FILM STUDY GROUP

SUBJECT INDEX AND REPORT

T.O.M. REEL NO. 77

Prepared by

PHILLIPS PETROLEUM COMPANY

Report 605-47P

PHILLIPS PETROLEUM COMPANY

BARTLESVILLE, OKLAHOMA

REPORT ON
UNITED STATES TECHNICAL OIL
MISSION MICROFILM REEL 77

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SUMMARY

This report on United States Technical Oil Mission Microfilm Reel No. 77 is a part of the American Petroleum Institute - Technical Oil Mission program devoted to the indexing, abstracting, and translation of the German literature collected by the Technical Oil Mission. Its distribution outside Phillips Petroleum Company includes all companies participating in the program and the Library of Congress. From the Library of Congress, photostat or microfilm duplication is available to anyone making a request.

This reel is concerned primarily with the operation of, and problems encountered at, Hydrierwerke-Politz. It includes data and information on the CV2B and DHD processes and products obtained therefrom, Alkazid process, license agreements, proposed toluene plant (T 16), and gasoline inhibitors. Numerous analytical methods are described in the latter portion of the reel.

Albert E. Miller, Chairman
A.P.I. - T.O.M. Study Group.

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ALKALINITY OF CV2B GASOLINE

Frames 6-15

Abstract

With regard to a complaint by the Research Center of RLM for Aviation Fuel on the alkaline behavior of CV2B gasoline from Gelsenberg Benzin A.G., I.G. Farbenindustrie points out that 7019 gasoline may contain aromatic amines resulting from the incomplete reduction of organic nitrogen and ammonia or by high pressure reaction of benzene and ammonia. However, tests show that those alkaline compounds do not corrode metals used in motors or unfavorably affect the storage stability of the gasoline. Furthermore, it is well known that aromatic amines, such as aniline and methyl aniline, are antiknock agents in concentrations of 2 to 6% and they also improve storage stability. Also, amines are used as gum inhibitors in the United States. Ill effects from the low concentrations (.1%) of organic bases in 7019 aromatized gasoline is therefore not feared.

STUDY OF SOLID AROMATICS FROM SLUDGE OF PITCH HYDROGENATION AT POLITZ

Frames 21-23

Abstract

The sludge contained:

49.6% solids (benzol insoluble)
50.4% oil
14.6% ash in solids
8.8% asphalt in oil

The following contents of hydrocarbons known to be especially stable under conditions of the technical hydrogenation process were found:

Carbozole - 0.2% of total sludge
Pyrene - none (this is usually present to the extent of 10-15% in oil from low temperature coal carbonization)
Coronen - 1.5% (Coronen melts at 432° and boils at 525° at 760 mm.)

It is concluded then that pitch sludge shows no special enrichment in these known stable hydrocarbons.

COAL - SLUDGE PROCESSING

Frame 24

Abstract

In the time from July 24-30, 1941, the following quantities were processed and obtained:

Coal (TK_2) 2363.087 tons = 14.1 tons/hr.

Coal sludge 3515.927 tons = 21.0 tons/hr. (25.4% solids)

Centrifuge mixture 5125.063 tons = 30.5 tons/hr. (16.4% solids)

Centrifuge oil 3999.164 tons = 23.8 tons/hr. (11.3% solids)

Centrifuge residue 1125.899 tons = 6.7 tons/hr. (39.5% solids)

Carbonization of the centrifuge residues gave the following:

Charge: 1125.899 tons = 2.23 tons/hr./oven
containing 446.574 tons solid and 679.325 tons oil

Production: 560.459 tons precooler oil = 3.35 tons/hr.
22.938 tons after cooler oil = 0.14 tons/hr.
583.397 tons total oil

Carbonization yield: 86.1%

Total loss in processing residues: 95.928 tons
= 4.1%

CENTRIFUGE PLANT - BUILDING 275, NOVEMBER, 1941

Frames 25-27

Abstract

In the month of November many centrifuges dropped out of operation simultaneously and suddenly. A centrifuge which had been running cleanly would stop completely in a few minutes. Four factors were considered in finding the reason for these interruptions:

1. The asphalt content. This had decreased from 7-8% to 4-5%. However, since good operation has been obtained with lower asphalt content, this factor was considered to have no effect.

2. The size distribution in the coal paste. This was concluded to have no effect on the running of the centrifuges, as the processing of coarser coal pastes and sludges failed to prevent the clogging of the centrifuges.

3. Temperature of the mixing cycle. It was found that this is important but not the determining factor. Operation was impossible as low as 130°C, poor at 145°C, better at 155-160°C, and 170-180°C would be desirable. The oil preheater was not ready for operation, and consequently this temperature could not be reached.

4. Nature of the pre-dilution oil. This was proved to be the decisive factor. An oil with an initial boiling point of 280°C inevitably resulted in trouble; one of 305°C initial was better; and the best was a catch pot heavy oil of 320°C initial. The explanation is that the light fractions of such an oil precipitate asphalt which entrains solid particles and deposits on the basket and nozzles of the centrifuge. The asphalt precipitation properties of the various oils were determined by tests. (Frames 29-30).

OIL RECOVERY FROM WASTE GAS

Frame 28

Abstract

Recovery of oil from the off gases of the centrifuge residue tanks is considered uneconomical because of the small amount recoverable and because cooling to -20°C is required. Oil scrubbing was ineffective.

DISTILLATION OF COAL SLUDGE

Frames 32-33

Abstract

Beginning in July, 1943, the coal sludge was distilled in the vacuum sludge distillation equipment. The process is operable and practical when the distillate approaches 42.9%. The softening point of the residue averages about 78.7°C. A calculated comparison of the clear oil loss between processing the coal sludge by centrifuging and carbonization on the one hand, and vacuum distillation on the other, shows a loss of 4.2% oil for the former and 15.7% for the latter method, when reduced to a common basis. Data for the comparative runs are given.

IRON SULFATE IN HEAVY OIL

Frames 34-36

Abstract

Settling tests of iron sulfate in catch pot heavy oil.

CONTRACT.

Frames 53-58

Abstract

Five Year Contract between RLM (Reichminister for Aeronautics) and Ammoniawerke Merseberg (Leuna) for the production of T52 Aviation Fuel. So far as isobutane yield from the hydrogenation plant and receipt from outside sources will permit, the production is to be 19,000 tons per year.

Properties of T52:

Density at 20°C = 0.700 to 0.725 kg/liter

Initial boiling point (Engler) = not under 85°C

Endpoint = up to 220°C

Reid vapor pressure (38°C) = 0.20 to 0.23 atm.

Octane No. (CFR) = Ca 96

Price is to be based on 20% amortization of RM 15,000,000 of outside capital and RM 10,500,000 of the company's own capital. The cost of operation of the T52 process is to be determined each year. A preliminary cost estimate is given.

LICENSE AGREEMENT

Frames 59-68

Abstract

License agreement between I.G. Farbenindustrie and Hydriewerke-Politz for the use of a process for the production of high antiknock fuel. The process utilizes gases produced by hydrogenation, or others containing C_4 hydrocarbons, and comprises the steps of dehydrogenation, polymerization and hydrogenation (T52 process). Where isobutane is used as raw material, the product has the following properties:

Octane Number (CFR Motor Method)	96
Specific gravity	0.700 - 0.720
Engler Distillation	
initial	below 85°C
at 130°C	not over 30% overhead
end point	not over 200°C
Reid Vapor Pressure (38°C)	0.20 - 0.23 atm. abs.

I.G. agrees to provide all technical know-how and the benefit of all future improvements and gives a non-exclusive, non-transferable license. The fee is 1.2 pfennig per kilo of T52. This is reduced 10% for production in excess of 20,000 tons/year and is reduced an additional 10% for production in excess of 30,000 tons/year. A grant-back of exclusive license to improvements made by the licensee to I.G. is provided.

The license does not cover the preparation of the necessary catalyst, which the licensee obligates himself to purchase from I.G. at cost plus normal profits. Licensee will observe strictest secrecy. I.G. will prepare and prosecute all patent applications.

Contract term 12 years.

LICENSE AGREEMENT

Frames 69-76

Abstract

License agreement between I.G. Farbenindustrie and Hydrierwerke-Politz for DHD process plant for the treatment of gasoline made by hydrogenation, with hydrogen, under pressure at temperatures above 400°C in the presence of catalysts that are periodically regenerated. The gasoline is aromatized (made poorer in hydrogen).

The licensee can enlarge its DHD plant or produce products other than fuel only by agreement with I.G.

The license does not cover the catalyst, which licensee agrees to purchase from I.G. However, if licensee invents new patentable catalysts he is entitled to manufacture them.

Fee - 0.7 pfennig per kg. scaled down with increased production as follows:

100,000 - 200,000 tons/yr.	0.63 pfg.
200,000 - 300,000 tons/yr.	0.56 pfg.
over 300,000 tons/yr.	0.50 pfg.

Licensee agrees to turn over to I.G. all its inventions and experience related to the process.

LICENSE AGREEMENT

Frames 77-93

Abstract

License agreement (3/29/1939) between I.G. Farbenindustrie and Hydrierwerke-Politz covering I.G.'s hydrogenation rights, including present and future rights of its direct and indirect licensees, such as Hydro Patents Co. (U.S.) and International Hydrogenation Patents Co. (The Hague) for the production of fuel including LPG, but excluding isooctane. I.G. agrees to train operating staff and impart all knowledge and experience gained from its own and its licensees' plants. The licensee agrees to reciprocal imparting of experience.

The license does not cover preparation of the necessary catalysts, which the licensee agrees to purchase from I.G., unless he develops a new and patentable catalyst. Licensee agrees to purchase the necessary liquor for the Alkazid plant from I.G., who agrees to furnish it on the same basis as he does to his other German licensees.

License fees	On the basis of coal	On the basis of crude oil
Motor gasoline	0.65 pfg/kg	0.60 pfg/kg
Aviation gasoline	0.75 pfg/kg	0.70 pfg/kg
L.P.G.	0.50 pfg/kg	0.45 pfg/kg
Diesel fuel	0.45 pfg/kg	0.40 pfg/kg

High octane gasoline, with better than 87 octane number, commands a higher fee according to its higher value. This will be arranged by agreement from time to time. With production in excess of 200,000 tons/yr., the fees are scaled down as follows:

200,000 - 300,000 tons/yr.	10%
300,000 - 400,000 tons/yr.	14%
400,000 - 500,000 tons/yr.	17%
500,000 - 600,000 tons/yr.	20%
600,000 - 700,000 tons/yr.	22%
700,000 - 800,000 tons/yr.	24%
800,000 - 900,000 tons/yr.	26%
900,000 - 1,000,000 tons/yr.	28%
over 1,000,000 tons/yr.	30%

Licensee grants I.G. an exclusive, transferable, royalty-free license to all of his developments and improvements of the process.

A "favored nations" clause is included.

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NOTES ON ALKAZID PROCESS

Frames 95-98

Abstract

Letter dated February 19, 1942, from Politz to Dr. Kimmerle, Leuna, replying to the following questions asked by Dr. Kimmerle:

Is it necessary to wash the lean gases from the two coal hydrogenation phases with water before passing to the Alkazid plant at 25 atm. in order to avoid stoppages in the pipe line due to ammonia salt deposition?

What is the ammonia content of the mixed gases?

Politz operated the Alkazid plant on the combined lean gases from the sump and gas phases without expansion. At first they experienced some difficulties from stoppages, but corrected this by injection of a small quantity of water at the most important places. This water is withdrawn from catch pots and the spherical gas holder ahead of the Alkazid plant. No stoppage has been experienced even in November.

In the total lean gas, the ammonia content is scarcely 1 g. per cu.m., even before water is injected.

Report 605-47P

DIRECTIONS FOR THE OPERATION AND CONTROL OF ALKAZID SCRUBBING PLANTS

Frames 101-102

Abstract

A. Treatment of the Aluminum Components of the Plant When Shut Down

Sometimes punctiform corrosion occurs in the aluminum parts of the Alkazid plant when it is not in operation and air has access to it. It is suspected that such corrosion centers as are formed during a shutdown can lead to progressive corrosion and serious damage to the apparatus during subsequent operation.

The following measures are taken to avoid such corrosion when the plant is shut down for an extended time, particularly when it is necessary to open it so that air can enter:

1. After emptying the solution, carefully and thoroughly rinse with distilled water.
2. Dry the aluminum apparatus thoroughly with warm air, so that no adhering moisture remains and protect the apparatus from "sweat" moisture which may collect as a result of atmospheric changes. The outside surface is also endangered by adhering water. Under insulation, the corrosion centers sometimes penetrate the walls completely. Exterior corrosion is avoided by keeping the apparatus hot or by protective coatings.

B. Removal of Corrosion Damage to Aluminum Parts

In the case of severe corrosion which makes a part unserviceable the part is either completely replaced or is welded or patched.

In very many cases in which smaller corrosion grains have appeared but the part is still serviceable, the grains may be removed and a surface which will be passive in further operation may be renewed by the following procedure:

In accessible places:

1. dry the surface
2. brush the surface vigorously with a steel brush or scrape off the corrosion product until a bright metal surface remains, being careful to avoid leaving steel splinters sticking in the aluminum, and to obtain a smooth surface free from scratches.

In difficultly accessible places such as heat exchangers:

1. rinse well with distilled water
2. let stand over night in contact with nitric acid chromium etching solution which has the following composition:

178 cc. HNO_3 (sp. gr. 1.39) and 49 g. $K_2Cr_2O_7$ in 1 liter solution.

3. rinse again well with distilled water
4. dry thoroughly

DIRECTIONS FOR CONTROL OF ALKAZID PLANT

Frames 103-105

Abstract

I. Method of Operation

The Alkazid process is a wet process for the removal of H_2S or CO_2, or both, with the help of certain salts which have the property of absorbing these gases in the cold and giving them up again unchanged when heated.

The technical operation of the process is carried out in such a manner that the Alkazid caustic is carried through a wash tower and a degasser (regenerating column, or separator) in a constant cycle with the steps as follows (Fig. 1).

The regenerated caustic (fresh solution) enters the top of the scrubbing tower and flows countercurrent to the gas being purified. The saturated solution is pumped from the bottom of the scrubbing tower through a heat exchanger to the regeneration column. In the heat exchanger, it is warmed indirectly by the regenerated solution, which is cooled thereby.

The saturated, preheated caustic flows through the regenerating column from top to bottom while being heated further and is finally boiled and blown with steam in the lower part of the column. The regenerated gases leave the top of the column in admixture with steam, which is separated in a cooler and the condensate returned to the system.

The hot regenerated solution is pumped through the heat exchanger and finally through a spray-cooled coil where it is reduced to the absorber temperature.

II. Directions for Operation

A. Transport and Storage of Alkazid Solution

Handle as a concentrated aqueous solution:

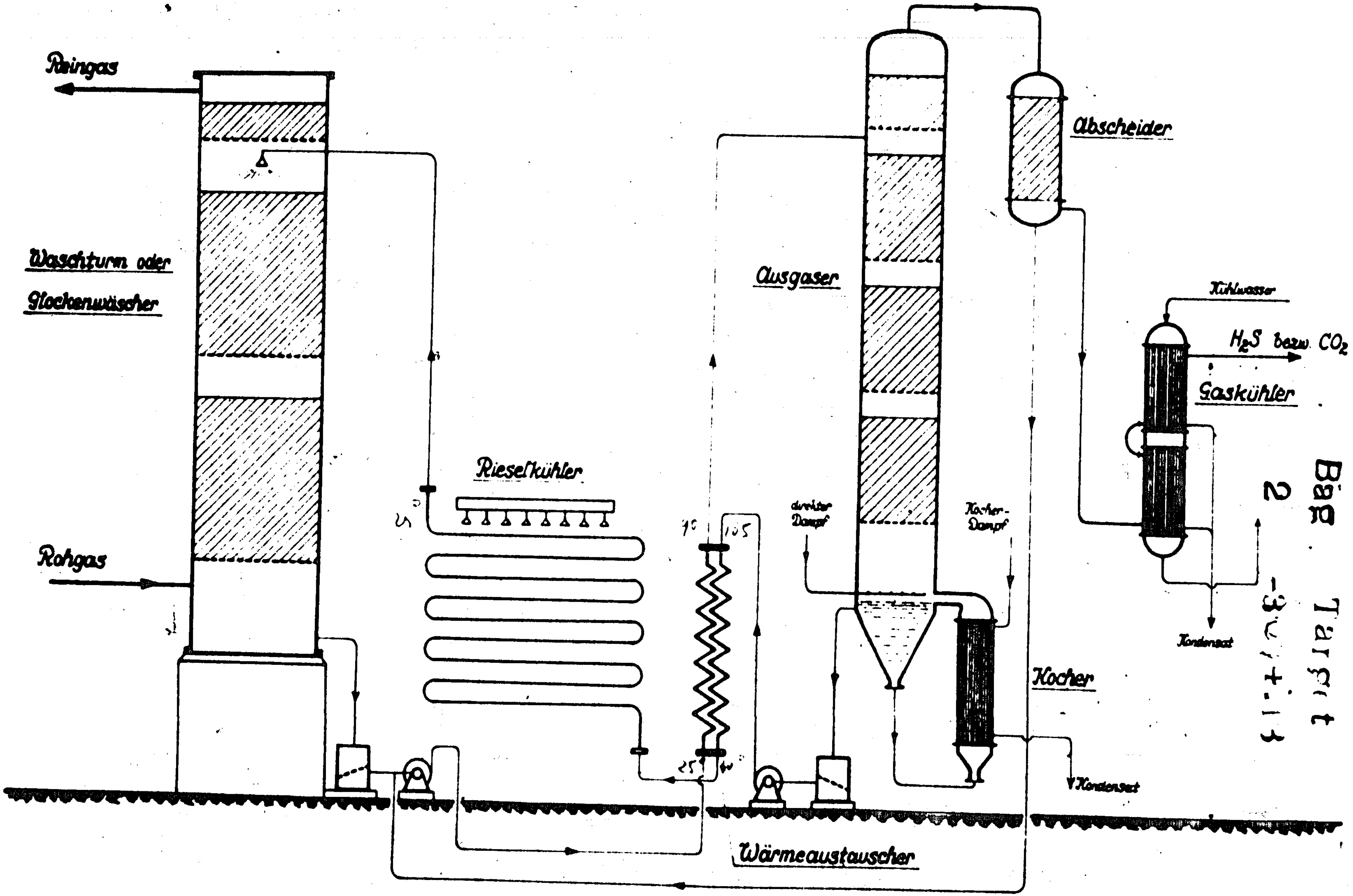
1. At low temperatures salt can crystallize out
2. At very high temperatures or very long storage small quantities of gas may be liberated.

B. Operation

For aluminum-containing plants, points 1-3 are very important.

Abb. 1

Schema einer Alkacid-Waschanlage.



1. A solution density of 1.16 at 20°C (degassed) should be maintained. It is desirable to avoid dilution because when the density falls much below this, there is some danger of the solution attacking the aluminum. If, however, a corrosivity towards aluminum develops, it is easily overcome by adding water glass (3 g. dry sodium silicate per cu. m. of solution is dissolved in a ten fold volume of solution and injected slowly into the cycle).

2. If aluminum parts are present in the plant, the system should be vigorously washed with a very dilute water glass solution by circulating for several hours. A solution of 0.1% water glass in condensate is suitable, and the temperature can be carried as high as 60°.

3. Metallic mercury should be scrupulously avoided because of its rapid attack on aluminum. Air must be kept out of the solution system.

4. Dust cannot be avoided, so a continuous or intermittent side stream may be removed and filtered to separate sludge.

RESULTS OF CORROSION EXPERIMENTS WITH IRON AND ALUMINUM (10-21-44)

Frames 107-111

Abstract

Composition of fresh commercial Alkazid Solution (caustic) is as follows:

Specific gravity	1.201
Potassium (K)	143 g./l.
CH_3CHNH_2COOH (Alanin)	396 g./l.
CO_2	22 g./l.
Calc.	69 g. K_2CO_3/l.
	155 g. KOH/l.
	396 g. Alanin/l.

The technical solution was diluted to a specific gravity of 1.16 and was used then for the absorption of H_2S and CO_2. The aminopropionic acid (Alanin) has weak acid and basic properties. When titrated with N/10 KOH at room temperature with phenolphthalein as an indicator, 1 g. Alanin only consumes 3.2 cc. N/10 KOH whereas the theoretical consumption approaches 112 cc. When titrated at 90°C, however, the consumption is 9.4 cc. N/10 KOH, which is also of a lower order of magnitude than the theoretical. Alanin is thus a much weaker acid than the other weak acids (acetic). It cannot be titrated electrometrically. When titrated with HCl toward methyl orange, 1 g. Alanin consumes 4 cc. N/10 HCl.

The Importance of KOH and K_2CO_3 Addition to Alanin and Operating Solutions for the Outgassing

The outgassing is conducted by boiling under a reflux condenser while passing nitrogen gas through the solution.

1. Pure Na_2S loses no sulfide-sulfur while boiling and bubbling N_2 through it.

2. An H_2S-containing Alkazid solution loses its H_2S content very quickly.

3. Addition of KOH makes outgassing much more difficult.

4. Addition of K_2CO_3 makes outgassing somewhat more difficult but not as difficult as an equivalent KOH addition.

5. The higher the Alanin concentration in proportion to H_2S content the more rapid is the outgassing.

6. When H_2S and CO_2 are present together in a solution, H_2S comes off first.

7. On passing H_2S into a pure Alanin solution, sulfur separation results. In the presence of KOH no sulfur separation occurs, although sulfur separation occurs if only K_2CO_3 besides Alanin is present.

8. Glycocoll and phenol behave similarly.

The Corrosion Behavior of Soft Iron

The aggressive components of Alkazid solution against iron are Alanin and sulfide sulfur. (Tests show that a pure Alanin solution attacks iron very strongly at 90°C.)

9. A K_2CO_3 addition to alkazid solution hinders the loss of metallic iron much more than an equivalent KOH addition.

10. An outgassed solution having a higher residual CO_2 content is therefore less corrosive to iron than one with lower CO_2 content.

11. Points 9 and 10 give two methods of preventing iron corrosion, which although they have the same effect on corrosion, differ with respect to steam consumption, because according to (9) the alkali content is increased by K_2CO_3 addition and outgassing becomes more difficult while by (10) the alkali content remains the same.

12. The higher the temperature, the stronger is the attack on iron. There exists a critical temperature at which corrosion sets in. It lies between 40 and 70°C, depending upon conditions.

13. In boundary line corrosion, the greatest attack occurs above the boundary line.

The Corrosion Behavior of Aluminum

The aggressive components of the Alkazid solution against aluminum are sulfide sulfur and alkalinity.

14. Alanin inhibits the solution of aluminum in alkali.

15. The Alanin content is so high that aluminum undergoes no loss of metal.

16. The aluminum corrosion is independent of the dilution of the solution.

Prevention of Corrosion of Aluminum when Plant is Shut Down

The aggressive materials, which attack aluminum in a shutdown plant, are salts, moisture and air. As these materials are practically always present, steps must be taken to protect the plant.

17. A maximum of 1% water glass solution protects aluminum even at 100°C.

18. The presence of foreign salts result in precipitation of silicic acid from water glass. Therefore the aluminum part of the plant should be well rinsed before filling with water glass solution.

OPERATION PROCEDURES AND HINTS FOR THE ALKAZID PROCESS (1939)

Frames 112-130

Abstract

A. Transport and Storage of Solution

Solution is normally non-corrosive, non-combustible and non-toxic. If it contains slight amounts of poisons at first, these are lost after short operation.

1. Avoid extreme temperatures.

2. Carefully vent the closures of containers to air from time to time. Separated salt may be put back in solution by steaming. Dilute solution with pure, distilled water.

B. Putting Plant into Operation

Before putting into operation, eliminate all leaks, test by filling with water, and pressuring. Wash out all dirt, etc. Add water glass to the last water-filling of aluminum parts of the plant (several hours contact). The plant is emptied of all water, and the concentrated Alkazid solution is pumped in cold. It is partially diluted by residual water in the system. This is allowed to circulate in the apparatus until all the air is displaced by deaeration from the sections containing solution. The air may be displaced with CO_2 and the latter displaced with heating gas, if a CO_2-free, inert gas is not available. The air may be displaced from the column by steam. If the degassing column is filled with H_2S or CO_2 from prior operation the solution should be heated before pumping in, to avoid rapid absorption and sucking of solution into the gas lines, etc. Heating of the degassing column should be slow. When the temperature rises to that desired in operation (100°C), the saturation of gas in the scrubbing tower should begin promptly to avoid corrosion.

C. Operation

The most favorable temperature of the solution entering the wash tower is 20-25°C for H_2S and 25-55°C for CO_2. Temperature lower than 20°C should be avoided because of danger of salt separation, particularly from solutions containing CO_2 or H_2S.

A part of the steam in the regenerator is introduced directly and the rest is used indirectly, the proportions being controlled to maintain the proper constant solution concentration. This is controlled by hydrometer tests. Indirect steam runs from 30-250 kg/cu.m. of solution. A specific gravity of 1.20 may be used without salt separation. In aluminum a specific gravity lower than 1.16 should be avoided.

New solutions frequently show a tendency to foam. This can be suppressed by addition of I.G. Antifoam Agent "I" (1 to 2 liters per 10 cu.m. of sol'n).

From time to time control tests should be made as follows:

(a) Gas saturation of the solution

(b) Performance of the regenerator

(c) Stability of the solution

If aluminum is present, it is desirable to test the solution every few days for corrosiveness. At the first sign of corrosion 300 g. water glass should be added per cu. m. Local dilution, as by returning condensate, must be avoided as this will lead to corrosion of aluminum.

Oxygen must be avoided in solutions containing H_2S because of the sulfate formation. High speed suction pumps may suck in air, so those at Leuna are encased in a chamber flushed with nitrogen. Thiosulfate is not regenerated and decreases the absorbing capacity of the solution.

Pump screens must be kept clean. As air oxidizes FeS, and therefore leads to corrosion of the metallic iron, air must be excluded as completely as possible. When shutting the plant down, the degassing column must be run until all the solution in the system is regenerated. Laboratory procedures for testing the solution are given. Apparatus and record forms are shown and an H_2S warning poster is given.

DETERMINATION OF WATER IN COAL PASTE

Frames 132-134

Abstract

The object was to determine how much water was evaporated during the preparation of the coal paste and while bringing it into the high pressure system. Both the xylol and heating-loss methods were used. The former gave higher values for water content. The average of 15 tests gave a water content of the dry coal of 2.37%. The average value for solids in the coal paste in the high pressure plant was 1.99%, a diminution of 16%.

ASH ANALYSES OF COAL CATALYSTS

Frames 136-142

Abstract

I. (Bayer Mass) II. (Ferrous Sulfate) and III. (Sulfigran = Na_2S)

Object: To determine how much ash is found in ashing the three catalysts and how the analytical values for the several mixtures vary from the theoretical when the proportions of II are varied.

Catalyst I: 11.4% H_2O and 83.2% ash (93.18% dry basis)

Catalyst II: 35.94% H_2O and 28.92% ash (45.07% dry basis)

Catalyst III: 0% H_2O and 171.04% ash (Na_2SO_4)

On mixing the three, a heat of reaction (between III and I) is observed, indicating a chemical reaction. Both the moisture and ash determinations of mixtures deviate widely from the theoretical values and the deviations vary with the content of II.

Catalyst I loses 9.7% H_2O in the plant dryer (heating from 30°C to 108°C) and catalyst II loses 37.13% H_2O .

WET CATALYST EXPERIMENT

Frames 144-145

Abstract

As the dry coal and catalyst may not be thoroughly mixed, it was suggested that wet catalyst addition be tried. The $FeSO_4.7H_2O$ dissolved readily in H_2O to give a 30-35% solution. However, the Bayer Mass was very difficult to grind and put into suspension. It was concluded that they had insufficient space for the additional dryer required and that the operation would be too costly.

MEMORANDUM OF A MEETING

Frames 216-217

Abstract

(Hydrierwerke - Politz, A.G., August 22, 1939)

Polymerization Plant for Cracked Gases from the Norddeutschen Oil Works at Politz.

This plant produces 9,000 tons/year of a gas having a heating value of 11,000 cal./kg. (15,000 cal./cu.m.), which is 800 cu.m./hour (12MM cal./hour).

Component	Vol. %
CH_4	16.5
C_2H_6	14.0
C_3H_4	8.0
C_4H_{10}	0.6
C_2H_4	30.0
C_3H_6	17.0
C_4H_8	5.0
i-C_4H_8	0.9
C_5H_{10}	5.0
Higher Olefins	0.85
H_2	1.5
CO	0.2
N_2	0.4
O_2	0.05
	100.0

With 70% conversion of propylene and butylene, hourly production of polymer (gasoline) would be 240 kg. Production costs, which total 328,000 marks for a year's production of 1,920 tons, a schematic flow sheet, and an estimate of the cost of the various plant items, are given. The location of the plant had not been determined.

CORROSION EXPERIMENTS FOR THE COAL CARBONIZATION PLANT

Frames 229-230

Abstract

(Hydrierwerke - Politz, A.G., September 17, 1942)

Results are presented of experiments made to determine the action of various catalysts on iron. Individual catalysts (200 parts) were heated, with stirring, to 220-230°C in admixture with oil or residue (800 parts). The loss in weight of an iron cube indicated the corrosion. Soda was most corrosive, "Sulfigran" was only slightly corrosive, and sulfur was not corrosive at all. Corrosion with the original oil was greater than with the residue. Futurs experiments will determine the relative corrosion on metals used for the reaction chambers as compared with wrought iron, and the influence of corrosion on the electrode materials.

MEMORANDUM ON ORIGIN OF PLUGGING OF TRANSFER LINES AND COOLERS OF BUILDING 280 (OF THE CARBONIZATION PLANT)

Frames 231-232

Abstract

(Hydrierwerke - Politz, A.G., June 30, 1941)

About the end of April, deposits were found in the oil in the precoolers of the carbonization plant which finally caused the plant to be shut down for removal of these solid products. The material was found to be anthracene. Investigation showed that it was produced in the high pressure conversion, and was partially decomposed in the carbonization plant.

Analytical Method for Determination of Crude Anthracene

15 g. of oil is dissolved in 10 cc. of benzene and cooled to 10°C, while being stirred with a thermometer. The mixture is filtered and the vessel washed with 5 cc. benzene. The precipitate is dried for 4 minutes at 105°C and weighed.

CHARGE DATA

Frames 233-234

Abstract

(Hydrierwerke - Politz, A.G., May 9, 1942)

Charge to each of 8 reactors, in tons/hour, for each day from April 1, 1942, thru May 7, 1942. It varies from about 0.5 to about 4.0

EXPERIMENTS WITH AND WITHOUT CATALYSTS

Frames 235-237

Abstract

(Hydrierwerke - Politz, A.G., February 10, 1942)

Experimental data are given for the experimental operation of reactor III without a catalyst (January 16-19, 1942) as compared with reactor IV with the usual catalyst (soda 0.268%, sulfur 0.299%, and Sulfigran 0.299%) during the same period, and as compared with normal operation of reactor III both before (January 12-15, 1942) and after, (January 20-22, 1942) this period. It is concluded that the two

disadvantages of omitting the catalyst are: (1) decreased conversion (from 87% down to 76.5%), and (2) a marked formation of coke which threatened to shut down the reactor.

COAL REQUIREMENTS TO PRODUCE 24,000 TONS/YEAR OF TOLUENE

Frame 312

Abstract

(Hydrierwerke - Politz, A.G., November 17, 1939)

Estimates of the coal requirements are given. The 360,000 tons of light naphtha which are needed, require 800,000 tons of coal. This exceeds the plant capacity of 660,000 tons per year. For the production of power, hydrogen, and the required light naphtha, the ratio of coal to toluene is 5.6/1. This requires 17 tons of coal per hour, or 30 car loads per 300 work days.

A COMMUNICATION ON TOLUENE FROM DR. PIER, OCTOBER 19, 1939

Frames 313-314

Abstract

(Hydrierwerke - Politz, A.G., November 23, 1939)

A tabular estimate of yields for (1) dehydrogenation and (2) aromatization. (1) produces 6,600 tons/year of toluene, and (2) 18,700 tons/year of toluene.

MINUTES ON TWO CONFERENCES RELATING TO T-16 ON THURSDAY, DECEMBER 28, 1939, AT THE MINERALOELBAU AND THE WIFO OFFICES

Frames 315-320

Abstract

(Hydrierwerke - Politz, A.G., January 3, 1940)

At the first meeting, representatives of Mineraloelbau Company, I.G. Ludwigshaven, and Hydrierwerke-Politz discussed a proposed plant T-16, to be built for production of toluene at Hydrierwerke-Politz. Reference is made to drawing N-7897-2 (Frame 321), which is a flow sheet for the dehydrogenation part of a plant (T-16) to produce 24,000 tons/year of toluene by dehydrogenation of the fraction boiling in the range of 90-160°C, from 5058/6434 (catalysts) light naphtha. The extraction plant is given on drawing N-7896-2, not presented. Various problems concerning the securing of equipment are discussed.

The second meeting was in the Wifo Offices in Berlin, Mohrenstrasse 35/37, between representatives of the above three groups and of Wifo. The Wifo representatives were surprised at the proposed cost, 30 million marks, and asked for initial production on July 1, 1941, and full production on October 1, 1941. Procurement of steel was discussed. An estimated breakdown of the cost is included.

INSPECTION DATA ON CV2B GASOLINE (46861)

Frames 337-338

Translation

Specific gravity	0.802
Aniline point I (calculated from mixture with normal gasoline), °C	+ 0.3
Aniline point I (direct determination), °C	- 7.7
Aniline point II, °C	+48.7
Copper strip test	slightly colored
Doctor test	negative
Iodine number (Hanus method)	6.9
Refractive index	1.4550
Octane number, motor	79.0
Octane number, + 0.12% Pb	91.5

Composition

Paraffins	13 wt. %
Naphthenes	31 wt. %
Aromatics	54 wt. % (50 vol. %)
Unsaturates	2 wt. %

Bombtest:

(a) Gasoline unleaded:

Pressure drop	none
Glass dish gum before aging	0.5
Glass dish gum after aging	4.0

(b) Gasoline + 0.12% lead:

Pressure drop	none
Glass dish gum before aging	1.3 mgr.
Glass dish gum after aging	13.0 mgr. (total evaporation residue)
Glass dish gum after aging (leaded)	11.5 mgr. (resin)
Phenol	0.012%

ASTM Distillation

Initial	4°
60°	1.9%
70°	5.9%
80°	11.2%
90°	18.9%
100°	30.1%
110°	45.9%
120°	59.0%
130°	70.9%
140°	81.0%
150°	90.0%
160°	95.0%
End point	166°/97.8%

IMPROVEMENT IN THE STORAGE STABILITY OF CV2B GASOLINE BY INHIBITORS

Frames 339-347

Abstract

(Hydrierwerke - Scholven, December 18, 1940)

Experiments, on improving storage stability of CV2B gasoline, were conducted at Scholven as a result of a conference on October 18, 1940, and a letter from I. G. Ludwigshaven on November 13, 1940.

It is known that condensation products between aminophenols and amines with aldehydes and ketones are inhibitors [Oel u Kohle, 45, 777 (1939)]. Fig. 1 (frame 343) shows results of adding benzyl-p-aminophenol. Pure phenol has no inhibiting value, Fig. 2 (frame 344). Addition of products removed from the caustic alkali wash improves the bomb test markedly, Fig. 3 (frame 345). Similar results were obtained by Henderson, Ross and Ridgeway [Ind. Eng. Chem., 31, 17 (1939)]. By repeatedly using the caustic alkali solution, it became saturated with inhibiting materials and no longer removed them from the gasoline although more strongly acidic substances, such as mercaptans, continued to be removed. Fig. 4 (frame 346) shows the results of two experiments. Two batches of CV2B gasoline, 48 liters and 68 liters, which had been washed free from H_2S with a soda solution, were washed in 1 liter portions with 100 cc. of a 4% NaOH solution. At first the bomb test was poor, but then it improved. The one to which phenol was added (0.05-0.1 gm./liter) was the best. According to these tests, it is clear that improvement will result in adding plant phenol to the gasoline.

This is explained on the basis that various phenols of various acidities are initially present. Addition of plant phenol allows the weakly acidic, inhibiting phenols to remain, unremoved, in the gasoline.

In plant practice, stabilized gasoline is washed with caustic alkali (10-12%) to remove residual traces of H_2S, and the wash solution is left in use until a trace of phenol appears in the washed gasoline. A portion of the old alkali is removed and fresh is added to maintain a phenol content of about 0.1 gm. phenol per liter. This requires considerable care in control, but yields a CV2B gasoline with greater oxidation stability.

AROMATIZATION OF BITUMINOUS COAL LIQUEFACTION MIDDLE OIL OVER CATALYST 7019 (at 250 atm.) FOLLOWED BY AFTER-REFINING CATALYST 7360

Frames 353-370

Abstract

Summary

Bituminous coal middle oil from Scholven was aromatized by passing it over used catalyst 7019 from Chamber 501 (used 1/2 yr.) in the 1-liter furnace followed by 0.33 liter 7360 as a refining catalyst with gas recycling. The test ran 1,000 hr. (6 weeks) without noticeable diminution of activity. At a 365°C refining temperature there was obtained, without additional loss and without decrease in aromatic content, a gasoline with better iodine number (1-3.4 as compared with 11.6 without refining), better oxygen bomb test, and better copper dish test. On the basis of 8 determinations of the bomb test and 7 copper dish tests with lead, one was not entirely without objection. This can be corrected by slight addition of inhibitor.

The octane number of the gasoline was the same as without the use of the refining catalyst. To further confirm these results in the 1-liter furnace, an aromatization test was carried out with new catalyst 7019 in a 6-liter furnace, with subsequent treatment with 7360 (2 liter) as a refining catalyst. This test also gave a CV2B with normal aromatic content, and at suitable refining furnace temperature, lower iodine number, good bomb test, and improved light stability.

The gasoline output during the test of 700 hr. of operation was about 0.25 of the 7019. Toward the end of the test, the output lay progressively lower; this should not be attributed to diminution of catalyst activity but to recycling of the distillate from the catch pot due to abnormal operation of the column.

According to this test in the 6-liter furnace, by operating a new aromatization chamber with a refining furnace it should be equally possible to obtain a good bomb test if the temperature of the refining furnace is kept as low as possible (under 305°C and if possible 255°C). It should be raised slowly when the iodine number exceeds 3. The bomb test of a product refined at too high a temperature was improved by addition of phenol (0.02%).

DHD EXPERIMENTAL DATA

Frames 372-377

Abstract

DHD test with liquid phase gasoline boiling up to 165°C and 5058 prehydrogenation gasoline boiling up to 165°C (1941).

As a shortage of charge stock existed at the time of the reconstruction of the DHD plant, tests were made to determine whether liquid phase gasoline, or 5058 gasoline without removal of its lower boiling fraction was suitable for conversion over catalyst 7360.

First a liquid phase gasoline of boiling range 46-165°C derived from pulverized coal, pitch, coal tar residue and generator tar was processed over catalyst 7360. This product gave trouble-free operation. The resulting gasoline boiling up to 165°C contained 60-63% aromatics. The iodine number was a maximum of 5.0, whereby it should be considered that the test was run without the refining furnace. The resultant gasification corresponded to 25% of the charge.

As it was thought from past experience that the sulfur contained in the liquid phase gasoline could damage the life of the 7360 catalyst in the regeneration, a product having a boiling range 67-160°C which had been prehydrogenated over catalyst 5058 was selected for the subsequent tests. This was a gasoline derived from a charge stock to the 5058 catalyst derived from 35% middle oil, 15% gasoline (both from liquid phase hydrogenation of pulverized coal, pitch, coal tar residue and generator tar) and a 50% middle oil from Rumanian crude oil. This product also gave satisfactory operation over the DHD catalyst in the presence of the lower boiling fractions. The resultant gasoline boiling up to 165°C contained 60-63% aromatics. Although the refining furnace was not used, the iodine number was less than 2.4. The gasification amounted to 22-24% of the charge. (Experimental data are given.)

Conditions:

Catalyst volume	1000 cc.
Direction of flow	top to bottom
Pressure	25 atm.
Temperature	29.2 mv (average)
Throughput	500 g/h = 0.5 kg./liter catalyst/hr.
Gas	500 liter fresh H_2/hr.

Report 605-47P

UTILIZATION OF GASES PRODUCED IN THE NEWLY CONSTRUCTED DHD PLANT (1940)

Frames 378-381

Abstract

The DHD gases are obtained from 5058 and 6434 hydrogenation gasoline. They are additional to the gases normally produced in the hydrogenation plant at Politz.

1. Lean gas

The DHD plant delivers, on the basis of 6,000 hours per year, 3900 cu.m. per hour of washed lean gas. This gas exists under a pressure of 50 atm. and contains around 60% H_2. According to I.G., the unsaturate content is under 3%. As it is obtained from hydrogenation gasoline, it is low in inorganic and organic sulfur compounds. This makes the lean gas best suited for production of pure H_2 by splitting. From it 8840 cu.m. per hour of pure H_2 can be produced. The 11 splitting furnaces of the Politz plant have a combined splitting capacity of 99,000 cu.m. per hour of pure H_2. Thus, the above lean gas will supply only 7% of the total capacity. This can be handled readily and will substitute for other gas which,according to older plans,would be taken from the propane recovery plant.

2. Rich gas

The DHD plant produces 1400-1700 cu.m. per hour on the average during 6000 hours per year. The rich gas comes out at a pressure about atmospheric and contains 4 to 5% H_2. The pentane content is about 6 to 8% corresponding to 3,500,000 liters pentane per year. The propane-butane content is 37 to 39% by volume. This corresponds on the average to 385 cu.m. propane/hr. If 65 vol. per cent of this is recovered by refrigeration, there is recovered 250 cu.m./hr., or 5,100,000 liters/6000 hr. year. The butane content is 206 cu.m./hr. The total butane, 5,350,000 liter/yr., can be recovered by refrigeration. It appears expedient to process the rich gas for liquefied gases. The residue gas after liquefaction will fall to about 1000 to 1200 cu.m./hr. of about the following composition:

8% H_2
23% CH_4
48% ethane
21% propane

3. By-product gas

100 cu.m. gas/hr is produced in the stabilization of wash gasoline. This contains 20% propane and butane. As it exists at a low pressure it can be charged to the rich gas.

The DHD plant can furnish lean gas for the production of about 9000 cu.m./hr. of H_2, rich gas for liquefaction, and residue gas of sufficient heat content to supply its own needs and that of the splitting plant.

STUDIES OF DHD - RESIDUAL GASOLINE (AFTER RAFFINATE SEPARATION OF AROMATICS) FROM CHAMBER NO. 1

Frames 382-384

Abstract

Comparative data are given for operating periods 30-35 and 38-39.

In the operating periods 30-35 the initial boiling point of the charge product dropped from 130° to about 105°C, with lower naphthene content of 24-27%. The resulting gasoline from the separator boiling up to 165°C contained 59-62 vol. % aromatics and had octane numbers of 80-83. In the raffinate (residual) gasoline with about 12-17% naphthenes, the octane number increased thereby from about 49 to about 61.

For the following raffinate gasoline tests, the resulting gasoline cut in the stripper up to 165°C was blended with a debutanized light gasoline (composition shown) to give a blend containing about 50% aromatics. A considerable improvement of the raffinate gasoline for periods 38 and 39 over that of periods 34 and 35 was obtained in that the naphthene content of the charge was increased from 28 to about 40% and the boiling curve of the lower fractions was lowered about 15°C. The antiknock value of the raffinate gasoline rose from about 65 to 70.5.

OPERATION RESULTS ON THE APPLICABILITY OF VARIOUS RAW GASOLINES FOR THE PREPARATION OF DHD GASOLINE (POLITZ, MARCH, 1942)

Frames 385-416

Abstract

It is assumed that all resulting gas recovery gasoline will be blended into the finished product. From the products of the liquid and the gas phase coal hydrogenation reactors there is obtained a gas-recovery gasoline having about 10-35 vol. % C_4, 30-70 vol. % pentane and a residual content of higher hydrocarbon. The higher hydrocarbon content depends upon the outside temperature, the separator temperature, and the cooling action of the gasoline condenser in the distillation. The volume of recovery gasoline of 35% C_4 is assumed to be 270 cu.m./day (175 cu.m. butane/hr.). By making reasonable assumptions, several cases are calculated to show how all the recovery gasoline can be utilized by blending into DHD gasoline, storage tank gasoline, etc. which will still meet octane number, aromatic content, and vapor pressure specifications. Most of the recovery gasoline has to be used in storage tank gasoline. Only a limited amount can be used in DHD gasoline. If 505 gasoline is the charge to DHD, much more recovery can be used than if the charge to DHD is predominantly 6434.

DETAILED STUDIES OF DHD GASOLINES P 1023 AND P 1031

Frames 417-428

Abstract

The DHD gasoline was first separated into the following fractions:

1. Fraction up to 75°C (aromatic-free forerunnings)
2. Fraction 75-95°C (boiling range of benzol)
3. Fraction 95-126°C (boiling range of toluol)
4. Fraction >126°C (boiling range of xylols and higher aromatics)

The compositions and octane numbers of the gasoline and the fractions were determined. Similarly, the aromatic-free residual gasoline (after separation of aromatics) was separated and studied. From the aromatic extract, benzol, toluol, and xylol plus higher aromatics were separated and their knock behavior was studied.

Results:

The forerunnings fractions of the whole gasoline and of the residual gasoline are in good areeement and consist of 31 wt. % naphthenes and 69% paraffins and have an octane number of 79.8. The second fraction contains 47% benzol with an octane number of 119.2 and a residual gasoline of octane number 63.5, which causes the octane number of this fraction to be only 76.

The knock resistance climbs in the third fraction, although the residual gasoline has fallen to an octane number of 48.6. The increase in the octane number agrees with the higher aromatic content of the fraction, which contains 79.6% toluol of octane number 105.

The highest octane number is shown by the fourth fraction. This fraction contains 95.5% xylol plus higher aromatics with an octane number of 101.9. The 4.5% residual gasoline with an octane number of 29.5 exerts a small lowering of the octane number. P 1023, with an octane number of 83.1, has the highest octane number of the heavy duty gasolines delivered to date. This gasoline, however, shows no essential distinction from the other gasolines with regard to its gross composition (aromatics, naphthenes, paraffins, etc.). The good residual gasoline with an octane number of 70.1 and particularly the light fraction up to 100°C, which is much better than the average of the other DHD finish tanks, is the decisive factor in the higher octane number of P 1023 gasoline.

A comparative study of DHD gasoline P 1031, which had the lowest octane number (79.8 as compared to 83.1 for P 1023), was made.

Comparing the starting gasolines, P 1023 contains 1.5% more aromatics than P 1031 (56.5 compared to 55 wt. %). Furthermore, P 1023 contains more material in the boiling range up to 70°C (21 vol. % vs. 18.5 vol. %) and more boiling up to 100°C (52.5 vol. % compared with 51 vol. %). The octane number of P 1023 exceeds that of P 1031 by 3.3 units. The better quality of P 1023 is attributed much less to the higher aromatics content than to its better residual gasoline, which has an octane number of 70.1 compared with 65.5 for P 1031. The larger low boiling fraction with its apparently higher fraction of isoparaffins is the decisive factor for the better quality of the residual gasoline and the whole gasoline. Complete comparative data are given for the fractions.

FILE NOTE ON DHD GASOLINE

Frames 429-432

Abstract

(Hydrierwerke - Politz, May, 1943)

Summary

Observations indicate that the falling of the overload curve in the rich region can be explained through the presence of intolerable quantities of normal heptane, contained as a result of too high an end point of the fore-running light oils in DHD gasoline.

Evaluation

(1) The light oils to be added to the DHD separator gasoline ought to have an octane number not under 76. The distillation end point should not be above 95°C.

(2) Through tests it will be determined how the normal heptane and other paraffin hydrocarbons influence the overload curve.

Detailed Statement

Four times in the last nine months, a falling of the overload curve in the rich range below the specified value has been observed. These occurred for tanks 1061-1065 in August, 1942, 1084-1086 in October, 1942, and for 1113-1115 in January, 1943, and for 1143-1145 in April, 1943. In the lean range the overload curve was always sufficient with a single exception, and that was for tanks 1114-1116 and 1117 in January, 1943, in which the value lay very slightly under the specification.

In October, January, and April the sinking of the overload curve in the rich range clearly coincided with high end point of the light oils, which values were all above 95°C, and some were as high as 100°C. Also in August the end point of the light oils lay between 96 and 98°C for several days.

The overload curve in the rich range lay very high above the specification value four weeks in the period December 1-30, 1942, even up to 1.5 p.m.e., and again in the period February 10 to March 10, 1943, when the value was around 1.0 p.m.e. At both times the end point of the light oils was remarkably constant at 90°C.

As normal heptane boils at 98°C and as normal paraffin hydrocarbons exert the strongest lowering of the overload curve, it is thought the sinking of the overload curve in the rich range is due to the presence of unduly high quantities of normal heptane. This must be the case particularly in working up highly paraffinic gasoline, for example, such as comes from petroleum, since in this the fraction of paraffin hydrocarbons is higher than in the hydrogenation gasoline.

Petroleum gasoline therefore requires a particularly good fractionation in the first distillation. Should there result from any temporary disturbance or trouble a forerunning fraction with too high an end point this should be returned to the tank preceding the first distillation. It is necessary to make the separation, so that normal heptane remains in the charge stock and is partly cracked and partly aromatized in the chamber.

REPORT ON THE INFLUENCE OF VARIOUS METHODS OF OPERATION OF DHD CHAMBERS ON THE CHARACTERISTICS OF THE END GASOLINE

Frames 433-440

Abstract

The striking phenomenon that the overload curve of many of our DHD finished products in the rich mixture range suddently drops below the position of the standard C-3 curve is always observed when the chambers 21 and 22 are running a separator gasoline of relatively low aromatic content, while separator gasolines with higher aromatic contents after their blending to the finished product do not show this undercutting of the standard curve. It was believed that this was due to the manner of operating the DHD chambers.

To test this, two separator gasolines were used; one taken from chamber 21 on August 6, 1943, while running with lower gasification and having an aromatic content of 56.5 vol. %, and the other from chamber 22 on August 12, 1943, while running with considerably higher gasification and having an aromatic content of 60.5 vol. %. It was assumed that the charge stock was the same, or nearly so. Actually they were very different, that to chamber 21 containing 10.5 aromatics plus unsaturates and 29% naphthenes, while that to chamber 22 contained 9.5% aromatics plus unsaturates and 48% naphthenes. The conclusions were:

(1) The manner of operation of the DHD chambers to give higher or lower aromatic content has decisive influence on the course of the overload curve, especially in the rich mixture range.

(2) The extract from a gasoline run to a higher aromatic content, as well as the gasoline itself, has a higher blending value. By working up a gasoline run to a lower aromatic content to a finished product it is always necessary to consider the weakness, namely the undercutting point of the standard curve and whether one can obtain a satisfactory curve by blending in very good light fractions. These tests and previous experience have shown this deficiency of a separator gasoline of the character of that of chamber 21 can be controlled only by increasing the aromatic content in the finished product; it was necessary in this case to increase the aromatic content to 53 vol. %.

(3) From the octane number and analysis of the residual gasoline and an investigation of the overload curve, it was concluded that this fraction has so little influence on the overall product that it is of interest only in special cases.

(4) The composition of the extract is of great significance, as the problem appears to be mainly one of the constitution of these hydrocarbons.

STUDY OF THE INFLUENCE OF TOLUENE WITHDRAWAL UPON THE OVERLOAD SENSITIVITY OF DHD GASOLINE

Frames 441-444

Abstract

By removal of the toluene fraction from DHD gasoline and substituting therefor higher aromatics by raising the end boiling point, it was found that with constant aromatic content there was never a lowering of the overload capacity; in fact, a slight improvement resulted.

DEHYDROGENATING RUMANIAN GASOLINE WITH 60% PARAFFINS

Frames 445-448

Abstract

During the period July 10 to August 10, 1943, Hydrierwerke-Politz dehydrogenated a mixture composted of 50% Rumanian crude oil gasoline and 50% of heavy gasoline (120-180°C) from hydrogenation. The run period lasted about 100 hours. The chamber gasification amounted to about 27 weight %. This value is about normal according to the yield curve for the product. This is true also for the usual DHD gasoline. It was also satisfactory in overload properties (see Tank 1175 studied by Ludwigshafen). This indicates that the conversion of the above product can be carried out and satisfactorily completed with the available catalyst (7360 granules).

Ludwigshafen obtained a run period of only 26 hr. in small tests with the conversion of 100% Rumanian gasoline to a separator gasoline with 60% aromatics. The charge stock had about the same composition as that processed here. Is this due to the catalyst used at Ludwigshafen? It is known that pilled catalyst has the characteristic of giving shorter operating periods, than granular catalyst . Ludwigshafen concluded that Rumanian gasoline should be used only in additions of 20% or this gasoline run to B-4 quality. The latter is impractical for HWP since then the light fractions to be added are lacking.

For us it is necessary to use a DHD catalyst which will permit conversion of 50% or more to gasoline of superior quality for sufficiently long running periods (100 hr. or more). If the pilled catalyst will not attain this, the granular catalyst must be substituted for it.

DHD CONVERSION OF CRACKED GASOLINE LUDWIGSHAFEN MAY 8, 1944

Frames 449-451

Abstract

Report on the tests of a sample of cracked gasoline submitted by HWP, July, 1943.

As a result of its high content of olefins and paraffins, 12 and 55% respectively, this gasoline is not particularly suitable for conversion by the DHD process. The high sulfur content (about 0.2% by weight) is especially troublesome. As you will see from the enclosed report, a sulfur content of 0.1% reduced the cycle length to one-half and even 0.03% sulfur brings it down to 70%. Sulfur, even in very small quantities, causes an irreversible damage to the catalyst activity. By the addition of only 15% of this cracked gasoline to hydrogenation gasoline, a 30% decrease in cycle length is brought about due to its sulfur content. The irreversible deterioration of the catalyst is still not taken into account, however.

On these grounds, we recommend that when conversion of such cracked gasoline becomes necessary, that it be given a prehydrogenation treatment sufficient to eliminate the sulfur, but to lower the aromatic content as little as possible. In Politz, the only method of desulfurization to be considered is a pretreatment in the prehydrogenation step.

Even when the cracked gasoline is pretreated in a satisfactory manner, we still conclude, with respect to its low aromatics and naphthene content and the poor octane number of the residual gasoline, that high additions of this gasoline to hydrogenation gasoline for the production of DHD gasoline of C-3 quality is undesirable. We conclude that the addition of this cracked gasoline should be held below 10% if at all possible.

REDUCTION OF DEHYDROGENATION ACTIVITY AND DURATION OF THE DAMAGE TO DHD CATALYST BY OXYGEN, SULFUR, NITROGEN AND UNSATURATES

Frames 452-453

Abstract

Hungarian heavy gasoline was passed over DHD catalysts of types 7360 and 7935 in the 100 cc. furnace and dehydrogenated under the ordinary DHD conditions (20 atm. H_2 pressure, 0.5 kg./liter and an hourly throughput of 0.8 cu. meter gas per kg. oil) in life tests. The aniline point of the product formed (about -5) was maintained approximately constant and the homogeneous reaction temperature of about 26 mv. (depending upon the activity of the catalyst used) rose to 28.5 mv. After reaching this temperature it was shut off. Through addition of phenol, aniline, CS_2 and diisobutylene to the charge stock, the limiting effect of oxygen, nitrogen and sulfur compounds and unsaturates was determined. Addition of phenol or aniline to Hungarian heavy gasoline exerted a reversible reaction limitation; addition of CS_2 caused irreversible catalyst damage which was not completely removed after many regenerations. An addition of 10 wt. % of the diisobutylene had no inhibiting effect upon the reaction.

A decrease in the cycle length by 50 per cent was produced by addition of about 0.2 to 0.4 wt. % oxygen, 0.05 wt. % nitrogen, or 0.1 wt. % sulfur to Hungarian heavy gasoline. Oxygen and nitrogen addition result in coke laydown which necessitates raising the temperature to maintain a given conversion.

TESTS ON FIRE BRICK

Frames 454-457

Abstract

Tests with unused and used fire brick from the DHD plant Politz showed no difference in catalytic activity toward the DHD reaction. The difference in color (yellow or very dark) was thought to be due to the different state of oxidation of the iron.

PRICE FORMATION OF THE DHD AVIATION GASOLINE SEPTEMBER 17, 1940

Frames 458-464

Abstract

Objective: To determine the profitableness of the DHD process.

Summary:

The process permits the production of 90 octane number aviation gasoline which contains no tetraethyl lead. If a 5058/6434 hydrogenation gasoline is assumed as starting material, the production cost is RM 325 per ton. If an imported petroleum gasoline is processed, the production cost is estimated to be RM 125 per ton.

In comparison, the cost of a 90 octane number aviation gasoline made from bituminous coal tar and crude oil by hydrogenation and addition of tetraethyl lead amounts to about RM 320 per ton.

Through the DHD process it is possible, with the aid of tetraethyl lead, to produce aviation gasoline having a considerably higher octane number than 90. The production cost of a DHD aviation gasoline with an octane number of 96-98 is estimated at RM 350 per ton from hydrogenation gasoline, and RM 150 per ton from petroleum gasoline.

90 octane gasoline from coal & crude oil

Raw material and process cost	RM 293.
Lead tetraethyl	19.80
Contingencies	7.20
	RM 320.

90 octane DHD gasoline from 5658/6434 hydrogenation gasoline

Raw material	RM 298.
Interest	10.
Salaries and wages	2.57
Power	10.
Contingencies	4.43
	RM 325.

DHD gasoline from imported (Gulf Coast) crude oil gasoline

Raw material	RM 94
Process cost	23
Contingencies	8
	RM 125

It is assumed that the cost of tetraethyl lead to bring either of the last two to 100 octane number gasoline will cost RM 25 per ton.

DHD LARGE SCALE RESEARCH AT LUDWIGSHAFEN
REPORT OF A VISIT TO LUDWIGSHAFEN TO OBSERVE DHD RESEARCH

Frames 465-467

Abstract

1. Chamber 504

This DHD test chamber contains 3 dehydrogenation furnaces with 930 liters of catalyst and a refining furnace with 315 liters of catalyst. The chamber processes 450 kg. of charging stock per hour. Each furnace is provided with a mantle for air cooling. An electric preheater is used.

2. Manner of Operation

The chamber has been in operation three months. It operates without difficulty. Run periods of 24 to 30 hr. and regeneration periods of 12 hr. were attained.

Temperatures in the dehydrogenation furnace are 27 mv. and in the refining furnace 14.5 to 16.5 mv. Aniline point of the catch pot product is determined hourly and is held at 0°C. The density of the recycle gas, which lies between 0.4 and 0.55 and rises with stronger dehydrogenation, is an indication of the operation.

3. Pressure

The pressure of the catch pot is held at 55 atm. which gives a pressure of 75 atm. after the recycle pump. This pressure is occasioned by the 25 atm. hydrogen partial pressure which is necessary in the refining furnace.

4. Catalyst

A molybdena-alumina catalyst is used. The density of the pilled catalyst is 0.815. It loses its activity in use due to deposition of polymers. Carbon and coke are essentially absent. It is regenerated by burning with air.

5. Yield of DHD gasoline and aromatic content

Currently the chamber is producing a catch pot product with 65 vol. % aromatics. When this is mixed with the fraction boiling below 90°C. one obtains a full range aviation gasoline with 50% aromatics. By this operation the yield of DHD gasoline is about 80 wt. % of the starting gasoline.

6. Isobutane in the end gas

As a by-product, the DHD gas consists of:

3.3 wt. % H_2
20.7 wt. % CH_4
24.7 wt. % C_2 (0.3% olefin)
26.0 wt. % C_3 (1.0% olefin)
17.6 wt. % nC_4 (2.5% olefin)
7.7 wt. % iC_4 (2.5% olefin)

7. Boiling point of DHD gasoline

50 vol % boils under 100°C and 98 vol. % below 165°C.

8. Octane number

Octane number of the product with 50% aromatics is 81 by the motor method and addition of 0.12 vol. % TEL brings it up to 91-94.

9. Overload capacity

The overload capacity of DHD gasoline is as satisfactory as that of 7019 gasoline or a 100 octane gasoline on the isooctane basis, although all three gasolines contain the same amount of TEL (.12%).

10. Iodine Number

The iodine number of DHD gasoline from a Scholven 5058-6434 coal liquefaction gasoline lies below the value of 4 set as a maximum for aviation gasoline. A Brabag gasoline, on the other hand, gives an iodine number of 7-8, which is generally dropped below 4 in the refining furnace.

11. Oxygen Stability

Bomb tests with 7 atm. O_2 for 4 hr. at 100°C show no pressure drop.

VISIT TO LUDWIGSHAFEN JANUARY 28, FEBRUARY 9, 1941 TO OBSERVE OPERATION OF DHD CHAMBER 504

Frames 468-471

Abstract

This chamber completed 82 periods. In the course of 17 operating days in January, it dehydrogenated 66% of the time, was regenerating 24% of the time, and was being repaired 10% of the time. The successive furnaces were being operated with stepwise increasing temperatures:

Furnace I at 25.5 mv. (average)
Furnace II at 27 mv. (average)
Furnace III at 28 mv. (average)

With 10% aromatics in the charge the aromatic content increased as follows:

At end of I 40% aromatics
At end of II 55% aromatics
At end of III 67% aromatics

Charge rate 0.32 to 0.4 kg./liter catalyst/hr.
Cycle gas rate 400-700 cu.m./hr.

A gasoline having a high naphthene content (45%) gives best results in the DHD process. Unsaturated compounds should be absent. Good results were obtained.

A test of 5058-6434 gasoline fraction from Politz, boiling 130-190°C and having a naphthene content of 55% gave a catch pot product having 65-75% aromatics. Gasification amounted to 15%.

DEHYDROGENATION - CHAMBER OR TUBULAR REACTORS

Frames 472-479

Abstract

An indistinct photostat of a paper concerning heat of reaction for the DHD process.

Report 605-47P

CONFERENCE REPORT ON DHD PROCESS, SEPTEMBER 9, 1941

Frames 481-488

Abstract

Details of materials of construction, items of equipment, and plant layout for Politz are included.

LUDWIGSHAFEN - I. G. REPORT ON PRODUCTION OF DHD GASOLINE FROM 6434 PREHYDROGENATION GASOLINE FROM BRABAG IN CHAMBER 504

Frames 489-516

Abstract

Summary:

In a semi-commercial apparatus (Chamber 504), 6434 brown coal tar prehydrogenation gasoline from Brabag's Magdeburg plant was dehydrogenated and the product obtained was tested in suitable blends in the overload motor. The charge was a gasoline fraction of 90-169°C, from which a DHD gasoline boiling up to 165°C with about 66 wt. % aromatics was obtained. By blending this with stabilized light gasoline from Brabag, or a light gasoline obtained by redistillation and stabilization of a coal liquefaction light gasoline, a finished gasoline of 50 vol. % aromatics was obtained. The dehydrogenation proceeded with a throughput of 0.48 kg./liter catalyst/hr. with increasing mean temperatures of about 510° and 516° for furnaces I and III. The yield of C_4-free catch pot product was 79-80%, with the 93% boiling point below 165°C.

The run period of satisfactory operation was 24-29 hr., and the regeneration time (from stopping of charge to starting) was 10-12 hr. Flow was from top to bottom in the furnaces. A charge of 935 liters of catalyst 7360 was used in furnaces No. 1-3 and 315 liters of the same catalyst was used in furnace No. 4. The charge rate was 0.48 kg./liter/hr. and the gas circulation rate was 1.2 cu.m./kg./hr.

Regeneration was by nitrogen containing 4% air at 50-60 atm. pressure and the temperature was held below 560°C. The heat of reaction in the dehydrogenation furnaces was -55 Cal. and was distributed as follows: 65% in furnace No. 1, 24% in furnace No. 2 and 12% in furnace No. 3. The heat of reaction in the refining furnace was + 20 Cal. Complete test data for the charge and products are given.

Report 605-47P

PRODUCTION OF DHD GASOLINE FROM 5058-6434 SCHOLVEN GASOLINE (LUDWIGSHAFEN, MARCH 31, 1941)

Frames 517-558

Abstract

A fraction boiling 90-180°C from Scholven 5058-6434 coal liquefaction gasoline was dehydrogenated in the semi-commercial DHD chamber 504, and the product was tested in overload test engines. A primary DHD product containing 65 wt. % aromatics was obtained, which after blending with the previously separated fraction boiling up to 90°C (20% of the original Scholven gasoline) and final redistillation to an end point of 165°C and stabilization gave a high value fuel with 55 wt. % aromatics. It had better overload performance than CV2B aromatization gasoline made from the same raw material. It had a motor method octane number of 79.5 clear and 94.5 with 0.12% TEL. Conditions were:

Throughput	0.48 kg./liter catalyst/hr.
Catalyst	7360
Furnace temperature	
No. 1	495°C
No. 3	525°C
Product/gas ratio	0.9 to 1.2
Run period	35 hr.
Regeneration	12 hr. (including drying, depressurizing and flushing)

The refining furnace accomplishes a rehydrogenation of unsaturates to lower the iodine number below the maximum of 4 for aviation fuel.

Hydrogen is formed in the process and must be removed to maintain a hydrogen partial pressure not greater than 20 atm.

Yield of product containing 92-95% boiling below 165°C was 79-81%. Dimensions of the apparatus are given.

Iodine number of the product increases during the last part of the make period and the conversion is stopped when the iodine number reaches 3.5. Hydrogen splitting is about 0.74-0.80%. Coke laydown is about 0.15-0.23%. Test data for the products, and values for the heat of reaction are given. A flow diagram of the plant is shown.

HYDROFORMING - LETTER DATED FEBRUARY 18, 1942, FROM LEUNA TO POLITZ

Frames 561-566

Translation

Hydroforming agreement

We send to you in the attached papers Section III, Quality, of our Hydroforming contract. The text is, as we have already told you, the result of many conferences with Herr Mucklich and his associates in RLM (Reichs Air Ministry). The very detailed presentation of this section results from the fact that RLM in the hydroformed gasoline establishes ever higher demands (up to 65 vol. % aromatics) while at the same time deteriorating the quality of the straight run gasoline used as starting material for hydroformed gasoline. In the paragraph on quality of hydroformed gasoline (HF) presented in table form, are the qualities which we believe it is possible to obtain, over against the RLM demands for quality, yield and production volume.

Reichsminister of Air Transportation. Firm of Ammoniakwerk Merseberg, Leuna. III Quality

Since the quality of HF gasoline is dependent upon the composition of the straight-run gasoline used as charge stock, an unequivocal establishment of the quality of the HF gasoline is not possible.

Within the limits of fluctuation of the gasoline which is practically available, according to the present technical state, the quality of the starting material and the quality of the HF gasoline, the throughput of straight-run gasoline or the production of HF gasoline stated below is to be expected:

Characteristics of Straight-run Gasoline

	I	II
Specific Gravity (20°C)	0.75	0.74
% boiling up to 100°C	20	20
% boiling up to 155-160°C	95	95
End point, °C	165	165
Aniline Point, °C	48	51
Octane No. (MM)	Not under 58	50-55
Sulfur content	Under 0.1 Wt. %	Under 0.1%

Characteristics of HF Gasoline

	1(a)		1(b)		2	
Specific gravity	0.785-0.800		0.770-0.785		0.770-0.785	
Initial, °C	Ca 45		Ca 45		Ca 45	
End point, °C	165		165		165	
Reid vapor pressure @ 37.8°C	Under 0.4 atm.		Under 0.4 atm.		Under 0.4 atm.	
Aromatic content, vol. %	About 63		50		55	
Gum, mg./100 cc.	Under 5		Under 5		Under 5	
Lead bomb test, mg./100 cc.	Under 5		Under 5		Under 5	
Sulfur content, %	Under 0.05		Under 0.05		Under 0.05	
Overload capacity	0.8	1.1	0.8	1.1	0.8	1.1
Mean effective pressure, $kg./cm.^2$	>23	13.5-14.5	17.0-18.0	10.5-11.5	18.5-19.5	12.0-13.0
Yield	71		81		72.3	
HF production	33,000		59,000		37,500	

From the best mixed-base, straight-run gasoline available (Gasoline I), the best quality HF gasoline 1(a) was produced in the Moosbierbaum plant at 71% yield, while with the same starting material as a lower limiting case, HF gasoline 1(b) was produced at 81% yield. The paraffinic starting material, Gasoline II, gave in the best case at 72% yield the HF gasoline of quality 2.

The throughput of straight run gasoline and production of HF gasoline named in Sect. I (delivery quantity) appear to be as follows, for previously planned expansion steps, according to the quality of the raw material and the HF gasoline demanded:

1st Expansion Step

Aromatic Content, %	50	55	60	50	55
Aniline Point of Charge Stock, °C	48	48	48	51	51
Charge, tons/year	87,500	77,000	63,000	80,500	70,000
HF Gasoline, tons/year	66,500	55,500	42,000	56,500	47,000

2nd Expansion Step

Aromatic Content, %	50	55	63	50	55
Aniline Point of Charge Stock, °C	48	48	48	51	51
Charge, tons/year	73,000	62,000	47,000	63,500	52,000
HF Gasoline, tons/year	59,000	48,000	33,000	48,000	37,500
Total Production, tons/year	125,500	103,500	75,000	104,500	84,500

Previously, the aromatic content of the finished HF gasoline was set at 50-55 vol. % for Moosbierbaum.

(Politz, February 16, 1942)

Memorandum on the Hydroforming Plant at Moosbierbaum

In a conference in Politz on February 12 and 13, 1942, participated in by Drs. Wels and Kaufmann of I.G. Farbenindustrie, Leuna, to inform us of their experiences with dehydrogenation, the following concerning the hydroforming process was communicated:

In September, 1939, Dr. Kaufmann visited New York and learned from Standard Oil Co. of New Jersey and Kellogg Co. of the development of the American process for the aromatization of gasoline hydrocarbons. At that time the process was in the laboratory development stage, in which furnaces with 4 liter throughput were running. The process was further developed by I.G. The first large plant in Moosbierbaum having an annual production of 60,000 tons DHD gasoline will begin operation in March, 1942, and is designed for Rumanian straight run gasoline. A second expansion step of the plant is planned and should be completed in 1943.

The HF process comprises catalytically dehydrogenating a gasoline fraction boiling at 90-165°C at 15 atm. pressure and temperatures up to 540°C. The large scale apparatus consists of 2 furnace groups of 3 furnaces each, of which each contains 7 cubic meters of catalyst.

Each furnace group can be connected to a dehydrogenation and regeneration circuit. The dehydrogenation circuit is equipped with regenerators and preheaters. A separate heating up of each furnace is possible by means of an intermediate preheater. The preheater consists of vertical tubes which are gas heated. The combustion gases are not recirculated. In the regeneration circuit there are regenerators but no preheaters. On the other hand, behind each furnace there is a steam heater for the purpose of utilizing the heat of regeneration.

In the furnaces there are 3 catalysts of different activity. Of them, the least active operates at 510° C, the intermediate at 490°C and the most active at 470°C. Furnace No. I contains 30% of poor, 30% of medium, and 40% of good catalyst. Furnace No. II is filled with 50% medium and 50% good, and Furnace No. III with 100% good catalyst. This arrangement makes it possible to equalize the heat evolution in the several furnaces. The heat of reaction is calculated to be 150-200 calories.

Gasoline from Rumanian crude oil having an aniline point of 51°C is used as the charge. It contains 25% naphthenes. The gasoline, which boils up to 165°C, is separated into a forerunnings fraction and a residue. The end point of the forerunning is about 75°C. The residue which is charged to the DHD chambers boils from 90° to 165°C.

A charge of 10.5 tons per hour is used and 11,000 cu.m. of recycle gas are circulated. The entrance temperatures in the several furnaces are about equal at 500°C at the beginning of the dehydrogenation period. The temperature is raised at the rate of about 3°C per hour. After 10-15 hours, the maximum temperature of 540°C is reached, and the furnace is shifted to regeneration. The gas density climbs from 0.35 to 0.50. The catch pot liquid contains 60-62 vol. % aromatics.

While one furnace group dehydrogenates, the second furnace group is on regeneration. Each furnace is treated with air simultaneously. The ignition temperature is about 400°C. Each furnace requires 4 to 5 hours for burning out. The total coke amounts to 0.5-1 wt. % of the charge. The dehydrogenation and regeneration times are about equal (1:1). The throughput during dehydrogenation is about 0.5-0.6 kg. charge/liter of catalyst.

The resulting liquid condensate is stabilized and refined with 0.2 to 0.4 wt. % of 98% H_2SO_4.

The gasoline loss amounts to about 1% by wt. After neutralization the product is redistilled, whereby about 4% residue remains. The gasoline, which contains about 60 vol. % aromatics is combined with the forerunning to make a finished gasoline of about 50 vol. % aromatics. The octane number of the finished gasoline is 65.

The switching of the converter groups from dehydrogenation to regeneration occurs automatically by means of hot valves. These are still not proved commercially. As each converter group operates on dehydrogenation only half of the time, the overall throughput is 0.25-0.3 kg. per liter per hour. At the pressure of only 15 atm. the relatively high coke formation and the short operating time of 10-15 hours results. As a result of the low hydrogen partial pressure, the H_2SO_4 refining cannot be replaced by a refining hydrogenation.

For the second expansion step of the Moosbierbaum plant, the following improvements are anticipated. In place of the previous converters, the intention is to use the heated converters used in the T-52 process. One such converter consists of a standing tube bundle (200 tubes) which is swept with hot combustion gases.

These converters operate isothermally. The hot valves therefore will be omitted since the heated tube converter will be connected to only one circuit. The regeneration will be conducted as in our DHD plant. The cycle gas will be expanded and the apparatus flushed and filled with inert gas before it is regenerated. After the regeneration, the gas will be removed and replaced with dehydrogenation gas. Furthermore the pressure will be raised to 30 atm. in the new Moosbierbaum plant.

INHIBITORS

Frames 568-569

Abstract

Transmittal letter, October 14, 1941, from I.G. at Ludwigshafen to Dr. Montfort of Hydrierwerke-Politz, covering a 500 cc. sample of Inhibitor A for testing. It is produced by the high pressure experiments at Ludwigshafen, boils in the gasoline range, is completely soluble in gasoline and leaves no residue upon evaporation. Suggested concentration is 0.01 to 0.02%; suggested experiments are bomb and storage tests with leaded gasoline, with particular attention to inhibition of lead precipitation.

DATA ON STABISOL AND INHIBITOR A

Frame 570

Abstract

Laboratory report of December 4, 1941, giving test data on use of Stabisol and Inhibitor A.

STUDY OF STABISOL AND INHIBITOR A IN AROMATIC GASOLINE

Frames 571-572

Abstract

Letter of January 29, 1942, from Dr. Monfort of Hydrierwerke-Politz to M. Pier at Ludwigshafen, presenting laboratory data to show that it took about 0.03% of Inhibitor A to be as effective as 0.01% of Stabisol, in aromatic gasoline.

ADDITIVE A

Frame 573

Abstract

Transmittal letter of February 23, 1942, from I.G. at Ludwigshafen to Hydrierwerke-Politz covering nine containers of Additive PA, together with one container of Additive A and one of Additive S (residual gasoline).

TEST OF ADDITIVE A

Frames 574-575

Translation

Summary by Lajus dated February 16, 1942, of a meeting in Politz on February 4 and 5, 1942, between Dr's. Wissel, Montfort, Steffen, and Huthwelker. It is necessary that the DHD gasoline produced at Politz and Ludwigshafen be completely stable during prolonged storage, but the air ministry (RLM) does not officially permit inhibitors to be used. We propose to use a small amount of an inhibitor-like substance which was originally in the gasoline but is destroyed or removed by the DHD process. Such a material (isobutyl oil) had been sent as "Stuff A" to Dr. Montfort for testing its inhibitor action. Meanwhile, we have established that in many instances fractions of residual (sumpf) gasoline have better inhibiting actions than Stuff A. Until the question of obtaining specific additives has been completely clarified in Politz, we propose that Politz add to its DHD gasoline a mixture of a Stuff A and sumpf gasoline boiling in the range of 160-230°C (in the ratio 3:2), as Ludwigshafen has been doing for a long time. A sample of this mixture, "PA", has been given to Dr. Montfort for immediate testing on Politz DHD gasoline.

The results obtained in Ludwigshafen with Politz DHD gasoline P 1018 are as follows:

Bomb test with P 1018 - 0.12 vol. % tetraethyl lead.

Additive	Pressure Drop	Glass Dish	Gum
None	0.6 atm.	43.1	66.6
0.02% Stuff A	None	26.5	22.3
0.05% Stuff A	None	7.2	4.7
0.02% Sumpf gasoline 160-230°	None	5.1	4.4
0.05% Sumpf gasoline 160-230°	None	5.5	4.6
0.02% Stuff PA	None	4.8	0.4
0.05% Stuff PA	None	3.1	3.6

Dr. Montfort stated that Politz had obtained similar results with Stuff A on P 1018.

Dr. Wissel agrees in principle. Further experiments should be undertaken in Politz and Ludwigshafen as to the effect of PA. The Politz DHD plant adds 0.01 to 0.05% Stuff PA, and with a production of 6000 tons/month of DHD gasoline Politz will need about 1.2 tons/month of Stuff PA.

AGING TESTS

Frame 576

Abstract

Test data, March 21, 1942, of Hydrierwerke-Politz, on 16 samples comparing storage and oxygen bomb tests.

INHIBITOR TESTS

Frames 577-582

Abstract

A file of correspondence dating from December 12, 1942, to February 3, 1943, pertaining to a requirement of the air ministry (RLM, or Reichsluftfahrtministerium) on December 9, 1942, that Stuff BJ - 5 (BI 5), made by Ruhroel, be used as an inhibitor for premium gasoline, and any objections thereto be made before December 31, 1942. This requirement was based upon the relatively poor storage stability of the base stock, and upon a desire to use a universal inhibitor. This apparently came as a complete surprise to Hydrierwerke-Politz and to I.G. at Ludwigshafen, and neither had had a chance to compare it to their PA. It was believed to be technical phenols obtained by extraction, and therefore similar to I.G.'s inhibitor.

INHIBITOR TESTS

Frames 583-593

Abstract

A file of correspondence dating from March 4, 1943, to May 25, 1944, on Inhibitor (Hemmstuff) V3. Initial comparative tests of the RLM's Stuff BI 5 showed it was no better than other materials and that V3 was far superior. In a meeting on June 9, 1943, at Hydrierwerke-Politz, it was stated that the RLM had retreated from its former position not to accept other inhibitors. The question of a patent application was discussed, and it was stated that the use, as inhibitors, of phenols from the Phenolsolvan process was new. The inhibitor put out by Ruhroel had been found to be phenols from the alkali wash of their process. They had a boiling range of 180 to 225°C and contained about 50% carbolic acid and 40% cresols. The price was 1675 R Marks/ton at Politz. According to experiments at Politz, the higher boiling phenols, 180 to 230 or 235°C, from the Phenolsolvan extract, were especially useful. This material was accepted by the RLM and designated as Inhibitor (Hemmstoff) ZV 3. People at Leuna (Ammoniakwerke-Merseberg) were also interested in using this inhibitor.

In a letter of May 25, 1944, were discussed the patentable differences between ZV 3 and the material of Ruhroel's secret patent 746/42. The latter comprised phenols obtained from the oily products of hydrogenation, especially that portion boiling from 200 to 235°C. On the other hand, ZV 3 was water soluble and was obtained from the waste water, as in the Phenolsolvan process, and it was concluded that this was an entirely different raw material.

FIRE DANGER FROM PYROFORIC IRON SCALE

Frames 598-599

Abstract

Exchange of telegrams between Hydrierwerke-Politz and I.G. Ludwigshafen on fire danger from pyroforic iron scaling off in the reaction chamber.

SUPERCHARGE KNOCK TEST DATA

Frames 600-602

Abstract

Supercharge knock test curves, and other inspection data, of DHD gasoline produced at Hydrierwerke-Politz in March, 1944. Data are shown in attached photostats of frames 601 and 602.

SUPERCHARGE KNOCK TEST DATA

Frames 603-611

Abstract

Supercharge knock test curves and other inspection data, of DHD gasoline produced at Hydrierwerke-Politz in September, 1943, on samples P 1170, P 1175, P 1185, P 1190, P 1195, P 1200, and in December, 1942, on P 1074 to P 1087, inclusive.

PHYSICAL PROPERTIES OF GASOLINE VT 708 AND VT 340

Frames 612-618

Abstract

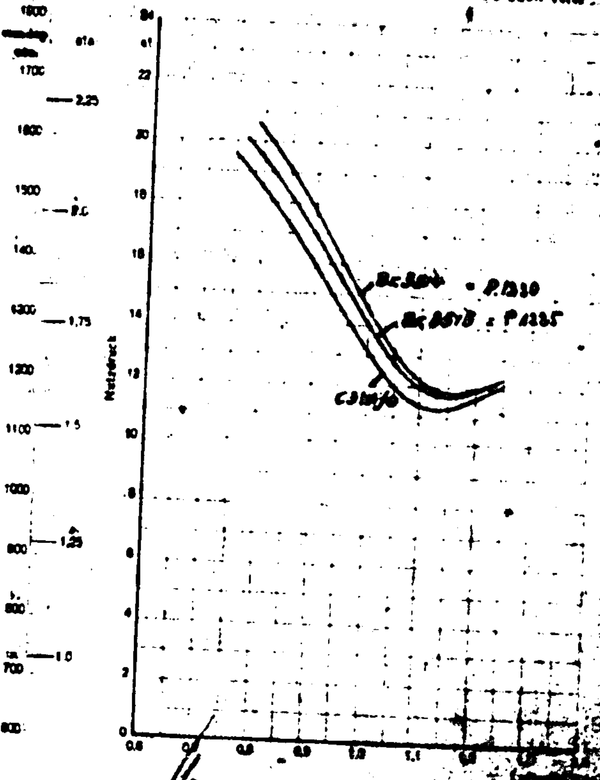
Detailed specifications for VT 708 and VT 340, which are the new designations of specific DHD gasolines made by Hydrierwerke-Politz, dated June and August, 1942.

Techn. Prüfstand Oppau

Klopfgrenzkurven nach dem Überladeverfahren

A.G. v. 14.3.44.

Motormuster: BMW 132 N
Motornummer:
Versuchstag: 28.5.44
Vergl.-Kraftstoff: C3 Alkylo, Br. 3290
Prüfkraftstoff: Br. K.M.K. Pölitz, Br. 3873 (Pölitz) Versuch Nr.: 2
Prüfkraftstoff: " " " , 3874 (Pölitz) Versuch Nr.:
Prüfkraftstoff: Versuch Nr.:
Prüfkraftstoff: Versuch Nr.:
Prüfkraftstoff: Versuch Nr.:
Verdichtungsverhältnis:
Ladelufttemperatur:
Zündzeitpunkt:
Versuch Nr.:



I.G. Farbenindustrie Aktiengesellschaft
Ludwigshafen am Rhein

Zur Kraftstoffprüfung Nr.

44 D 308-4112

		225	[illegible]
		81	0,780
		5	82
		5	95
			4[illegible],5
			30
			14
			56
		)	0,8
		13	0,[illegible]
		ativ	[illegible]gativ
		t	gut
		5	45
		5	[illegible]
		2	13
		1	2[illegible]
		1	31
		0	40
		9	48
		9	58
		8,5	68,5
		8	79
150°	86,5	87	87,5
160°	92	94,5	94
170°	95	96,5	96,5
Endpunkt °C/%	182/98	174/98	175/98

Anlage zum Brief an Hydrierwerke Pölitz A.G. v. 14.3.44.

Untersuchung der DHD-Benzinproben P 1220, P 1225 und P 1230 von Pölitz.

	P 1220	P 1225	P 1230
Spezifisches Gewicht bei 15°	0,777	0,781	0,780
Oktanzahl Motor-Methode	81,5	81,5	82
" " " + 0,12 Vol.% Pb	92,5	92,5	93
Aromaten-Gehalt Vol.%	48,5	50	43,5
Zusammensetzung in Gew.%			
Paraffine	29	30	30
Naphthene	16	14	14
Aromaten + Ungesättigte	55	56	56
Jodzahl	0,6	0,9	0,8
Schwefel in Gew.%	0,41	0,43	0,45
Doctortest	negativ	negativ	negativ
Kupferstreifentest	gut	gut	gut
ASTM-Siedekurve			
Siedebeginn °C	47	46	45
bis 60° Vol.%	5	5	[illegible],5
70°	13,5	12	13
80°	24	21	22
90°	34	31	31
100°	44,5	40	40
110°	55	49	48
120°	64	59	58
130°	73	68,5	67,5
140°	80	78	79
150°	85,5	87	87,5
160°	92	94,5	94
170°	95	96,5	95,5
Endpunkt °C/%	182/98	174/98	175/98

INSPECTION DATA OF DHD GASOLINE P 1040 TO P 1060

Frames 619-630

Abstract

Supercharge knock test curves and other inspection data, of DHD gasoline produced at Hydrierwerke-Politz, 1942, for samples P 1040 to P 1060, inclusive.

On frame 627 are detailed instructions for determining the iodine number.

RECOVERY OF DHD GASOLINE

Frames 631-634

Abstract

Letter of June 3, 1942, to Mr. Pier from Wissel and Diepenbruck of Hydrierwerke-Politz, with two diagrammatic flow sheets for production of DHD gasoline. The first employs as charge stock pure "5058 Benzin" (produced with catalyst 5058) which was produced from 30-35% natural petroleum and 60-65% coal and tar. The second employs as charge stock a mixture of "35% 6434- and 65% 5058-Benzin". The diagrams include material balances and stream inspection data.

DHD GASOLINE PRODUCTION FROM A RUMANIAN CRUDE

Frames 635-642

Abstract

Letter of November 18, 1943, to Hydrierwerke-Politz from I.G. Ludwigshafen, reporting on results of DHD treatment of a paraffinic Rumanian petroleum.

INFORMATION EXCHANGE

Frame 643

Abstract

Exchange of information between Hydrierwerke-Politz and DHD plants at Scholven and at Gelsenberg. Copies of minutes of various meetings, from the middle of 1940 through 1944.

DATA FROM POLITZ AND GELSENBERG COAL CHAMBERS

Frame 644

Abstract

Comparison of run data on a coal reactor at Hydrierwerke-Politz and Gelsenberg on October 16, 1941

INFORMATION EXCHANGE

Frames 645-647

Abstract

Minutes of a meeting of representative of the three plants on 15-17 of October, 1941. (poorly reproduced).

INFORMATION EXCHANGE

Frames 648-650

Abstract

Notes on meeting with Dr. Huther from Gelsenberg, on October 14 and 15, 1941.

PRODUCTION AT POLITZ AND GELSENBERG

Frames 651-653

Abstract

Comparison of production of Politz and Gelsenberg for January to May, 1942, at a meeting on June 19, 1942, of Dr. Jakob and Dr. Berger.

CENTRIFUGE DIFFICULTIES

Frame 654

Abstract

Notes on accident in Scholven, about June 24, 1941, due to failure of centrifuge used to purify charge stock.

ELECTRIC PREHEATERS ON HIGH PRESSURE REACTORS

Frames 655-656

Abstract

Information exchange on electric preheater on the high pressure reactors from I.G. Ludwigshafen, February 25, 1942.

INFORMATION EXCHANGE ON INJECTING COLD SLURRY TO HIGH PRESSURE HYDROGENATION CHAMBERS

Frames 657-676

Abstract

Information exchange progressive multipoint introduction of cold charge stock November 10, 1941, at Scholven. Flow sheets are given on frames 675 and 676.

PRODUCTION AT POLITZ AND GELSENBERG

Frames 677-678

Abstract

Duplicate of Frames 652 and 653.

INVESTIGATION REPORT

Frames 679-682

Abstract

Notes on a three-week visit to Gelsenberg, written October 21, 1940, discussing operations.

SCHOLVEN OPERATIONS

Frames 683-685

Abstract

Discussion, October 26, 1942, of operations at Scholven.

PLANT VISIT REPORT

Frame 686

Abstract

Discussion, November 4 and 5, 1942, of a visit to Wesseling and Gelsenberg.

PLANT VISIT REPORT

Frames 687-694

Abstract

Discussion, September 8, 1942, of a visit to Scholven and Wesseling.

PLANT VISIT REPORT

Frames 695-700

Abstract

Discussion of a visit on April 25 and 26, 1944, to the DHD plant at Scholven.

PLANT VISIT REPORT

Frames 701-711

Abstract

Discussion of visit to Scholven and Gelsenberg on December 17 and 18, 1940. Frame 711 has a drawing of a gas recycle pump operating at 700 atmospheres.

REPORT ON GELSENBERG OPERATIONS

Frames 712-723

Abstract

A discussion of the hydrogenation plant at Gelsenberg, June 25, 1941.

HYDROGENATION DATA, STALL NO. 7

Frames 725-760

Abstract

Detailed monthly reports of the nature of charge stocks, incidents of operation control tests, etc. Examples are shown in prints of frames 726 and 746 attached.

PREHEATERS FOR HIGH PRESSURE CHAMBERS

Frames 761-768

Abstract

Discussion and data for various designs of hairpin tube coal paste preheater (coal paste flowing through tubes heated externally by combustion gases). Temperature gradients and heat transfer coefficients are given.

UPPER SILESIAN COAL TESTS

Frames 769-772

Abstract

Data obtained in large scale test (stall 804) with Upper Silesian coal.

HYDROGENATION PLANT DATA

Frames 773-883

Abstract

Monthly reports of Hydrogenation plant analyses.

Examples for February, 1944 are shown in prints of frames 803 to 823 which are attached.

HYDROGENATION FLOW DIAGRAMS AND OPERATING DATA

Frames 884-895

Abstract

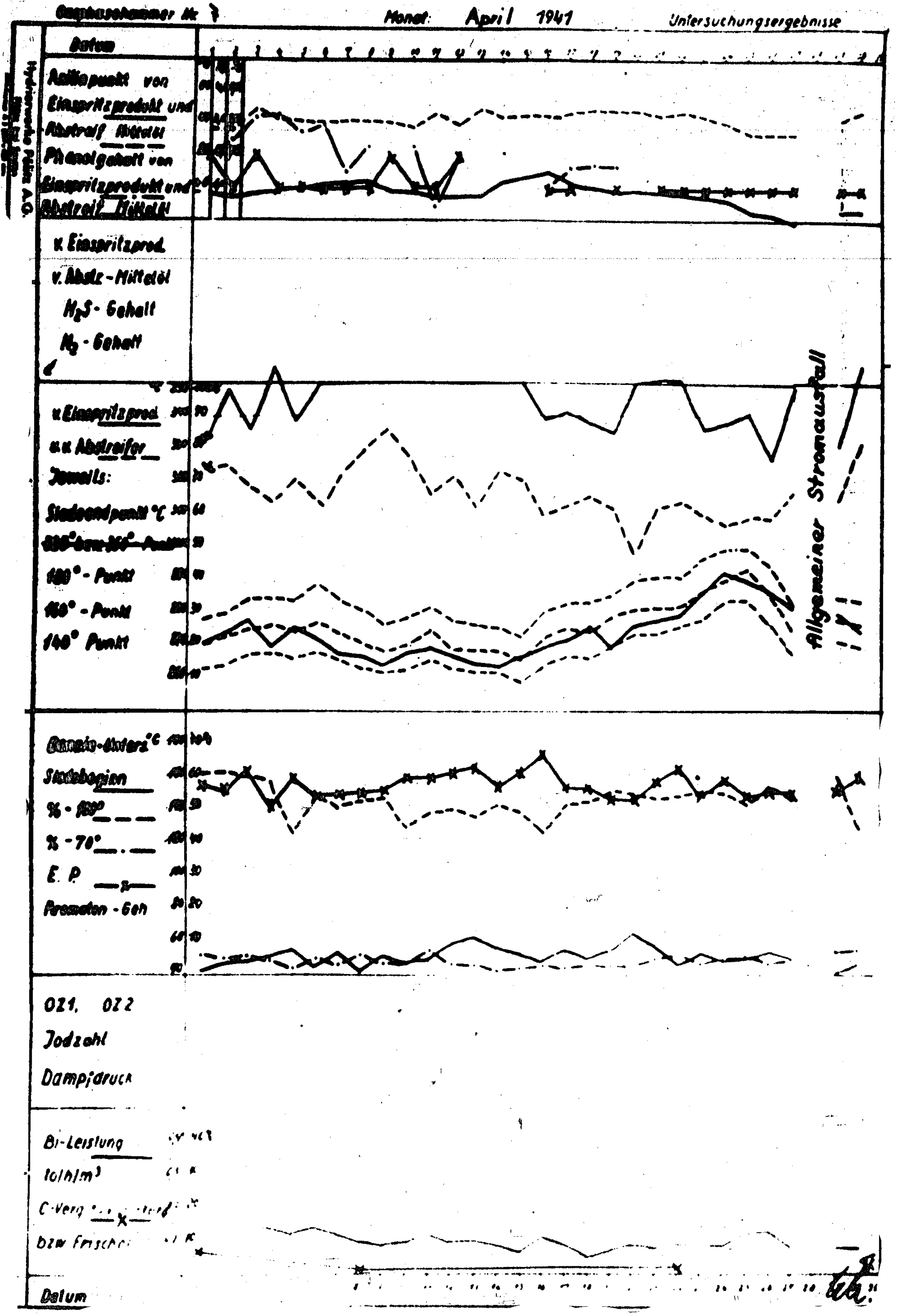
More data on heat regeneration in preheating coal paste(charts for several cases).

HYDRIERWERKE PÖLITZ
AKTIENGESELLSCHAFT

Kammer 7 Januar 1941

Datum	1.-3. Januar	4.-10. Januar	21.-25. Januar	**Gesamt**
Art der Einspritzung	R 403/4 am+bm 3:7	R 403/4 am 1:1 +bm R 403/1/6sm 1:1 3:7	R 403/4 am + 422 bm 3:7	-
Durchsatz to	1524 to	5172 to	1265 to	**7961 to**
Frischoel to	457 to	1552 to	380 to	**2389 to**
Abstreifer to	1545 to	5172 to	1317 to	**8034 to**
Mitteloel to	1434 to	4718 to	1229 to	**7381 to**
Benzin to	111 to	454 to	88 to	**653 to**
Verlust to	-	-	-	-
% Verlust d. Frischoel	-	-	-	-
% C-Vergasung	-	30.1. 9,12 %	23.1. 19,05 %	-
O.Z. vom Benzin	71,7/-	72,6/89,7	68,4/-	-
Benzin-Konzentration	ca. 11,3 %	ca. 9,5 %	ca. 7, %	-
Vom Benzin - 70°	ca. 5,2	ca. 7,5 %	ca. 5,5 %	-
" " - 100°	ca. 53,0 %	ca. 60,0 %	ca. 55,0 %	-
Endpunkt	ca. 161	ca. 165	ca. 156	-

Gasphasekammer Nr 7 Monat: April 1941 Untersuchungsergebnisse



HYDRIERWERKE PÖLITZ A.G.
- Hauptlaboratorium -

Pölitz, den 30. März 44 Sch
BL

Ausführliche Produktuntersuchung der Kammern

vom Monat Februar 1944

Bag 4
Target 30/4.13

Untersuchung der:

1.)	Kohlekammern	13 - 16	Seite	1 - 4
2.)	Erdölkammer	11	„	5 - 6
3.)	Vorhydrierungsk.	5 - 8	„	7 - 9
4.)	Benzinierungsk.	2 - 4	„	10 -12
5.)	Kammer	1a	„	13
6.)	Fertigbenzine		„	14
7.)	Wasseranalysen		„	15
8.)	Gasanalysen		„	16 -20

HAUPTLABORATORIUM
Betriebsleitung

Verteiler:
Dr. Schmitt
BH
BD
BK
BL

Kohlekammern 13 - 16

Tagesdurchschnitt vom 7.2.1944

	Naßkohle	Trockenkohle
	Monatsdurchschnitt	+ Kat. I + II
Rohkohle		
Wasser	8,7 %	3,2 %
Wasserfreie Kohle		
Asche	6,10 %	9,26 %
Alkalität/g H_2SO_4/kg	-	38,7
% S gesamt	-	1,07
Reinkohle		
Flüchtiges	36,5 %	36,7 %
unt.Heizwert/kgcal/kg	-	7428
Elementaranalyse		
% C	81,5	80,9
% H	4,8	4,6
% O		12,3
% N		1,15
% S	-	0,75
g H/100 C	-	5,7
g H/disp.		3,4
	Tagesdurchschnitt	Tagesdurchschnitt
Gesamt	Dickbrei	Dünnbrei
Spez.Gewicht/100°C	1,198	1,161
Wasser	0,9 %	0,4 %
Festes	53,6 %	41,4 %
Asche i.Festen	7,5 %	8,6 %
	Monatsdurchschnitt	Monatsdurchschnitt
	Anreibeöl	Spülöl
Spez.Gewicht/50°C	1,117	1,053
Wasser	0,0 %	0,0 %
Festes	5,2 %	0,4 %
Asche i.Festen	25,3 %	-
Öl (entfestet)		
Spez.Gewicht/50°	1,105	1,053
Gew.% - 325°	1,5	4,6
Asphalt	11,0	≈1.1
Vacuumsiedekurve		
Druck mm Hg	11	11
Siedebeginn	153°	90°
10 Vol.%	201	184
20 "	217	209
30 "	229	221
40 "	240	228
50 "	256	236
60 "	280	246
70 "	325	260
80 "	-	282
90 "	-	325
- 325°	70 %	90 %

2

Abstreiferprodukt

Zerlegung (Kolonne)	**Kammer 13**	**Kammer 15**	**Kammer 16**
Spez.Gewicht/20°	1,023	1,024	1,022
Bi+Mi - 325°C	39,2	41,4	39,2
Rückstand über 325°	60,8	58,6	60,8

Kammer 13-16 im Anfallverhältnis gemischt

Gesamt	**Monatsdurchschnitt**	**Monatsdurchschnitt** **I.Hälfte**	 **II.Hälft**
Spez.Gewicht/20°	1,025	1,025	1,027
Anfall - 325° Gew.%	41,4	41,3	41,9
Elementaranalyse:			
% C	88,2		
% H	8,6		
% O	2,3		
% N	0,82		
% S	0,03		
g H/100 C	9,8		
g H/disp.	9,3		

	Bensin - 165°C	**Mittelöl 165-325°C**
	- Labordestilliert -	
Spez.Gewicht/20°	0,782	0,964
Alkalilösliches	14,0 g/100	27,5 Vol.%
Siedebeginn	57°	171°
10 Vol.%	86	198
20 "	95	209
30 "	101	220
40 "	108	230
50 "	115	242
60 "	122	254
70 "	130	267
80 "	138	280
90 "	151	297
95 "	161	310
- 100°	28 %	-
Endpunkt	166°/ 97 %	329°/99,0 %
Rückstand	1,0 %	1,0 %
Verlust	2,0 %	0,0 %.

		Benzin - 165°C	Mittelöl 165-325°C
Vom alkalilöslichen Anteilen befreit			
Spez.Gewicht/20°		0,768	0,958
Anilinpunkt I		+23,2	-10,3 ber.
Anilinpunkt II		+53,2	
Aromaten + Unges.	Vol.%	29,5	
Naphthene	"	39,0	
Paraffine	"	31,5	

	Rückstand über 325°
Spez.Gewicht/50°	1,054
Stockpunkt	-25°
Viskosität/50°	6,4°E
Vacuumsiedekurve	
Druck mm Hg	11
Siedebeginn	140°
10 Vol.%	180
20 "	198
30 "	209
40 "	218
50 "	228
60 "	237
70 "	251
80 "	273
90 "	310
- 325°	92,5 %

Abschlamm	Kammer 13	Kammer 15	Kammer 16
Spez.Gewicht/100°	1,185	1,181	1,179
Festes	17,4 %	16,0 %	16,1 %
Asche i.F.	40,4 %	42,3 %	42,0 %
Erw.Pkt.(Krä.Sa.)	37,3	32,5	29,7

	Monatsdurchschnitt
Spez.Gewicht/100°	1,198
Festes	19,5 %
Asche i. Festen	42,6 %
Erw.Pkt. (Krä.Sa.)	34,5°
Öl (entfestet)	
Spez.Gewicht/100°C	1,112
Asphalt	16,8 %
Gew.% - 325°	2,1 %
Vaccumsiedekurve	
Druck mm Hg	11
Siedebeginn	130°
10 Vol.%	209
20 "	231
30 "	246
40 "	265
50 "	290
- 325°	58 %

4

Gesamt	**Schleuderöl**	**Schleuderrückstand**	**Schwel**
Spez.Gewicht/100°	1,101	1,302	1,044
Festes	5,6 %	37,8 %	1,8 %
Asche i.Festen	31,1 %	55,6 %	-
Öl (entfestet)			
Spez.Gewicht/100°	1,082	1,082	1,044
Gew.% - 325°	2,5	1;5	2,1
Asphalt	1~~4~~3,2 %	13,4 %	1,2 %
Vacuumsiedekurve			
Druck mm Hg	11	11	11
Siedebeginn	164°	165°	138°
10 Vol.%	207	207	201
20 "	224	224	219
30 "	237	237	229
40 "	255	250	237
50 "	273	271	246
60 "	306	300	257
70 "	-	-	275
80 "	-	-	305
- 325°	63 %	66 %	85 %

	Schwelrückstand (wasserfrei) **Monatsdurchschnitt**
Festes	99,2 %
Asche i.Festen	38,8 %
u.Heizwert; kcal/kg	5135
Elementaranalyse:	
% C	57,0
% H	1,5
% O	0,6
% N	0,66
% S fl.	1,48
% S ges.	3,58
g H/100 C	2,6
g H/disp.	2,1

Sumpfphasekammer 11

Monatsdurchschnittsproben v. Februar 1944

Einspritzung: R 422 + BM 48 : 52

Einspritzung

Gesamt

Spez.Gewicht/50°	0,910
Festes	0,1 %

Öl (entfestet)

Spez.Gewicht/50°	0,910
Asphalt	0,4 %
Vacuumsiedekurve	
Druck mm Hg	11
Siedebeginn	195°
10 Vol.%	237
20 "	250
30 "	265
40 "	281
50 "	301
60 "	325
- 325°	60 %
Elementaranalyse:	
% C	86,9
% H	11,9
% O	0,6
% N	0,37
% S	0,34
g H/100 C	13,7
g H/disp.	13,5

Abstreiferprodukt

Zerlegung(Kolonne)	Monatsdurchschnitt	Halbmonatsdurchschnitt I.Hälfte	Halbmonatsdurchschnitt II.Hälfte
Spez.Gewicht/20°	0,870	0,871	0,871
Anfall - 325° Gew.%	35,2	36,7	33,5
Rückstand über 325°	64,8	63,3	66,5

Elementaranalyse: vom Anfall - 325°C

% C	86,7
% H	12,5
% O	0,4
% N	0,26
% S	0,12
g H/100 C	14,4
g H/disp.	14,2

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Gesamt:	**Bensin -165°C**	**Mi.-Öl 165-325°C**	**Rückstand über 3**
		- labordestilliert -	
Spez.Gewicht/	0,728	0,820	0,895/50°
Alkalilösliches	0,8 Vol.%	3,0 Vol.%	-
Viskosität/50°	-	-	2,8°E
Stockpunkt	-	-	+ 30°C
Siedekurve			
Druck mm Hg			11 mm
Siedebeginn	58°	160°	172°
10 Vol.%	86	189	206
20 "	95	204	220
30 "	102	216	232
40 "	109	227	242
50 "	116	240	253
60 "	123	252	265
70 "	130	264	280
80 "	138	277	298
90 "	148	297	-
95 "	158	312	-
- 100°	27,5 %	-	[illegible]
- 325°	-	97,5 %	88 %
Endpunkt	165/97 %	334/99,0 %	-
Rückstand	1,0 %	1,0 %	-
Verlust	2,0 %	0,0 %	-

Von alkalilöslichen Anteilen befreit

Spez.Gewicht/20°	0,726	0,816
Anilinpunkt I	+53,8°	+61,5°
Anilinpunkt II	+63,5°	
Aromaten + Unges.	17,0 Vol.%	
Naphthene	18,0 "	
Paraffine	65,0 "	

Abschlamm

Gesamt

Spez.Gewicht/100°	0,968
Festes	16,5 %
Asche i.Festen	52,0 %

Öl (entfestet)

Spez.Gewicht/50°	0,918
Gew.% -325°	3,9
Asphalt	0,7 %
Vacuumsiedekurve	
Druck mm Hg	11
Siedebeginn	112°
10 Vol.%	194
20 "	221
30 "	239
40 "	255
50 "	272
60 "	291
70 "	317
- 325°	72,5 %

Elementaranalyse:

	Festes
% C	45,2
% H	1,3
% O	0,4
% N	0,40
% S ges.	6,67
% S fl.	1,54
% H/100 C	2,9
% H/disp.	2,5

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Vorhydrierungskammern

Tagesdurchschnitt vom 3.2.1944

Einspritzung: 50 %) 5 % R 213/215/216/217 o.Bi
45 % R 213/215/216/217 asm 72:28

50 %) 7 % R 422 am + a Bi
43 % R 422 am + a Bi

Einspritzprodukt	Tagesdurchschnitt	Monatsdurchschnitt
Spez.Gewicht/20°	0,883	0,877
Anilinpunkt	+34,0° C	+36,3° C
Alkalilösliches	7,3 Vol.%	0,0 Vol.%
Siedebeginn	55°	65°
10 Vol.%	122	[illegible]
20 "	153	170
30 "	183	195
40 "	209	219
50 "	230	240
60 "	249	260
70 "	270	282
80 "	291	303
90 "	314	325
95 "	325	339
- 140°	15,5 %	10,5 %
- 165°	23,5 %	18,5 %
- 325°	95,0 %	90,0 %
Endpunkt	341°/ 98,0 %	348°/ 98,0 %
Rückstand	2,0 %	2,0 %
Fe	0,7 mg/l	-

Fraktionen:	Sp.Gew./20°	A.P.X (entph.)
140-160°	0,791	+41,4
190-210°	0,864	+33,4
240-260°	0,917	+36,1
290-310°	0,940	+35,8

Elementaranalyse:	
% C	86,8
% H	11,4
% O	1,1
% N	0,41
% S	0,25
g H/100 C	13,1
g H/disp.	12,8

Abstreiferprodukt	Kammer 1	Kammer 5	Kammer 6	Kammer 7	Kammer 8
Spez.Gewicht/20°	0,803	0,822	0,812	0,804	0,808
Anilinpunkt	+56,2	+59,6	+58,4	+58,1	+55,9
Phenole g/100 ccm	u.0,01	u.0,01	u.0,01	u.0,01	0,01
Siedebeginn	56°	94°	89°	66°	62°
10 Vol.%	105	131	122	111	118
20 "	131	157	146	137	137
30 "	153	179	165	159	158
40 "	176	199	183	179	178
50 "	197	219	201	198	198
60 "	217	237	220	218	220
70 "	239	256	239	239	[illegible]
80 "	262	270	262	262	267
90 "	294	302	[illegible]	290	[illegible]
95 "	314	[illegible]	[illegible]	310	317

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	Kammer 1		Kammer 5		Kammer 6		Kammer 7		Kammer 8
- 140°	24,5 %		13,0 %		17,5 %		21,5 %		21,5 %
- 165°	35,5 %		23,5 %		30,0 %		33,0 %		33,5 %
Endpunkt	3250/98,0 %		332/98 %		324/98 %		329/98 %		331/98 %
Rückstand	1,5 %		1,5 %		1,5 %		1,5 %		1,5 %
Verlust	0,5 %		0,5 %		0,5 %		0,5 %		0,5 %
Fraktionen:	Sp.Gew./20°	A.P. I	Sp.Gew./20°	A.P. I	Sp.Gew./20°	A.P. I	Sp.Gew./20°	A.P. I	Sp.Gew./20
90-110°	0,742	+51,1	0,735	+51,2	0,742	+48,8	0,728	+50,3	0,726
140-160°	0,744	+50,4	0,785	+51,5	0,785	+51,4	0,774	+51,4	0,778
190-210°	0,821	+52,3	0,821	+54,1	0,820	+54,3	0,819	+54,2	0,822
240-260°	0,848	+59,2	0,849	+61,6	0,852	+61,3	0,843	+62,5	0,851
290-310°	0,860	+71,2	0,865	+72,1	0,854	+73,4	0,854	+74,8	0,864

	Techn.Benzin laborstabilisiert v.4.2.44	Techn. Mittelöl v.4.2.44
Spez.Gewicht/20°	0,7165	0,838
Anilinpunkt I	+50,8	
Anilinpunkt II	+53,9	
Aromaten + Unges.	4,5 Vol.%	
Naphthene	50,5 "	
Paraffine	45,0 "	
Dampfdruck/37,8°C	0,46	
Cu-Streifentest	negativ	
Doktortest	negativ	
Klopfwert	69,5	
" + 0,115 Vol.% Pb	90,5	
Siedebeginn	42°	170°
5 Vol.%	58	188
10 "	65	194
20 "	73	203
30 "	79	212
40 "	83	222
50 "	87	233
60 "	91	249
70 "	94,5	265
80 "	98,5	280
90 "	104	300
95 "	111,5	322
- 70°	17,0 %	-
- 100°	83,0 %	-
- 300°	-	90,0 %
- 325°	-	96,0 %
Endpunkt	116°/ 97,0 %	334°/ 99,0 %
Rückstand	1,5 %	1,0 %
Verlust	1,5 %	-
% C	-	86,5
% H	-	13,5

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Anhang zur Untersuchung
der Vorhydrierungskammern vom 3.2.1944

Benzin - 165°
laborddestilliert und stabilisiert

	Kammer 1	Kammer 3	Kammer 6	Kammer 7	Kammer 8	
Spez.Gewicht/20°	0,741	0,754	0,756	0,747	0,742	
Anilinpunkt I	+50,8	+49,2	+49,9	+50,6	+50,6	
Anilinpunkt II	+57,4	+54,2	+54,8	+55,5	+55,7	
Aromaten +Unges.	8,0	5,0	6,0	5,5	5,5	Vol.%
Naphthene	38,0	49,0	47,0	45,0	44,5	"
Paraffine	54,0	46,0	47,0	49,5	50,0	"
Klopfwert	58,5	53,8	54,4	57,0	57,2	
" + 0,125 % Pb	81,0	77,6	78,2	79,9	80,1	
Dampfdruck	0,303	0,164	0,120	0,199	0,272	
Siedebeginn	54°	79°	91°	66°	62°	
10 Vol.%	78	94	99	87	82	
20 "	90	101	104	95	92	
30 "	99	106	108	102	99	
40 "	106	112	113	108	107	
50 "	112	117	118	113	112	
60 "	119	123	124	120	119	
70 "	127	130	132	128	127	
80 "	136	140	142	137	137	
90 "	153	155	155	150	152	
95 "	164	165	165	160	161	
- 70°	5,5 %	-	-	1,0 %	3,0 %	
- 100°	31,0 %	17,5 %	11,5 %	26,5 %	29,0 %	
Endpunkt	169°/97,0 %	169/97 %	169/97 %	169/98 %	169/98 %	
Rückstand	1,0 %	1,0 %	1,0 %	1,0 %	1,0 %	
Verlust	2,0 %	2,0 %	2,0 %	1,0 %	1,0 %	

Mittelöl über 165°
labordestilliert

Spez.Gewicht/20°	0,839	0,[illegible]	0,[illegible]	0,83[illegible]	0,845
Anilinpunkt	+59,8	+63,0	+62,7	+62,4	+60,0
Siedebeginn	181°	184°	183°	182°	185°
10 Vol.%	198	203	198	195	200
20 "	207	212	207	205	209
30 "	216	222	216	214	219
40 "	225	233	225	224	229
50 "	235	245	236	234	240
60 "	249	259	247	246	253
70 "	264	271	268	260	266
80 "	280	287	277	277	284
90 "	302	308	299	299	304
95 "	320	325	316	316	320
- 325°	96,5 %	95,0 %	96,5 %	97,0 %	96,5 %
Endpunkt	338/98,0 %	343/98 %	334/98,0 %	331/98 %	336/98,0 %
Rückstand	2,0 %	2,0 %	2,0 %	2,0 %	2,0 %

Benzinierungskammern

Tagesdurchschnitt vom 3.2.1944

Einspritzung: R 213/215/216/217/451/422 b + cm 70:30

Einspritzprodukt	Tagesdurchschnitt	Monatsdurchschnitt
Spez.Gewicht/20°	0,825	0,827
Anilinpunkt	+61,3	+61,3
Phenole g/100 ccm	u.0,01	0,01
Siedebeginn	162°	165°
10 Vol.%	180	188
20 "	191	198
30 "	200	206
40 "	208	214
50 "	219	224
60 "	234	236
70 "	251	253
80 "	270	275
90 "	294	301
95 "	314	318
Endpunkt	338°/98,0 %	332°/98,0 %
Rückstand	2,0 %	2,0 %

Fraktionen:	Sp.Gew./20°	A.P.I
190-210°	0,817	+55,8
240-260°	0,841	+63,0
290-310°	0,853	+75,5

Elementaranalyse:		
% C	-	86,0
% H		13,9
g H/100 C		16,2

Abstreiferprodukt	Kammer 2	Kammer 3	Kammer 4
Spez.Gewicht/20°	0,744	0,744	0,742
Anilinpunkt	+58,1	+58,6	+58,7
Siedebeginn	38°	39°	37°
10 Vol.%	67	70	67
20 "	89	87	87
30 "	108	103	103
40 "	126	119	120
50 "	145	137	136
60 "	164	154	154
70 "	184	174	173
80 "	207	195	194
90 "	238	225	225
95 "	-	-	258
- 140°	47,5 %	51,5 %	54,0 %
- 165°	61,0 %	65,5 %	66,5 %
Endpunkt	269°/ 95,0 %	260°/95,0 %	270/96,5 %
Rückstand	1,5 %	1,5 %	1,5 %
Verlust	3,5 %	3,5 %	2,0 %

Fraktionen:	Sp.Gew./20°	A.P.I	Sp.Gew./20°	A.P.I	Sp.Gew./20°	A.P.I
90-110°	0,722	+54,1	0,721	+55,1	0,720	+54,7
140-160°	0,773	+54,6	0,772	+55,4	0,772	+55,6
190-210°	0,805	+60,1	0,806	+61,5	09806	+61,2

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	Techn.Benzin [illegible] v.4.-[illegible]44	Techn.Mittelöl v.4.-[illegible]1944
Spez.Gewicht/20°	0,7135	0,797
Anilinpunkt I	+55,8	
Anilinpunkt II	+58,7	
Aromaten + Unges. Vol.%	5,0	
Naphthene "	35,5	
Paraffine "	59,5	
Dampfdruck/37,8°C	0,44	
Cu-Streifentest	negativ	
Doktortest	negativ	
Klopfwert	71,6	
" + 0,115 Vol.% Pb	91,3	
Siedebeginn	41°	149°
5 Vol.%	56	159
10 "	60	164
20 "	68,5	173
30 "	77	178
40 "	85	184
50 "	94	191
60 "	102	198
70 "	112	209
80 "	124	224
90 "	139	250
95 "	151	274
- 70°	22,0 %	-
- 100°	58,0 %	-
Endpunkt	153°/ 98,0 %	293°/ 98,0 %
Rückstand	1,0 %	1,0 %
Verlust	1,0 %	1,0 %

Anhang zur Untersuchung der Benzinierungskammern vom 5.2.1944

Benzin - 165°
labordestilliert und stabilisiert

	Kammer 2	Kammer 3	Kammer 4
Spez.Gewicht/20°	0,716	0,723	0,715
Anilinpunkt I	+55,2	+55,7	+55,9
Anilinpunkt II	+59,1	+60,3	+59,4
Aromaten + Unges.	6,5 Vol.%	6,5 Vol.%	5,5 Vol.%
Naphthene	33,5 "	29,5 "	32,5 "
Paraffine	60,0 "	64,0 "	62,0 "
O.Z.	69,4	69,5	69,4
+ 0,125 % Pb	89,7	89,2	89,8
Dampfdruck	0,470	0,431	0,453
Siedebeginn	39°	49°	42°
10 Vol.%	58	64	58
20 "	69	74	69
30 "	79	85	79
40 "	89	92	88
50 "	99	100	98
60 "	110	109	108
70 "	120	119	119
80 "	133	132	132
90 "	151	149	148
95 "	163	160	160
- 70°	21,0 %	15,0 %	21,0 %
- 100°	51,0 %	50,0 %	52,5 %
Endpunkt	169°/ 97,0 %	169°/98,0 %	169°/ 98,0 %
Rückstand	1,0 %	1,0 %	1,0 %
Verlust	2,0 %	1,0 %	1,0 %

Mittelöl über 165°
labordestilliert

	Kammer 2	Kammer 3	Kammer 4
Spez.Gewicht/20°	0,807	0,803	0,804
Anilinpunkt	+63,1	+63,5	+63,8
Siedebeginn	177°	171°	174°
10 Vol.%	184	178	183
20 "	188	183	187
30 "	192	188	190
40 "	196	192	194
50 "	200	196	199
60 "	207	202	206
70 "	216	211	214
80 "	230	224	227
90 "	254	249	252
95 "	274	268	272
Endpunkt	301°/98,0 %	287°/98,0 %	298°/ 98,0 %
Rückstand	2,0 %	2,0 %	2,0 %

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Kammer 1a

Tagesdurchschnitt vom 5.2.1944

	Einspritzprodukt	Abstreiferprodukt
Spez.Gewicht/20°	0,7290	0,7041
Anilinpunkt	+46,3	+77,5
Klopfwert (M.M.)	-	96,7
Brechungsindex/20°	1,4171	1,3979
Siedebeginn	98°	94
5 Vol.%	100,5	98
10 "	101,5	99
20 "	103	101
30 "	104,5	102
40 "	106	103
50 "	107,5	104
60 "	110	105,5
70 "	114	108
80 "	124	114
90 "	170	135
95 "	178	178
- 100°	3,5 %	15,0 %
- 130°	83,0 %	88,0 %
Endpunkt	184°/ 98,0 %	186°/ 98,0 %
Rückstand	1,0 %	1,0 %
Verlust	1,0 %	1,0 %
Reaktion des Rückst.	neutral	neutral
Jodzahl	193	0,3
Kupferstreifentest	negativ	negativ
Doktortest	negativ	negativ
Abblastest	0,4 mg	0,6 mg

Abstreiferprodukt

Fraktionen	- 130°C	über 130°C
Ausbeute (Gew.%)	88,5	11,5
Spez.Gewicht/20°	0,696	0,763
Anilinpunkt	+76,5	+80,5
Jodzahl	0,0	0,4
Klopfwert	97,8	88,8
Brechungsindex	1,3939	1,4272

Fertigprodukte (Benzin).

	VT 708 Tank 217 P.IX/401 v.29.2.44	DHD Tank 220 P. 1235 v.28.2.44	HP 110 Tank 151 P. v.24.2.44
Spez.Gewicht/20°	0,7072	0,7683	0,7015
Anilinpunkt I	+55,4	+5,4	+77,7
Anilinpunkt II	+58,9	+60,6	
Arom.+Unges. Vol.%	4,5	47,5	
Naphthene "	35,0	16,5	
Paraffine "	60,5	36,0	
Jodzahl(Meth.38)	1,0	1,4	0,0
Kupferstreifentest	negativ	negativ	negativ
Doktortest	negativ	negativ	negativ
Dampfdruck/37,8°C	0,45	0,46	0,18
Schmelzpunkt	u.-60°	u.-60°	-
Klopfwert (M.M.)	70,5	78,8	97,1
" + 0,115 Vol.% Pb	92,4	93,0	-
Siedebeginn	42°	39°	87°
5 Vol.%	55,5	51,5	97
10 "	59	58	98
20 "	64,5	68,5	100
30 "	71	83	101
40 "	78	98	102
50 "	86	114	103
60 "	93,5	124	105
70 "	101	137	107
80 "	110	145	113
90 "	124	156	154
95 "	138	165	179
- 70°	29,0 %	21,5 %	-
- 100°	69,0 %	41,0 %	-
- 130°	-	-	88,0 %
Endpunkt	148°/98,0 %	173°/98,0 %	187°/98,0 %
Rückstand	1,0 %	1,0 %	1,5 %
Verlust	1,0 %	1,0 %	0,5 %

Wasser - Untersuchungen : g/Ltr.

Gasphase - Kammern

Prod.	Datum	Kammer	NH_3	Cl	Phenole	H_2S	CO_2	Abdampf-rückst
Abstr. Wasser	3.2.44	1	30,09	0,018	0,15	31,96	3,52	1,73
		5	29,92	0,021	0,015	19,38	3,52	0,20
		6	38,59	0,024	0,12	26,52	4,40	0,95
		7	34,40	0,011	0,14	21,42	3,52	0,35
		8	30,77	0,007	0,15	1,70	7,92	0,14
"	5.2.44	1a	2,21	0,011	0,007	0,27	1,06	0,73
"	5.2.44	2	0,41	0,007	0,006	0,68	0,88	0,35
		3	0,34	0,004	0,001	0,14	0,93	0,05
		4	0,34	0,005	0,001	0,27	0,35	0,20

Sumpf - Kohle - Kammern

Prod.	Datum	Kammer	NH_3	Cl	Phenole	H_2S	CO_2	Abdampf-rückst
Abstr. Wasser	10.2. 1944	11	4,76	0,011	0,30	2,04	5,28	0,23
Einspr. Wasser	"	11	0,34	0,011	0,01	-	-	-
Abstr. Wasser	8.2. 1944	13	2 21,98	0,008	3,5	0,85	27,28	2,3
		15	8,09	0,004	4,0	1,87	20,24	2,2
		16	20,94	0,002	4,0	0,68	26,84	1,9

Gas - Untersuchungen

Reichgase der Sumpfkammern

		24.-25.2.1944	17.-18.2.1944		
Kammer		11	13+14	15	16
CO_2		3,6	13,4	11,0	11,0
H_2S	Vol.%	1,9	0,7	0,6	0,6
H_2S	g/nm^3	24,7	9,96	8,03	8,03
NH_3	Vol.%	1,1	10,0	13,3	13,3
NH_3	g/nm^3	7,69	70,8	94,2	94,2
CO		0,8	0,4	0,5	0,8
H_2		28,7	7,5	5,5	8,0
N_2		2,9	3,9	4,2	0,6
C_1		14,4	7,7	7,8	8,4
C_2H_6		11,5	10,3	10,9	9,3
C_2H_4		0,0	0,0	0,1	0,1
C_3H_8		12,4	12,7	14,6	17,0
C_3H_6		0,1	0,4	0,4	0,5
i.C_4H_{10}		i.+n.10,4	3,7	3,3	3,9
i.C_4H_8 + nC_4H_8		0,6	1,0	1,0	0,8
n.C_4H_{10}		-	7,1	8,5	8,2
C_5	Vol.%	11,6	21,2	18,3	17,5
C_5	g/nm^3	519,0	926,5	800,5	775,0
KW		61,0	64,1	64,9	65,7
C-Z		2,90	3,44	3,35	3,01
Dichte		1,4087	2,0106	1,9351	1,8980

Gas - Untersuchungen

Armgase der Sumpfphasekammern

		24.-25.2.1944	17.-18.2.1944			
Kammer		11	13	14	15	16
CO_2		0,8	3,4	4,9	4,7	3,7
H_2S	Vol.%	0,2	0,1	0,2	0,2	0,1
H_2S	g/nm³	3,03	1,82	1,37	2,39	1,7
NH_3	Vol.%	0,1	0,6	0,8	0,7	0,3
NH_3	g/nm³	0,50	4,36	5,57	4,62	2,4
CO		1,4	2,0	2,0	2,0	2,4
H_2		62,4	51,7	53,9	51,3	50,4
N_2		6,8	7,0	5,2	7,1	3,[illegible]
C_1		18,1	21,4	20,2	20,8	21,[illegible]
C_2		5,2	7,4	7,0	7,1	9,7
C_3		3,3	4,1	3,7	4,4	5,1
$i.C_4H_{10}+n.C_4H_{10}$		1,2	1,6	1,4	1,3	2,0
C_5	Vol.%	0,5	0,7	0,7	0,4	0,9
C_5	g/nm³	20,8	32,2	29,4	18,9	37,8
KW		28,3	35,2	33,0	34,0	39,3
C-Z		1,61	1,66	1,63	1,63	1,75
Dichte		0,4397	0,4340	0,5893	0,5955	0,62

Ausgänge der Sumpfphasekammern

		24.-25.2.44	17.-18.2.1944			
Kammer		11	13	14	15	16
CO_2		0,2	0,4	0,5	0,4	0,5
H_2S	Vol.%	0,0	0,0	0,0	0,0	0,0
H_2S	g/nm³	0,59	0,25	0,30	0,25	0,3
NH_3	Vol.%	0,0	0,0	0,0	0,0	0,0
NH_3	g/nm³	0,001	0,04	0,09	0,07	0,0
CnHm		0,2	0,2	0,1	0,6	0,1
CO		2,1	1,5	2,4	2,2	2,1
H_2		75,9	75,4	69,3	74,6	69,[illegible]
N_2		8,8	7,4	9,0	7,3	8,[illegible]
KW		12,8	15,1	18,7	14,9	16,[illegible]
C-Z		1,39	1,49	1,43	1,31	1,[illegible]
Dichte		0,3026	0,3137	0,3621	0,3223	0,[illegible]

Gas - Untersuchungen

17.-18.2.1944

Behälter		Schwelrestgas	Abschlammgas
CO_2		1,5	1,0
H_2S	Vol.%	0,4	0,0
H_2S	g/nm³	3,24	0,31
NH_3	Vol.%	0,1	0,2
NH_3	g/nm³	0,53	1,72
CO		0,9	2,3
H_2		83,6	68,5
N_2		0,2	6,6
C_1		8,7	14,1
C_2		[illegible] 2,7	3,6
C_3		[illegible] 7,3	1,8
C_4		[illegible] 0,4	0,8
C_5	Vol.%	0,2	1,1
C_5	g/nm³	8,3	49,7

Waschanlage

		24.-25.2.44	24.-25.2.44	18.-25.2.44	24.-25.2.44
		Behälter 1 + 2	Behälter 3 + 4	Reichgas	Ausgang
CO_2		0,7	0,4	2,4	0,2
H_2S	Vol.%	0,1	0,0	0,7	0,1
H_2S	g/nm³	0,15	0,63	9,24	0,17
NH_3	Vol.%	0,0	0,0	0,0	0,0
NH_3	g/nm³	0,01	0,01	0,14	0,02
CnHm		-	-	-	0,0
CO		2,0	2,0	0,7	1,8
H_2		56,3	53,7	8,9	76,0
N_2		9,8	7,4	0,9	8,7
C_1		21,2	24,6	19,3	-
C_2		6,9	8,4	25,1	-
C_3		2,5	3,0	26,4	-
C_4		0,5	-	-	-
$i.C_4H_{10}$		-	} 0,4	2,2	-
$n.C_4H_{10}$		-	-	10,2	-
C_5	Vol.%	0,0	0,1	4,1	-
C_5	g/nm3	1,5	3,1	151,5	-
KW		31,1	36,5	87,3	13,3
C-S		1,43	-1,44	2,30	1,33
Dichte		[illegible]		1,366	[illegible]

16
3,7
0,1
1,76
0,3
2,46
2,4
50,4
3,8
21,8
9,7
5,1
2,0
0,9
37,8
39,3
1,75
0,628
16
0,5
0,0
0,33
0,0
0,09
0,1
2,1
69,8
8,8
18,7
1,58
0,36

55

Gas - Untersuchungen:

Reichgase der Gasphasekammern

		5.-6.2.1944			3.-4.2.1944	
Kammer		1a	2	3+4	1+5+6	7+8
CO_2		0,2	0,2	0,2	0,1	0,0
H_2S	Vol.%	0,0	2,7	2,3	0,8	0,0
H_2S	g/nm³	-	36,3	30,4	10,74	0,45
NH_3	Vol.%	0,0	0,0	0,0	0,6	0,9
NH_3	g/nm³	0,0082	0,005	0,01	4,33	6,2835
CO		0,3	0,2	0,2	0,5	0,1
H_2		76,5	12,4	10,5	27,7	23,4
N_2		3,6	4,6	2,8	5,0	8,0
C_1		7,6	6,0	4,1	22,4	27,0
C_2		1,3	4,1	2,1	8,9	11,6
C_3		0,4	21,6	21,1	15,0	12,2
iC_4H_{10}	+ iC_4H_8	8,1	24,8	33,8	8,1	6,1
nC_4H_{10}	+ nC_4H_8	0,4	11,4	11,6	5,1	5,7
C_5	Vol.%	1,6	12,0	11,7	5,8	5,0
C_5	g/nm³	65,4	438,0	430,8	112,0	204,0
KW		19,4	79,9	84,4	65,3	67,6
C-Z		2,75	3,55	3,69	2,56	2,35
Dichte		0,4591	1,9256	2,1065	1,1758	1,1613

Abstreifer der Gasphasekammern

	5.-6-2-1944			3.-4.2.1944	
Kammer	1a	2	3+4	1,5,6	7+8
CO_2	0,2	0,1	0,1	0,1	0,1
H_2S	0,0	0,3	0,4	0,7	0,8
NH_3	0,0	0,0	0,0	0,0	0,6
CO	0,0	-	-	0,0	0,0
H_2	0,0	-	-	0,0	0,0
C_1	0,0	0,0	0,0	0,0	-
C_2	0,1	0,4	0,2	0,7	0,5
C_3	0,2	7,7	10,8	4,4	3,2
$i.C_4H_{10}{}^{+8}$	10,7	40,2	46,5	7,9	5,7
$nC_4H_{10}{}^{+8}$	1,0	16,1	26,9	7,2	7,9

Gas - Untersuchungen

Armgase der Gasphasekammern

	5.-6.2.	5.-6.2.1944			3.-4.2.1944			
Kammer	1	2	3	4	5	6	7	8
CO_2	0,0	0,1	0,1	0,1	0,0	0,0	0,0	0,0
H_2S Vol.%	0,1	0,6	0,4	0,8	0,0	0,1	0,0	0,0
H_2S g/nm³	0,71	8,17	5,88	10,27	0,16	1,18	0,23	0,19
NH_3 Vol.%	0,2	0,0	0,0	0,0	0,2	0,2	0,2	0,2
NH_3 g/nm3	1,4	0,0088	0,0052	0,0033	1,58	1,76	1,14	1,0564
CO	0,7	1,2	0,9	0,8	0,1	0,4	0,1	0,4
H_2	69,2	66,7	68,2	68,5	67,7	71,4	70,5	70,3
N_2	14,7	17,3	15,3	13,6	15,6	13,4	13,0	13,5
C_1	10,3	7,2	6,8	7,1	12,5	11,5	13,7	12,0
C_2	2,6	0,7	1,4	1,7	2,8	2,0	1,2	2,7
C_3	1,3	2,3	2,6	2,8	0,4	0,6	0,9	0,4
C_4	0,7	2,7	3,4	3,8	0,5	0,2	0,3	0,3
C_5 Vol.%	0,3	1,2	1,1	0,9	0,2	0,2	0,1	0,2
C_5 g/nm³	9,7	44,3	42,1	33,0	8,8	7,0	4,0	8,2
KW	15,2	14,1	15,3	16,3	16,4	14,5	16,2	15,6
C-Z	1,56	2,29	2,40	2,37	1,36	1,32	1,27	1,33
Dichte	0,3889	0,4865	0,4811	0,4795	0,2824	0,3436	0,3428	0,2154

Ausgänge der Gasphasekammern

	5.-6.2.	3.-4.2.	5.-6.2.44			3.-4.2.1944			
Kammer	1a	1	2	3	4	5	6	7	8
CO_2	0,03	0,00	0,0	0,02	0,03	0,02	0,00	0,00	0,01
H_2S Vol.%	-	-	0,0	-	-	-	-	-	-
H_2S g/nm³	0,00	0,19	0,34	1,41	1,75	0,02	0,04	0,01	0,01
NH_3 Vol.%	-	-	0,0	-	-	-	-	-	-
NH_3 g/nm³	0,0001	0,0603	0,0005	0,004	0,0014	0,062	0,1024	0,0919	0,0732
CO	-	-	0,2	-	-	-	-	-	-
H_2	-	-	81,3	-	-	-	-	-	-
N_2	-	-	13,6	-	-	-	-	-	-
KW	-	-	8,9	-	-	-	-	-	-
C-Z	-	-	2,04	-	-	-	-	-	-
Dichte	-	-	0,2865	-	-	-	-	-	-

INCREASING THROUGHPUT IN TAR CHAMBERS 11 AND 12

Frames 896-905

Abstract

Report from I.G. calling attention to errors in temperature measurements and methods by which production of the tar hydrogenation stalls 11 and 12 could be improved.

GAS ANALYSIS OF PRODUCTS AT HYDRIERWERKE-POLITZ

Frames 906-958

Abstract

Data on gas formation in various hydrogenation operations.

SAMPLING OF STALL PRODUCTS POLITZ, SEPTEMBER 4, 1942

Frames 959-960

Abstract

Communication from head laboratory calling attention to improper sampling techniques which are being employed. Open containers are used and permit loss of volatile components from gasoline samples. The daily samples do not represent an average composition since they are not taken at a constant rate. Sample lines are not blown out and dirty containers are used in some instances. It is suggested that a reliable man be placed in charge of the sampling program.

STUDY OF RECYCLE GAS LEAVING THE STALLS DURING APRIL POLITZ, MAY 9, 1942

Frames 961-965

Abstract

The cycle gas from the coal stalls showed a relatively high CO_2 content of 0.26 vol. % on the average; in the collecting mains for the cycle gas from the coal stalls there is, however, only a very low ammonia content after the water wash (average 2.3 mg./nm^3) so that the formation of $(NH_4)_2CO_3$ is improbable.

Similar proportions occur also in the cycle gas from the tar stalls with an average CO_2 content of 0.13 vol. % and NH_3 of 9.0 mg./nm^3.

The NH_3 content of the 5058 cycle gas (stall 5) is relatively high (205 mg./nm^3); however, as a result of the lower CO_2 content (ave. 0.02 vol. %), the formation of ammonium carbonate is not expected.

In 6434 cycle gas the CO_2 and NH_3 contents of 0.01 vol. % and 1.4 mg./nm^3 respectively are particularly low.

REPORT ON CONDITION OF THE CHAMBERS AS OF JUNE, 1942

Frames 966-972

Abstract

The condition of the chambers in June, 1942, is described.

PLANT PRODUCTION DATA, HYDRIERWERKE-POLITZ

Frames 974-1000

Abstract

The production data of Hydrierwerke-Politz is shown.

Examples are shown in the attached prints of frames 974, 976 and 981.

PLANT PRODUCTION DATA, HYDRIERWERKE-POLITZ

Frames 1001-1041

Abstract

Daily production reports to Arsyn of Hydrierwerke-Politz are shown.

Examples are shown in attached prints of 1031, 1035, 1036 and 1037. The bombing effect is shown in the latter 3 prints.

[illegible] Pölitz Aktiengesellschaft

Stettin-Pölitz, den [illegible] Juni 1941 [illegible]

PRODUCTION PROGRAM FOR 1941 AND FIRST QUARTER OF 1942 ON THE BASIS OF THE CHAMBER DATA OF JUNE 6, 1941 (CONFERENCE IN R.[illegible]A. BERLIN)

Vertraulich

	Januar bis Mai	Juni	Juli	August	September	Oktober	November	Dezember	Summe Juni - Dezember 1941	Januar	Februar	März	Summe Januar-März 1942
Kammer 11 (oooo) 35 m³		9 800 7 500	9 800 7 500	9 800 7 500	9 800 7 500	9 800 7 500	9 800 7 500	9 800 7 500	68 600 52 500	9 800 7 500	9 800 7 500	9 800 7 500	29 400 22 500
Kammer 12 (oooo) 35 m³		9 800 7 500	9 800 7 500	9 800 7 500	9 800 7 500	9 800 7 500	9 800 7 500	9 800 7 500	68 600 52 500	9 800 7 500	9 800 7 500	9 800 7 500	29 400 22 500
Kammer 13 (oooo) 35 m³	Einspritztermin 1.6.	4 600 2 600	9 250 5 200	9 250 5 200	9 250 5 200	9 250 5 200	9 250 5 200	9 250 5 200	60 100 33 800	9 250 5 200	9 250 5 200	9 250 5 200	27 750 15 600
Kammer 14 (oooo) 35 m³		Einspritztermin 25.7.	-	6 200 3 500	9 250 5 200	9 250 5 200	9 250 5 200	9 250 5 200	43 200 24 300	9 250 5 200	9 250 5 200	9 250 5 200	27 750 15 600
Kammer 15 (oooo) 35 m³	für Kohle: Bi + M'Öl Leistung 0,23 für Teer + Pech: Bi + M'Öl Leistung 0,33			Einspritztermin 15.9.	1 150 650	8 100 4 550	9 250 5 200	9 250 5 200	27 750 15 600	9 250 5 200	9 250 5 200	9 250 5 200	27 750 15 600
Kammer 16 (oooo) 35 m³							Einspritztermin 15.12.	1 150 650	1 150 650	8 100 4 550	9 250 5 200	9 250 5 200	26 600 14 950
Einsatz:	f. S-Kammern	19 600	19 600	19 600	19 600	19 600	19 600	19 600	137 200 Teere +Pech usw.	19 600	19 600	19 600	58 800 Teere + Pech usw.
Produkt:		15 000	15 000	15 000	15 000	15 000	15 000	15 000	105 000 S-M'Öl	15 000	15 000	15 000	45 000 S-M'Öl
Einsatz:	f. Kohle-Ka.	4 600	9 250	15 450	19 650	26 600	27 750	28 900	132 200 Kohle	35 850	37 000	37 000	109 850 Kohle
Produkt:		2 600	5 200	8 700	11 050	14 950	15 800	16 250	74 350 S-M'Öl	20 150	20 800	20 800	61 750 S-M'Öl
Ges. Einsatz	f. Gasphase	17 600	20 200	23 700	26 050	29 950	30 800	31 250	179 350 S-M'Öl	35 150	35 800	35 800	106 750 S-M'Öl ges.
Kammer 3 (ooooo) 24 m³		5 000 3 750	5 000 3 750	5 000 3 750	5 000 3 750	5 000 3 750	5 000 3 750	5 000 3 750	35 000 26 250	5 000 3 750	5 000 3 750	5 000 3 750	15 000 11 250
Kammer 2 (ooooo) 24 m³	Leistung 0,24 eingesetzt Ausbeute 75 % eingesetzt		Einspritztermin 1.8.	2 500 1 900	5 000 3 750	5 000 3 750	5 000 3 750	5 000 3 750	22 500 16 900	5 000 3 750	5 000 3 750	5 000 3 750	15 000 11 250
Einsatz:		5 000	5 000	7 500	10 000	10 000	10 000	10 000	57 500 S-M'Öl	10 000	10 000	10 000	30 000 S-M'Öl
Produkt:		3 750	3 750	5 650	7 500	7 500	7 500	7 500	43 150 CV_2-Bi	7 500	7 500	7 500	22 500 CV_2-Bi
Kammer 6 (ooo) 24 m³		10 000 9 300	10 000 9 300	10 000 9 300	10 000 9 300	10 000 9 300	10 000 9 300	10 000 9 300	70 000 65 100	10 000 9 300	10 000 9 300	10 000 9 300	30 000 27 900
Kammer 7 (ooo) 24 m³		2 600 2 400	5 200 4 800	6 200 5 800	6 050 5 600	9 950 9 250	10 600 9 850	11 250 10 450	51 850 46 150	10 000 9 300	10 000 9 300	10 000 9 300	30 000 27 900
Kammer 8 (ooo) 24 m³	Frischdurchsatz max. 0,71 eingesetzt, Ausbeute 93 % eingesetzt								Einspritztermin 1.1.	5 150 4 800	5 800 5 400	5 800 5 400	16 750 15 600
Einsatz:		12 600	15 200	16 200	16 050	19 950	20 600	21 250	121 850 S-M'Öl	25 150	25 800	25 800	76 750 S-M'Öl
Produkt:		11 700	14 100	15 100	14 900	18 550	19 150	19 750	111 250 B-M'Öl	23 400	24 000	24 000	71 400 B-M'Öl
Kammer 4 (oooo) 32 m³	Leistung 0,48 Ausbeute 85 % eingesetzt	10 000 8 500	10 000 8 500	10 000 8 500	10 000 8 500	10 000 8 500	10 000 8 500	10 000 8 500	70 000 59 500	12 000 10 200	12 000 10 200	12 000 10 200	36 000 30 600
Kammer 5 (oooo) 32 m³		1 700 1 450	4 100 3 500	5 100 4 350	4 900 4 150	8 550 7 300	9 150 7 750	9 750 8 300	43 250 36 800	11 400 9 700	12 000 10 200	12 000 10 200	35 400 30 100
Einsatz:		11 700	14 100	15 100	14 900	18 550	19 150	19 750	113 250 B-M'Öl	23 400	24 000	24 000	71 400 B-M'Öl
Produkt:		9 950	12 000	12 850	12 650	15 800	16 250	16 800	96 300 L-Bi	19 900	20 400	20 400	60 700 L-Bi
Kammer 1 (ooooo) 24 m³		3 600 3 000	3 600 3 000	3 600 3 000	3 600 3 000	3 600 3 000	3 600 3 000	3 600 3 000	25 200 21 000	3 600 3 000	3 600 3 000	3 600 3 000	10 800 9 000
Kammer 21 (ooooo) 24 m³	Leistung 0,25 bzw. 0,38 eingesetzt (bei 500 Std/Monat)				Einspritztermin 15.10.	900 750	6 300 5 250	7 200 6 000	14 400 12 000	7 200 6 000	7 200 6 000	7 200 6 000	21 600 18 000
Kammer 22 (ooooo) 24 m³	Ausbeute 83 % eingesetzt						Einspritztermin 15.12.	900 750	900 750	6 300 5 250	7 200 6 000	7 200 6 000	20 700 17 250
Einsatz:		3 600	3 600	3 600	3 600	4 500	9 900	11 700	40 500 L-Bi	17 100	18 000	18 000	53 100 L-Bi
Produkt:		3 000	3 000	3 000	3 000	3 750	8 250	9 750	33 750 DHD-Bi	14 250	15 000	15 000	44 250 DHD-Bi

[illegible]	Januar bis Mai	Juni	Juli	August	September	Oktober	November	Dezember	Summe Juni - Dezember 1941	Januar	Februar	März	Summe Jan.-März 1942		Summe [illegible]
DHD - Bi	eff. 19 540	3 000	3 000	3 000	3 000	3 750	8 250	9 750	33 750 DHD-Bi	14 250	15 000	15 000	44 250	DHD-Bi	[illegible]
CV_2-Bi		3 750	3 750	3 750 1 900	6 250 1 250	7 500	7 500	7 500	40 000 CV_2-Bi	7 500	7 500	7 500	22 500	CV_2-Bi	[illegible]
L-Bi	eff. [illegible]	[illegible]	[illegible]	9 250	[illegible]	11 700	[illegible]	[illegible]	[illegible] L-Bi	[illegible]	[illegible]	[illegible]	[illegible]	L-Bi	[illegible]
[illegible]	eff. [illegible]	13 100	15 [illegible]	17 900	19 550	22 950	[illegible]	[illegible]	[illegible] Fertigprodukte	[illegible]	[illegible]	[illegible]	[illegible]	Fertigprodukte	[illegible]
I.SZ	-	-	-	-	-	[illegible]	[illegible]	[illegible]	[illegible] I.SZ	[illegible]	[illegible]	[illegible]	[illegible]	I.SZ	[illegible]
Gesamtproduktion	eff. [illegible]	13 100	15 [illegible]	17 900	19 550	22 [illegible]	[illegible]	[illegible]	[illegible] Gesamtproduktion	[illegible]	[illegible]	[illegible]	[illegible]	Gesamtproduktion	[illegible]

HYDRIERWERKE PÖLITZ AKTIENGESELLSCHAFT

Stettin-Pölitz, den 17. Mai 1941 [illegible]

Produktionsprogramm Pölitz 1941 nach dem Stand vom 17.5.1941.

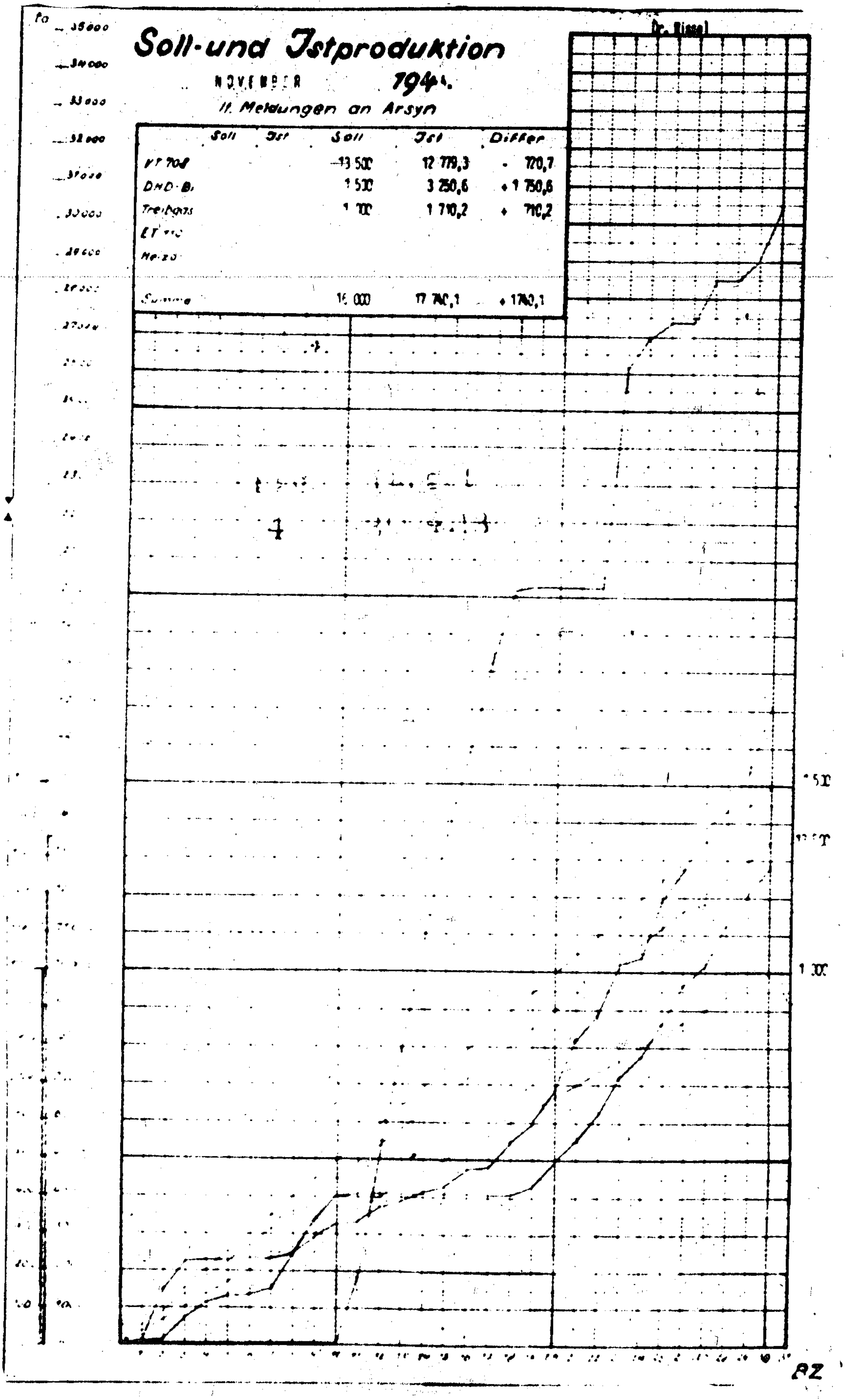
	Januar	Februar	März	April	Mai	Juni	Juli	August	September	Oktober	November	Dezember	Summe Mai - Dezember 19
DHD - Benzin	-	-	} eff. } 2 169	} eff. } 4 402	2 500	2 500	2 500	2 500	2 500	4 900	9 700	12 100	39 200 to
CL$_2$b - Benzin	-		}	}	3 000	3 000	3 000	3 000	5 000	6 000	6 000	6 000 ÷	35 000 to
L - Öl	eff. 6 674	eff. 6 738	eff. 7 633	eff. 15 183	4 900	6 950	9 550	12 150	12 210	11 750	6 600	3 000	67 100 to
Summe	eff. 6 674	eff. 6 738	eff. 9 802	eff. 19 585	10 400	12 450	15 050	17 650	19 710	22 650	22 300	21 100	141 300 to
T$_{52}$	-	-	-	-	-	-	-	-	-	300	600	1 000	1 900 to
	eff. 6 674	eff. 6 738	eff. 9 802	eff. 19 585	10 400	12 450	15 050	17 650	19 710	22 950	22 900	22 100	143 200 to

Dr. Rissel

Soll-und Istproduktion

NOVEMBER 1944.

II. Meldungen an Arsyn



	Soll	Ist	Soll	Ist	Differ
VT 708			13 500	12 779,3	- 720,7
DHD-B.			1 500	3 250,6	+ 1 750,6
Treibgas			1 000	1 710,2	+ 710,2
ET [illegible]					
Heizö					
Summe			16 000	17 740,1	+ 1740,1

BZ

[illegible] PÖLITZ AKTIENGESELLSCHAFT

Monat: Dezember 1943 [illegible] AB

Tägliche Meldungen an Arbeitsgemeinschaft [illegible]

Dr.[illegible]

Dat.	Flugbenzin (to)					DHD-Benzin (Ad)					ET 110 (Fr)					Treibgas					Heizöl (Dz)				
	Produktion to		Versand to		Bestand	Produktion to		Versand to		Bestand	Produktion to		Versand to		Bestand	Produktion to		Versand to		Bestand	Produktion to		Versand to		Bestand
	tägl.	Summe	tägl.	Summe	täglich	tägl.	Summe	tägl.	Summe	täglich	tägl.	Summe	tägl.	Summe	täglich	tägl.	Summe	tägl.	Summe	täglich	tägl.	Summe	tägl.	Summe	täglich
1.	[illegible]	[illegible]	585	585	2824	211	211	1347	1347	1113	44	44	-	-	708	127	127	236	236	230	-	-	403	403	1696
2.	[illegible]	[illegible]	1415	2000	2230	110	321	-	1347	1223	38	82	-	-	1456	320	447	199	435	351	-	-	-	403	1696
3.	[illegible]	1878	1315	3315	1287	489	810	-	1347	1712	41	123	-	-	788	111	558	166	601	298	-	-	-	403	1696
4.	[illegible]	2987	-	3315	2381	772	1582	-	1347	2484	49	172	-	-	837	274	832	179	780	399	-	-	443	846	1253
5.	[illegible]	[illegible]	1323	4638	2048	592	2174	711	2058	2365	44	216	-	-	881	187	1019	205	985	373	-	-	-	846	1253
6.	[illegible]	4691	-	4638	2782	522	2696	692	2750	2195	46	262	-	-	927	269	1288	250	1235	392	-	-	-	846	1253
7.	[illegible]	5336	1352	5990	2075	449	3145	-	2750	2644	42	304	-	-	969	164	1452	236	1471	320	-	-	366	1212	887
8.	[illegible]	[illegible]	-	5990	2864	408	3553	-	2750	3052	43	347	-	-	1012	190	1642	247	1718	263	-	-	-	1212	887
9.	[illegible]	[illegible]	1650	7640	[illegible]	544	4097	-	2750	3596	36	383	-	-	1048	199	1841	199	1917	263	-	-	-	1212	887
10.	[illegible]	[illegible]	-	7640	2880	359	4456	1941	4691	2014	49	432	-	-	1097	168	2009	180	2097	251	838	838	-	1212	1725
11.	[illegible]	[illegible]	1251	8891	2219	555	5011	541	5232	2028	38	470	-	-	1135	194	2203	123	2220	322	-	838	469	1681	1256
12.	[illegible]	9118	977	9868	1980	461	5472	-	5232	2489	52	522	-	-	1187	149	2352	151	2371	320	-	838	-	1681	1256
13.	742	[illegible]	346	10214	2376	389	5861	1263	6495	1615	46	568	-	-	1233	345	2697	70	2441	595	1305	2143	-	1681	2561
14.	[illegible]	10579	1353	11567	1841	470	6331	572	7067	1513	41	609	-	-	1274	110	2807	110	2551	456	-	2143	-	1681	2561
15.	778	[illegible]	-	11567	2620	498	6829	-	7067	2011	39	648	961	961	352	176	2983	164	2715	468	-	2143	-	1681	2561
16.	[illegible]	[illegible]	1300	12867	2211	340	7169	-	7067	2351	43	691	-	961	395	170	3153	170	2885	481	-	2143	359	2040	2202
17.	[illegible]	13301	541	13408	2652	384	7553	1349	8416	1386	44	735	-	961	439	109	3262	152	3037	438	-	2143	-	2040	2202
18.	627	[illegible]	733	14141	2746	459	8012	-	9416	1845	41	776	-	961	480	225	3487	118	3155	545	-	2143	-	2040	2202
19.	[illegible]	15011	1148	15289	2451	417	8429	-	9416	2262	41	817	-	961	521	93	3580	245	3400	393	-	2143	-	2040	2202
20.	[illegible]	[illegible]	732	16021	2822	507	8936	395	8811	2376	43	860	-	961	564	136	3716	193	3593	336	-	2143	488	2528	1714
21.	[illegible]	[illegible]	-	16 021	3520	195	9131	1050	9861	1519	44	904	-	961	608	141	3857	162	3755	315	-	2143	494	3022	1220
22.	[illegible]	[illegible]	1429	17450	2883	510	9641	-	9861	2029	44	948	-	961	652	185	4042	140	3895	360	-	2143	412	3434	808
23.	[illegible]	[illegible]	1334	18784	2287	524	10165	656	10517	1897	41	989	-	961	693	137	4179	140	4035	357	-	2143	-	3434	808
24.	[illegible]	[illegible]	[illegible]	19418	2270	501	10666	496	11013	1902	41	1030	-	961	734	171	4350	138	4173	390	-	2143	543	3977	265
25.	[illegible]	[illegible]	-	19418	3196	149	10815	-	11013	2051	44	1074	-	961	778	166	4516	6	4179	550	-	2143	-	3977	265
26.	[illegible]	[illegible]	1191	20609	2928	-	10815	693	11706	1358	40	1114	-	961	818	207	4723	228	4407	529	130	2273	395	4372	-
27.	778	[illegible]	[illegible]	[illegible]	3807	527	11342	675	12381	1210	44	1158	-	961	862	141	4864	281	4688	389	-	2273	-	4372	-
28.	[illegible]	[illegible]	[illegible]	22778	2203	667	12009	-	12381	1877	38	1196	-	961	900	231	5095	148	4836	472	-	2273	-	4372	-
29.	[illegible]	[illegible]	[illegible]	24073	[illegible]	195	12204	-	12381	2072	44	1240	-	961	944	204	5299	258	5094	418	449	2722	449	4821	-
30.	[illegible]	24 [illegible]	-	24073	2786	607	12811	1105	13486	1020	44	1284	-	961	988	225	5524	209	5303	434	470	3192	470	5291	-
31.	[illegible]	24 814	1334	25407	2136	684	13495	304	13790	1854	38	1322	-	961	1026	142	5666	123	5552	453	505	3697	505	5796	
effektiv		24 814,0					13 494,6					1 321,8					5665,7					3697,7			

Produktion im Dezember 1943	
VT 7[illegible] - Bi	24 814,0 to
DHD - Bi	13 494,6 "
ET 110	1 321,8 "
Treibgas	5 665,7 "
Heizöl	3 697,7 "
effektiv	48 993,8 to

Bel

Dr. Riesel

																								Versand t. tägl.	Versand t. Summe	
1.	1 356	1 356	1 367	1 362	3 124	213	213	723	723	1 963	65	65	-	-	1 070	203	203	329	329	1 145	-	-	-	-	-
2.	1 311	2 667	1 248	2 610	3 287	320	533	642	1 365	1 541	41	106	-	-	1 111	341	544	337	666	1 149	-	-	-	-	-
3.	1 392	4 059	599	3 209	3 980	453	953	-	1 365	1 391	46	152	-	-	1 157	260	804	317	983	1 092	-	-	-	-	-
4.	1 219	5 278	985	4 194	4 214	388	1 348	-	1 365	2 356	50	202	446	446	761	387	1 191	378	1 361	1 101	-	-	-	-	-
5.	1 234	6 512	-	4 194	5 448	492	1 840	741	2 106	2 107	34	236	443	889	352	242	1 433	288	1 649	1 055	-	-	-	-	-
6.	1 603	8 115	1 990	6 187	5 058	135	1 975	634	2 740	1 608	41	277	-	889	393	274	1 707	237	1 886	1 092	-	-	-	-	-
7.	1 782	9 897	2 041	8 228	4 799	-	1 975	-	2 740	1 608	41	318	-	889	434	298	2 005	139	2 025	1 251	-	-	-	-	-
8.	1 433	11 330	2 492	10 720	3 740	364	2 339	-	2 740	1 972	51	369	-	889	485	152	2 157	277	2 302	1 126	-	-	-	-	-
9.	1 259	12 589	1 721	11 441	3 278	672	3 011	-	2 740	2 644	55	424	-	889	540	416	2 573	322	2 624	1 220	714	714	714	714	-
10.	1 066	13 655	956	12 397	3 388	556	3 567	-	2 740	3 200	57	481	-	889	597	272	2 845	326	2 950	1 166	-	714	-	714	-
11.	1 345	14 998	547	12 944	4 186	460	4 027	1 321	4 061	2 339	60	541	-	889	657	298	3 143	364	3 314	1 100	-	714	-	714	-
12.	1 523	16 521	2 289	16 233	3 420	393	4 420	661	4 722	2 071	55	596	-	889	712	300	3 443	259	3 573	1 141	-	714	-	714	-
13.	932	17 455	1 199	17 432	3 153	201	4 621	-	4 722	2 272	33	629	-	889	745	157	3 600	294	3 867	1 004	-	714	-	714	-
14.	1 243	18 698	1 140	18 572	3 256	549	5 170	709	5 431	2 112	54	683	-	889	799	300	3 900	222	4 089	1 082	-	714	-	714	-
15.	1 344	20 042	1 946	20 518	2 654	428	5 598	663	6 094	1 877	63	746	-	889	862	420	4 320	303	4 392	1 199	-	714	-	714	-
16.	1 224	21 266	1 216	21 734	2 662	439	6 037	-	6 094	2 316	52	798	-	889	914	366	4 686	461	4 853	1 104	631	1 345	631	1 345	-
17.	1 270	22 536	1 085	22 819	2 847	443	6 480	-	6 094	2 759	60	858	-	889	974	382	5 068	470	5 323	986	640	1 985	640	1 985	-
18.	1 164	23 700	1 170	23 989	2 841	346	6 826	-	6 094	3 105	71	929	-	889	1 045	353	5 421	375	5 698	994	-	1 985	-	1 985	-
19.	1 114	24 814	1 621	25 610	2 334	178	7 004	-	6 094	3 283	37	966	757	1 646	325	153	5 574	410	6 108	737	523	2 508	523	2 508	-
20.	1 399	26 213	1 037	26 647	2 696	430	7 434	-	6 094	3 713	60	1 026	-	1 646	385	409	5 983	312	6 420	834	-	2 508	-	2 508	-
21.	1 305	27 518	1 213	27 860	2 788	381	7 815	-	6 094	4 094	60	1 086	-	1 646	445	281	6 264	359	6 779	756	-	2 508	-	2 508	-
22.	1 646	29 164	962	28 822	3 472	426	8 241	-	6 094	4 520	60	1 146	-	1 646	505	247	6 511	295	7 074	708	-	2 508	-	2 508	-
23.	1 482	30 646	1 423	30 245	3 531	376	8 617	1 446	7 540	3 450	66	1 212	-	1 646	571	364	6 875	432	7 506	640	1 056	3 564	1 056	3 564	-
24.	1 425	32 071	1 283	31 528	3 673	322	8 939	549	8 089	3 223	84	1 296	-	1 646	655	280	7 155	249	7 755	671	878	4 442	878	4 442	-
25.	1 414	33 485	1 401	32 929	3 686	248	9 187	1 470	9 559	2 001	71	1 367	-	1 646	726	390	7 545	348	8 103	713	-	4 442	-	4 442	-
26.	1 267	34 752	1 254	34 183	3 699	254	9 441	909	10 468	1 346	79	1 446	-	1 646	805	304	7 849	279	8 382	738	-	4 442	-	4 442	-
27.	1 537	36 289	2 198	36 381	3 038	330	9 771	-	10 468	1 676	66	1 512	-	1 646	871	328	8 177	324	8 706	742	-	4 442	-	4 442	-
28.	1 430	37 719	1 461	37 842	3 007	319	10 090	-	10 468	1 995	68	1 580	-	1 646	939	300	8 477	258	8 964	784	337	4 779	-	4 442	337
29.	414	38 133	-	37 842	3 412	122	10 212	-	10 468	2 117	-	1 580	-	1 646	939	53	8 530	137	9 101	700	553	5 332	-	4 442	890
30.	-	38 133	-	37 842	3412	-	10 212	-	10 468	2 117	-	1 580	-	1 646	939	-	8 530	-	9 101	700	-	5 332	-	4 442	890
31.	-	38 133	-	37 842	3 412	-	10 212	-	10 468	2 117	-	1 580	-	1 646	939	-	8 530	-	9 101	700	-	5 332	-	4 442	890

VT 708	38 133,1
DHD-81	10 212,1
ET 110	1 580,5
Treibgas	8 530,4
Heizöl	5 332,2
	63 788,3

HYDRIERWERKE PÖLITZ AKTIENGESELLSCHAFT

Monat: Juni [illegible]

Tägliche Meldungen an Arbeitsgemeinschaft Hydrierung, Synthese und Schwelung, Berlin.

Dr. [illegible]

Dat.	Flugbenzin (Av)					OHD-Benzin (Ad)					ET 110 (Pr)					Treibgas (Ge)					Heizöl (Hz)				
	Produktion to tägl.	Produktion to Summe	Versand to tägl.	Versand to Summe	Bestand täglich	Produktion to tägl.	Produktion to Summe	Versand to tägl.	Versand to Summe	Bestand täglich	Produktion to tägl.	Produktion to Summe	Versand to tägl.	Versand to Summe	Bestand täglich	Produktion to tägl.	Produktion to Summe	Versand to tägl.	Versand to Summe	Bestand täglich	Produktion to tägl.	Produktion to Summe	Versand to tägl.	Versand to Summe	Bestand täglich
1.	-	-	-	-	3 412	-	-	-	-	2 117	-	-	-	-	939	-	-	-	-	700	-	-	288	288	687
2.	-	-	-	-	.	-	-	-	-	.	-	-	212	212	727	-	-	2	2	698	-	-	474	[illegible]	213
3.	-	-	-	-	.	-	-	-	-	.	-	-	495	707	232	-	-	9	11	689	-	-	213	[illegible]	-
4.	-	-	-	-	.	-	-	-	-	.	-	-	-	.	.	-	-	4	15	685	-	-	-	.	-
5.	-	-	-	-	.	-	-	-	-	.	-	-	-	.	.	-	-	12	27	673	-	-	-	.	-
6.	-	-	-	-	.	-	-	-	-	.	-	-	-	.	.	-	-	16	43	657	-	-	-	.	-
7.	-	-	654	654	2 758	-	-	-	-	.	-	-	-	.	.	-	-	17	60	640	-	-	-	.	-
8.	-	-	708	1 362	2 050	-	-	-	-	.	-	-	-	.	.	-	-	29	89	611	-	-	-	.	-
9.	-	-	-	.	.	-	-	-	-	.	-	-	-	.	.	-	-	20	109	591	-	-	-	.	-
10.	-	-	-	.	.	-	-	976	976	1 141	-	-	-	.	.	-	-	10	119	581	-	-	-	.	-
11.	-	-	713	2 075	1 337	-	-	383	1 359	758	-	-	-	.	.	-	-	1	120	580	-	-	-	.	-
12.	-	-	908	2 983	429	-	-	-	.	.	-	-	-	.	.	-	-	12	132	568	-	-	-	.	-
13.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	6	138	562	-	-	-	.	-
14.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	6	144	556	-	-	-	.	-
15.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	4	148	552	-	-	-	.	-
16.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	1	149	551	-	-	-	.	-
17.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	12	161	539	-	-	-	.	-
18.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	2	163	537	-	-	-	.	-
19.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	-
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23.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	-
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27.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	-
28.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	-
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30.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	.	-	-	-	.	-
	28,9 von OHD							28,9 an VT 70								103,6 von Gelsenberg Benzin A.G.									
				2 983,4	406,7				1 359,5	728,5		-	705,8		233,3				163,9			-		988,2	-
Verlust				+ 25,0					+ 28,9				(0,4 Abfüllverlust)				f. Katastr.eins.		+ 17,7						
f. Katastropheneins.				+ 28,1																					

HYDRIERWERKE PÖLITZ AKTIENGESELLSCHAFT

Monat: Juli

Tägliche Meldungen an Arbeitsgemeinschaft Hydrierung, Synthese und Schwelung, Berlin.

Dr. Wesel

Dat.	Flugbenzin (Ak)					OHD-Benzin (Ad)					ET 110 (Pr)					Treibgas (Go)					Heizöl (Oz)				
	Produktion to		Versand to		Bestand	Produktion to		Versand to		Bestand	Produktion to		Versand to		Bestand	Produktion to		Versand to		Bestand	Produktion to		Versand to		Bestand
	tägl.	Summe	tägl.	Summe	täglich	tägl.	Summe	tägl.	Summe	täglich	tägl.	Summe	tägl.	Summe	täglich	tägl.	Summe	tägl.	Summe	täglich	tägl.	Summe	tägl.	Summe	täglich
1.	-	-	-	-	487	-	-	-	-	729	-	-	-	-	223	-	-	-	-		-	-	-	-	-
2.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
3.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
4.	-	-	-	-	-	-	-	651	651	78	-	-	-	-	-	-	-	-	-		-	-	-	-	-
5.	78	[illegible]	-	-	565	-	-	78	729	-	-	-	-	-	-	-	-	10	10		-	-	-	-	-
6.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
7.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
8.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	2	12		-	-	-	-	-
9.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
10.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
11.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
12.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
13.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
14.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
15.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
16.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
17.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
18.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
19.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	1	13		-	-	-	-	-
20.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
21.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
22.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
23.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
24.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
25.	-	-	-	-	-	929	929	-	-	929	-	-	-	-	-	-	-	1	14		-	-	-	-	-
26.	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		-	-	-	-	-
27.	-	-	-	-	-	493	1 422	708	1 437	714	-	-	-	-	-	-	-	-	-		-	-	-	-	-
28.	-	-	-	-	-	757	2 179	-	-	1 471	-	-	-	-	-	-	-	2	16		-	-	-	-	-
29.	-	-	-	-	-	33	2 212	402	1 839	1 102	-	-	-	-	-	-	-	-	-		-	-	-	-	-
30.	-	-	-	-	-	73	2 285	846	2 185	439	-	-	-	-	-	-	-	-	-		-	-	-	-	-
31.	-	-	-	-	-	474	2 759	-	-	913	-	-	-	-	-	-	-	1	17		-	-	-	-	-