</u>
1:2	ca. 3 mg	175-215	Once-through 2- stages	Water evaporating from Chamber Wall
1:2	3 mg	180-210	"	"
1:1	3 mg	>300	Gas Recirculation 2-stages	Gas Recirculation
1:1	0.2-0.3 mg	300	Liquid Phase Liquid & Gas Recirculation	Liquid Phase With Oil Recycle
1:1	20 mg	210-230	Once-through 2- stages	Water Evaporating from Wall of Chamber
1:0.8	20 mg	200-220	"	"
1:2	3 mg	180-210	"	"
1:2.3		400-450	Gas + Product Recycle	
1:2.3		360-400	Gas Recycle Without Product	

CARBON SYNTHESIS FROM CO + H₂*
CUBIC METER

Yield in Kg/1000 Nm³ of Reaction Gas
H₂ + CO

<u>Technically Possible Yield</u>	<u>Practically Obtained Yield</u>	<u>C₃ + C₄ Gasoline</u>	<u>200°C. Gasoline</u>	<u>200-300°C. Diesel Oil</u>	<u>300°C. Paraffin</u>
160	147	18.1% Olef. 50%	50.9% 35%	18.1% 18%	12.9% 18%
160	147	6.5% Olef. 40%	39.3% 18%	23.4% 10%	30.8% 10%
160	149	31% Olef. 85% Alcohol	54% ----- 7.5%-----	13.5% 50%-----	1.5%
160	150 (130)	15% Olef. 75% Alcohol	48% ----- 10%-----	20% 40%-----	20%
170	155	18.1% Olef. Alcohol	45.8% 48% 34%	19.0% 37% 26%	17.1% 20% 5%
187	170	13.8% Alcohol	58.8% 47.8%	14.7% 34.0%	15.7% 10-20%
170	150	7.1% Olef.	11.7% ----- -----	16.2% 22%-----	65.4%
<u>PARAFFIN</u>					
250	Methanol	C ₄ -Alcohol 50%	C ₃ +C ₄ -C ₁₄ 40%	Alcohol	Dimethyl Ether
452	445				1.5%

<u>Hard Paraffin M.P.100°C.</u>	<u>Catalyst Capacity Ton Product/ M³ Cat./Day</u>	<u>Necessary After - Treatment of Product</u>	<u>Economics - Used Per Ton Primary Product</u>		
			<u>Nm³ CO + H₂</u>	<u>Theoretical Coke Con- sumption-Tons</u>	<u>Theor. Coke Needed For Oxy. Removed From Gas</u>
3%	0.353	77.5% Cracking and Distillation; 22.5% Polymer.	6800	3.92	2.71
7%	0.353	70% Cracked, 6.5% Poly. Distilla- tion.	6800	3.92	2.71
	0.882	48% Dehydrated 31% Poly. Dis- tillation.	6710	3.87	2.86
	4.24	40% Dehydrated 20% Crack. 13% Poly. Distilla- tion.	6670	3.85	2.64
	1.41	Distillation Extraction	6440	3.715	2.505
	1.06	Distillation Extraction	5880	3.39	2.18
65.4%	0.353	Distillation 11% Cracked 10.6% Poly.	6670	3.85	2.64
		Distillation Refining.	3870	2.23	1.02
		Distillation	12100	1.21	0.0

Items #24-30 contain discussions of various phases of concurrent hydrocarbon and alcohol synthesis using iron catalysts. Such phases as construction, process outline, gas procurement, catalyst plant, product quality and product uses are discussed. The only points which the reviewer deems worthy of summary are: The alcohols in the aqueous phase are recovered by distillation and those in the hydrocarbon phase (particularly above 200°C.) are fractionated into the appropriate boiling ranges, esterified with boric acid, the hydrocarbons are distilled overhead and the ester residue split with hot water and the boric acid recovered by chilling. Dehydration of the alcohols and acids in the gasoline by treatment with Al_2O_3 causes a partial shift of the double bond to the center of the molecule with increased antiknock properties. The edible fats produced by oxidation of the paraffins (320-450°C.) are physiologically the same as natural fats. The production of edible fats from paraffin produced from coal is not recommended because the traces of aromatics might incite cancer.

Items #30-33 contain very little significant information, though the graphical representation of the solubility of monoolefins in an ethanalamine-caproic nitrate solution in Item #31 might be of some interest.

Item #34 - Production of Paraffin from Water Gas over Iron Catalyst (Experiments) 3-25-41. This item contains a description of the liquid phase or bubble process ("Schaumfahrweise") developed by Michael. This is translated in part -

Just as for the gasoline production, gas recirculation was developed, so during the later experiments, it was found that the gas oil synthesis should be carried out in a liquid phase with colloidal catalyst. This process has other advantages besides simplicity, such as a large decrease in gas and an increase in gas oil yield over that with gas recirculation. This process also improves the quality of the gasoline. The research (I.G.) octane number of the refined gasoline boiling up to 190°C. was 90, whereas it was 84 from the gas recycle operation. The capacity of the process per chamber is slightly better than with the gas phase operation. At 250°C. with the middle oil or gas oil process, it is about 0.4 kg/liter/day, at 280°C. it was 0.8; and by the gasoline process, it was 0.8-1 kg/liter/day at 310°C. The catalyst was suspended in a liquid pool contained in about 60% of the chamber.

Comparative product distributions are:

	Gasoline Synthesis at 310°C.	Middle Oil & Paraffin Synthesis at 250°C.	
		Cat. X	Cat. II
% Gasoline 200°C.	60	30	20
% Middle Oil 200-350°C.	25	30	30
% Paraffins 350°C.	15	40	50

The olefin content of the 250-350°C. fraction from the middle oil synthesis was 50-60%. The $C_3 + C_4$ fraction from the liquid phase operation was 4%, gas 10%.

The product water from both processes contained 4-6% alcohols. The oil contained a few per cent oxygen. The alcohol yield could be raised by increasing the pressure. The $C_1 + C_2$ yield was below 5% for the liquid and above 5% for vapor-phase operation. The calculated poly-stage yields were 160-170 gm/m³ of ideal gas.

The 250-325°C. fraction from the middle oil synthesis was suitable for direct sulfonation or for production of fatty acids by inclusion of CO. A third of the paraffin fraction was in the oxidation boiling range. The oxidation of the 250-325°C. fraction produced high pressure oils for leather and varnish for book printing. By careful cracking of the paraffins, it was possible to obtain a 60% yield of 230-350°C. material containing 50-70% olefins.

Items #35-46 contain further discussions on the iron catalyst hydrocarbon and alcohol (synol) synthesis. The phases covered are reactors used, operating conditions, fuel production, product distribution, product analyses, alcohol recovery, carbon adsorption of the light hydrocarbons (gasol), and CO₂ removal by high pressure water washing. The results had apparently all been obtained on an experimental scale and only a few points are of particular interest. The types of reactors which were used were tubes with heat transfer to diphenyl-diphenyl oxide or to high pressure water, large reactor with shallow catalyst bed for the hot gas recycle process, and large reactors containing a liquid pool with agitation obtained either mechanically or by the foaming action of the fine gas bubbles. The catalyst for the liquid phase operation was 1-2 microns in size and was produced by decomposition of iron carbonyl. For alcohol production the H₂:CO ratio was 0.7-0.8:1. The statement is made that precipitated catalysts gave more higher alcohols but were not as stable as the fused. The gasoline refining consisted of clay treating at 300°C. and atmospheric pressure using four liters of clay (which could be regenerated) for one liter of gasoline. Alcohol recovery was accomplished by fractionation, boric acid esterification and distillation. The analytical method for determining alcohols in the products (hydrocarbon phase) was fractionation followed by determination of the hydroxyl number of each fraction by esterification with acetic anhydride and pyridine.

Item #47, 7-17-41, consists of explanatory notes on a Fischer-Tropsch report which had been given elsewhere.

Item #48 - Meeting on Carbon Monoxide-Hydrogen Synthesis 7-1-41. This item consists of discussion on various iron catalyst processes, the Michael hot gas recycle, Michael liquid phase, oil recycle (Dr. Dufschmidt, Oppau) and fused iron in tubes (Dr. Wenzel, Merseburg). Mention is also made of the use of cobalt and of cobalt-aluminum-zinc oxide catalysts for wax production. A few additional points of interest are brought out. The gasoline from the gas recycle process is 84-85 octane no. (I.G. Research) after deoxidation with clay and after further treatment with bleaching earth and addition of alpha-naphthol it is storage stable. In the original reactors the flow from the small recycle conduits into the wider catalyst chamber resulted in an undesirable gas whirl. For the recycle (gasoline) process strong sintered iron catalysts were used (precipitated cat. too friable) and in the liquid phase (gas oil) process iron from iron carbonyl was used after grinding in oil. The diesel oil from the recycle process was 50-55 cetane number and 60-70 from the liquid phase operation. The E₁ fraction from gas recycle was 60-65% iso. The oil recycle process of Dr. Dufschmidt operated with fused catalysts and heat removal was accomplished by the boiling of the relatively light oil fractions. The experiments covered operation at 25-180 atm. At 180 atm. a large amount of low boiling and only a small amount of higher alcohols were produced. The alcohols decreased at high conversion and with carbonization of the catalyst. Dr. Wenzel at Merseburg obtained a product containing 60-65% alcohols in the 100-380°C. fraction. These operations were in tubes with a fused iron (NH₃) catalyst. It is stated that in order to obtain high alcohol yields it is necessary to operate at low temperature and low contact time as the alcohols may decompose.

Items #49-56 contain discussion of miscellaneous phases of the iron catalyst processes such as contact discussion; preparation and reduction of fused catalysts (no details); description of a proposed large scale synol plant; detailed analyses of fractions for alcohols, ketones, aldehydes, acids, olefins; gas procurement; discussion of oxidation of higher alcohols to acids during distillation; and results of experiments on an iron catalyst designated as F1-1249 (no analysis given). In the opinion of the reviewer no additional pertinent information is included.