CHAPTER VI.

Questions raised during the study of the process by Partners with answers by Dr Alberts.

November 1st 1938.

Several questions were discussed, which Dr Alberts answered as follows:

1. Throughput of ideal gas per ton Co per hour in first stage ?

This depends on the amount of inert gas.

a) At atmospheric pressure: 1300 m3 per hour total gas, including inerts. This is an average and remains the same whether more or less inerts are present. Production: 100 g liquid product + 12 g gasol per m3 ideal gas.

b) At higher pressures: the pressure can probably vary between 5 and 10 atm. At higher pressures the trend goes towards more paraffins and more saturated hydrocarbons. Therefore, at 5 atm. less paraffins will be produced and the power consumption will also be less than at 10 atm. From 10 - 15 atm. neither yield nor paraffin content vary (this means a variation of not more than 2 %). Above 15 atm. both yield and paraffin content drop appreciably.

Difference in yield and paraffin content between 7 and 10 atm. is very small.

At higher pressures more oxygen-containing products are formed and at still higher pressure the reaction goes more and more towards the methanol synthesis.

Note: Numbers omitted relate to questions which have been dealt with in other chapters.

Throughput in case of pressure synthesis is about 20% higher that at atmospheric pressure. Yield: 125-130 g liquid product + 7 g gasol per m3 ideal gas.

Note.

Figures at atmospheric pressure are actual plant data over long periods. Figures at higher pressures are data mostly from laboratory and pilot plant tests.

2. Throughput per ton Co per hour in second stage ?

Both at atmospheric and higher pressures the throughput in the second stage is about 10% more synthesis gas (not ideal gas) than in the first stage.

Yield at atmospheric pressure: 20 g liquid product + 3 g gasol per m3 ideal gas entering first stage.

Yield under pressure: 25 to 30 g liquid product + 3 g gasol per m3 ideal gas entering first stage.

Therefore, total yield:
atmospheric pressure: 100 + 20 = 120 g liquid product and 12 + 3 = 15 g gasol per m3 ideal gas;
under pressure: 125 to 130 + 25 to 30 = 150 to 160 g liquid product and 7 + 3 = 10 g gasol per m3 ideal gas.

3-4. With and without pressure 90 - 92% of the CO is being converted. Difference in yield between the two processes has to be explained by a larger or smaller production of methane.

Theoretical yield from 1 m3 ideal gas is 210 g per m3.

Holten use gas with about 15 - 17% inerts. The contraction in the first stage is about 60 - 65%, for 70 - 75% conversion of CO, whereas the contraction in the second stage is about 30 - 40% on the gas entering the second stage.

5. Total production in first stage ?

Atmospheric pressure: Catalyst life in both stages is about 4 months or 122 days. Throughput at Holten is about 1300 m3 total gas per ton Co per hour or about 1100 m3 ideal gas per ton Co per hour. Therefore, total production in first stage is 122 x 24 x 1100 x 100 = 322 tons liquid product per ton Co.

Pressure: Catalyst life 6 months, or 183 days. Total production, therefore, in first stage : 183 x 24 x 1100 x 125 x $\frac{12}{10}$ (20% more throughput) = 720 tons liquid product per ton Co.

6. Total production in second stage ?

This depends on the number of first-stage ovens connected to one second-stage oven. During its whole lifetime a catalyst remains in the stage in which it has been put originally. Generally 3 first-stage ovens are connected to 1 second-stage oven. Supposing that 1000 m3 total synthesis gas, containing 17% inerts, enters each of the first-stage ovens, that is, 830 m3 ideal gas (CO + 2 H_2), at a conversion of 76% on CO in the first stage 630 m3 ideal gas will thus be converted. The remaing gas per oven will then consist of 200 m3 CO + 2 H_2 and 170 m3 inerts + some methane formed in the first stage; in all, therefore, about 400 m3 per oven. The feed to a second-stage oven connected to 3 first-stage ovens will, therefore, be about 1200 m3 total gas containing about 50% inerts.

Production in the second stage is said to be about 20% of the production in the first stage. Three ovens in the first stage will produce : 3000 x 0.83 (total ideal gas) x 100 = 250 kg liquid product per hour. 20% of this quantity, i.e. 50 kg, will be produced in the second stage. Assuming that the gas entering the second stage contains 50% inerts, making from 1200 m3 total gas 600 m3 ideal gas per hour, then the production is about 85 g per m3 ideal gas in the second stage. Actually the production in the second stage amounts to 90 - 100 g of liquid product per m3 ideal gas.

The conclusion which can be drawn from the above is that irrespective of stage, the conversion per converter amounts to about 100 g liquid product per m3 ideal gas.

Further information on the two stages : Poisonous resins deposited on spent catalyst appear to be the same in both stages.

It has been found that the percentage of Co in the catalyst could be decreased, without affecting the production per converter. Originally one oven filled with 3 tons of catalyst contained about 1400 kg Co; later on this was reduced to 850 - 900 kg Co. This refers to atmospheric pressure ovens.

When starting the pressure experiments the latter catalyst was used. This is also the catalyst now being used in the few commercial pressure ovens which are at present in operation under atmospheric pressure. Experimentally, the amount of Co has even been reduced in pressure experiments to an equivalent of 400 kg Co per converter filling.

Every 4 months an oven is taken out of service. It takes about 2 days in all to discharge the oven, refill it again with catalyst and put it into operation.

On an average, 2.5 times during the 4 months' running periods a catalyst is regenerated in situ. Usually the first time this is done is after about 1200 operating hours. Total time for regeneration: 10 - 12 hours. Therefore, total regenerating time per 4 months' running period is 2.5 x 12 = 30 hours. The overall out-of-production time per oven (regenerating, discharging and refilling) is, therefore, 48 + 30, or, say, 80 hours, or per year 240 hours = 10 days. This is about 97% service factor. For the design of the present German plants a service factor of 92 - 93% has been taken.

For design purposes an overall throughput of 820 - 850 m3 ideal gas per converter (1st + 2nd stage) is taken.

The second-stage ovens do not have to be regenerated, or, at the most, only once during a 4 months' period, because less paraffin is absorbed by the catalyst in this stage, owing to, firstly, less paraffin being produced and, secondly, the scavenging action of the larger amount of inert gases present.

Thus far it has not been necessary to regenerate the catalyst in the pilot plant working under higher pressures. These catalysts have now been 6 months in operation and have, therefore, shown a 100% time efficiency.

7.&9. Number of ovens operating in first and second stage ?

There are 52 atmospheric pressure ovens at Holten. Usually 48 - 50 of these ovens are in operation, about 36 being in the first stage and about 13 - 14 in the second stage.

11. Catalyst volume per oven ?

Atmospheric pressure: total free volume 12.5 m3; catalyst volume: 10 - 11 m3. Pressure ovens: free volume 10 m3; catalyst volume: 10 m3.

This shows that the overall apparent density of the catalyst is a little higher in the pressure ovens than in the atmospheric pressure ovens.

12. Tons of Co per oven ?

850 - 900 kg Co per oven for both atmospheric and higher pressures.

13. Apparent density of catalyst ?

280 - 300 kg/m3.

14. Number of spare ovens ?

At Holten 3 - 4 out of 52 converters

15. Weight % Co in catalyst ?

In green contact (that is unreduced catalyst) roughly 20%. This latter catalyst still contains about 7 - 10% water and also a little absorbed CO_2. In final catalyst 24 - 26%, viz. 850 - 900 kg Co per 3 - 3.1 tons oven charge.

19. Exact readings from thermocouples in catalyst bed ?

The temperature difference between lower and upper layer of the catalyst in an oven is about 3 - 5°C. Temperature difference, when working with a tube of about 20 mm dia. in the laboratory between the wall and the centre of the catalyst (therefore, about 10 mm distance), is about 15°C.

Practice has shown that the temperature control of the commercial ovens can be done by the steam pressure only. Even in the laboratory no temperatures are taken any longer with a pyrometer tube in the catalyst bed. However, it is realized that the actual temperature on the catalyst surface may be from 100 to a few 100°C higher than the temperature actually measured.

<u>21.</u> Are ovens started singularly or in block ?

All converters in one block are started and shut down together.

<u>22.</u> What is the preferred number of ovens operating in one block from a process point of view and from a plant-size point of view ?

Oven blocks with 2, 4 and 6 ovens are in use at Holten. Dr Alberts does not think that there is any <u>objection from a process point of view to combining 6 ovens</u> in one block. For small plants, however, it will be advisable to reduce this number, as otherwise the service factor will be adversely affected.

<u>23.</u> What is the total amount of fin cooling surface and steam-pipe cooling surface in both low and high pressure ovens ?

Low pressure ovens: 4000 m2 total cooling surface . per oven. 10% of this is water-cooled tube surface

High pressure ovens: 2100 m2 total cooling surfa'
all being water-cooled surface.

<u>25.</u> Weight of metal per oven exclusive of boiler drum and connections ?

Low pressure ovens : 41 kgt.

High pressure ovens : 48 - 49 kgt, all without insulation.

<u>26.</u> Why are ovens used with a depth of catalyst of 2.5 m ?

Pilot-plant converters have a depth of 1.5 m, commercial-plant pressure converters of 4.5 m. ~~Do not~~ know how far one can go, but too high a catalyst bed would give too much gas on the first layer of catalyst where the gas enters, which could cause trouble through too much heat evolution. Permissible catalyst depth will, therefore, depend on cooling surface and efficiency of cooling means. For this reason Ruhrchemie do not think that the depth of their low pressure converters could be much increased. On one occasion, a fairly large amount of carbon was found to have deposited on the entrance side of the catalyst chamber and warping of the vane tops was noticed. This ~~was not due to too much gas, but was explained at the~~ time by too high an initial activity of the catalyst.

28. Pressure drop per foot of catalyst versus age of catalyst in first- and second-stage ovens ?

Pressure drop does not increase with age of catalyst. Normally the pressure drop is 100 - 200 mm water per oven, both for atmospheric and high pressure ovens, such when using 2 - 3 mm catalyst. It is evident that variation in the size of catalyst will vary the pressure drop.

29. What experiments have been made to prove that oven has been evenly filled ?

Once the pressure drop of all tubes on two cross lines in a pressure converter was measured. The highest variation found was 10%.

31. Have you ever operated high pressure ovens at atmospheric pressure ?

High pressure ovens were operated at atmospheric pressure with same throughput as for the atmospheric pressure ovens. No difference in operation and yield was found. No tests were made in pressure ovens at atmospheric pressure with higher throughputs, but Dr Alberts believes that this will be possible. This question is of importance because the cooling is supposed to be much better in the pressure ovens than in the atmospheric pressure ovens.

32. Why did you choose 15 ft depth of catalyst bed for high pressure ovens instead of using 3 ovens in parallel with 5 ft depth of catalyst bed ?

This was purely a matter of capital costs.

~~In the pilot plant it was found that with equal~~ space velocity no difference could be determined in composition of the products when varying the linear velocity from 1 to 2 or 1 to 4.

However, different products are obtained when using

a) 1000 m3/h containing 800 m3 CO + 2 H_2

versus

b) 2000 m3/h containing 800 m3 CO + 2 H_2.

Lighter products are obtained when the gas contains more inerts.

In general it can be said that at the same temperature but with higher throughputs the products obtained are lighter.

With same throughput but varying temperatures lighter products are obtained at higher temperatures.

33. Have you ever emptied an oven layer by layer and examined the catalyst from each layer in the laboratory to determine whether the decline in activity of the catalyst is the same at top and bottom ?

Activity decreases from top to bottom. A comparatively fresh catalyst will show much higher activity in the bottom layers than in the top layers, the latter ones being poisoned, of course, first. After some time of running, however, the bottom layers will get poisoned also, so that the activity of the catalyst at top and bottom will then be about equal.

34. What is your present recommendation for ideal construction of low pressure ovens ?

Converters with catalyst in tubes.

When not considering yield per m3 gas but only yield per g Co, one can use wider tubes when increasing the linear velocity. Dr Alberts believes that when circulating gas, such as, for instance, tail gas or CH_4, to remove the reaction heat it will be necessary to recirculate about 5 or 6 times the volume of fresh feed.

When using the same gas quantity the conversion with a 2 - 3 mm catalyst will be about 70% CO in the first stage, whilst when using a 0.5 - 1.0 mm catalyst the conversion will be about 90% CO.

One can reach also in the latter case a conversion of 70 - 75% CO by increasing the throughput by 35%. This is due to better heat dissipation.

For commercial ovens Dr Alberts would recommend the following:

Tubes of 10 mm dia., thickness of wall: 0.75 mm or as thin as possible. Material of tubes, if possible, Al; length of tubes: 4 - 5 m. Catalyst: hard as stone if possible, less than 1 mm.

However, it is pointed out that a small quantity of dust is worst for a 0.5 - 1 mm than for a 2 - 3 mm catalyst. The shorter the tubes, the smaller the pressure drop and the more difficult it is to obtain an even gas distribution over the tubes. However, apart from gas distribution 500 tubes of 1 m are to be preferred to 100 tubes of 5 m.

35. Have you ever visualized connecting a number of cooling tubes to different headers in order to control temperature in direction of flow ?

Trials have been made in connecting a number of cooling tubes to different headers, in order to control temperature in direction of flow. No great difference was found in the results. A real advantage could not be ascertained; probably the products are somewhat more saturated.

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36. Have you ever made experiments with upflow through catalyst bed tending to show that you can operate with 1/2 or 1/4 of the normal catalyst bed thickness ?

Experiments with upflow through catalyst bed have been made. It was stated that in this case no great difficulties had been encountered with clogging up of the bed. However, after a fortnight the experiments with a low pressure oven had to be terminated on account of leakage. A h.p. pilot tube clogged up after 2 to 3 months. Probably the upflow will produce some more paraffins, but with slight difference as compared with downflow. Dr Alberts does not think much of the cooling effect of the reflux.

37. Analysis of synthesis gas entering first stage, leaving first stage and leaving the second stage ?

The analyses of the gas entering and leaving first stage and leaving second stage are as follows:

	entering first stage	leaving first stage	leaving second stage
CO	28	17	6
H_2	55	32.5	10
CH_4	0.4	8	17.8
CO_2	13.5	35.5	55
N_2	3.1	6.5	10
Olefines	-	0.5	1.2

Conversion after 1st stage : 72 - 75% CO
Total conversion after 2nd stage : 92 - 95% CO

38. What are the properties of the gasol fraction after first stage and after second stage ?

The properties of the gasol fractions after firs and second stage do not change.

Some time ago activated charcoal was used after the first stage; this was discontinued 7 months ago, as there was no difference to be found.

39. Have you any results when operating with a synthesis gas as produced directly in the water gas generator ?

A synthesis gas, consisting of 1 CO - 1.5 H_2 has been used.

It appeared that with l.p. operation, with Co-Th as a catalyst, the catalyst life was shorter (about 4 weeks) and that the liquid products were lighter.

The olefine content in the light gasoline was increased from 40 to 60% with an octane number of 68.

It is expected that more suitable catalysts may be found for the purpose of operating with a synthesis gas as produced directly in a watergas generator.

An experiment with Co-Mg catalyst under pressure is in progress. Yield is the same as with synthesis gas 1 : 2. During 2 months catalyst activity did not increase. There is some difference in olefine content of products between running at 5 atm. and at 10 atm.

It is interesting to note that under 7 atm. pressure working with watergas, about the same products (as far as boiling range and olefine content are concerned) are obtained as at normal pressure when working with gas containing $CO:H_2$ in ratio 1 : 2.

As Ruhrchemie are less interested in paraffin production, they, therefore, intend running their ~~full-scale plant under pressure~~ with watergas if the experiment should prove successful. In the second stage, they will then add some H_2, in order to obtain a better CO conversion.

The pressure of units in the second stage, also in the case of watergas as raw material, leads to a lighter product.

In case it should prove advantageous to run the second at a higher pressure, this will be possible, because the h.p. compressors can deliver the gas up to 15 atm. on account of the tail gas of the first stage having a higher specific gravity than the synthesis gas.

41. Reactions involved in organic sulphur removal ?

The reactions involved in organic sulphur removal are as follows:

After primary purifying, the gas contains 12 - 20 g/100 m3 sulphur in the form of about 60% carbon disulphide (CS_2) and about 40% carbon-oxy-sulphide (COS).

These figures, of course, depend on quality of coke and the way the generators are run. These compounds are converted into H_2S with the aid of H_2 present, and are taken up by the sodium carbonate; the reaction is promoted by the iron oxide.

A small percentage of oxygen (0.1 - 0.2%) is necessary to oxidize the sulphides into sulphates; air may be used for this purpose.

In the beginning no oxygen was added, but in that case the H_2S "breaks through" more easily, and the saturation could not be carried on higher than 7%, whereas now with the addition of oxygen, the saturation may go on up to 10%.

It is important that the purifying mass should be very porous, in which respect lux mass is quite satisfactory. It was stated that any kind of porous iron hydroxide is suitable, such as, for instance, "Raseneisenerz". Al-content is immaterial. This purifying mass consists of 66% FeO and 34% Na_2CO_3. The temperature when purifying is about 200 - 300°.

Cyclic-bound sulphur will not be removed, but mercaptans are.

Cyclic sulphur compound is contained in gases when produced in generators operating at relatively low temperature, but will be cracked at 1100 - 1200°C; consequently, the Holten synthesis gas does not contain such compounds.

The price of lux mass was stated to be RM 14.50 per ton, containing 50% water; calcinated soda RM 100. per ton and the price of finished purifying mass was said to be RM 90.- to RM 100.- per ton.

Questions 45 and 46:

45. Why is the addition of oxygen necessary ?
46. If the answer to 45 is to produce sulphates why do you have to make sulphates ?

have also been answered by the above.

44. Is a reducing atmosphere necessary for sulphur removing ?

Investigations will be carried out to answer the question whether a reducing atmosphere is necessary for sulphur removing.

47. What is the maximum space velocity used successfully in sulphur removal towers ?

As regards the question of the maximum tolerable space velocity in organic sulphur-removing towers, the following answer is given:

Two types of towers are in use:

– 110 m³
– 1
~ 190

	set of two towers	
a) screen towers (trays)	cap. 20,000 m3 gas/h	cont. 50-60 m3 mass each tower
b) cylindrical towers	cap. 25,000 - 28,000 m3/h (probably more)	cont. 90-100 m3 mass each tower.

The towers are used in sets of 2, placed one after the other, for the sake of greatest possible safety and economy.

When the first tower is saturated up to 10-11%, the set is taken out of operation and, after re-filling, the second tower is placed first.

The specific gravity (app.dens.) of the mass is 0.8.

A high porosity is necessary, therefore, no balls or pellets can be used.

48. You have two designs of towers; which do you prefer ?

For handling the heavy bells of the H_2S removal, hoisting equipment is necessary and this crane is also available for the second purifying step, thus making the use of screen (tray) towers, which are slightly less expensive than the cylindrical towers, possible. One tray at Holten has a weight of about 30 tons.

49. Do you have both types of towers in operation at Holten ?

At Holten 5 screened towers and 5 of the other type are available, i.e. in all: 5 sets of 2 towers. In operation there is no difference.

50. Do you recommend a heat-exchanger or cooler between towers ?

A heat-exchanger between the towers is unnecessary, as about the same temperature in both towers is satisfactory.

51. What is the analytical procedure for organic sulphur determination ?

The sulphur content of the purified gases is determined by combustion. The gas is passed through an analysis apparatus at the rate of 100 L/h , yielding after 4 hours sufficient sulphur to be determined as Ba sulph. This procedure has been described in "Brennstoffchemie". Heating the gas at 1000° in a quartz tube filled with quartz chips with the hydrogen present and determining the H_2S as CdS, is a shorter method.

52. How many grams of sulphur are deposited on catalyst when towers are changed from second stage to first stage and towers are removed ?

Regarding the shifting of towers, the following is stated:

The life of one tower is 83 days.

If one tower is taken out for renewing of the mass, also the second tower goes out of operation.

A cooling of the mass with circulating synthesis gas or tail gas is applied; otherwise the hot mass would glow when coming into contact with air, which would de detrimental to the transport belts and further equipment.

Part of the removed mass, which apparently has not yet been completely saturated, is separated, granulated and used again in the H_2S removal. The cooling, discharging and refilling of a tower takes 24 hours.

~~Dr~~ Alberts mentioned the possibility that in future or in other installations both purifying steps may be replaced by other systems, e.g. a Linde-Fränkl process, in which the gas is purified from CO_2,H_2S, organic sulphur and gum by regenerative cooling down to -180° / -190°. In this process a scavenging gas is applied to remove the condensed impurities, for which Linde require nitrogen, involving high costs. Dr Alberts thought it possible to use air for this purpose, but Linde refuse to build such a plant.

53. What is the total utility consumption and the net utility consumption, exclusive of the watergas generators and synthesis product-treating units (steam, electrical power, water and fuel gas) ?

The utility consumptions are given as follows:

1) gas consumption for organic sulphur removal 4% of calor. value of the gas to be treated; the flow chart ZM 106/19 estimates this figure to be 116 cal. per m3 synthesis gas.

2) for distilling fuel consumption 2%.

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3) when working under pressure, the heating of the fat oil from the oil absorption plant will require 1 - 1½% of the calor. value of the products condensed.

4) steam consumption for the final purifying: nil.

5) steam consumption of the synthesis: merely for starting up the processes (heating of the ovens), this quantity is about 1 ton/h steam of 20 atm. for a plant of 100 converters.

6) for the condensation of the heavy products in a plant working at atmospheric pressure, no steam is necessary. The activated charcoal plant requires 2 - 2.5 kg low-pressure steam of 2½ atm./kg gasoline produced.

If C_3 and C_4 are produced in large quantities, the steam consumption is to be set at 4½ kg/kg C_3 and C_4.

An average figure for Holten is about 3 kg/kg light benzine, incl. C_3 and C_4.

The steam produced in the synthesis at Holten amounts to 5 kg per kg liquid products.

7) When working under pressure no steam is used for condensing heavy benzine, which needs only indirect water cooling; the light parts are absorbed in oil; for removing the vapours from the fat oil, steam cannot be used, it having an insufficient temperature; the fuel consumption is about $1\frac{1}{2}$% of the cal. value of the oil produced.

8) The electrical energy consumption, assuming that no steam is used, may be stated as follows:

For H_2S removal, a small quantity for circulating air during the emptying of the vessels.

For organic S-removal about 20 kW for each set ~~of purifiers with a capacity of 20,000 m3/h . This~~ figure includes the blowers for air and circulating flue gas.

The electrical consumption for the synthesis is about 0.75 kW per oven for feed pumps, hydrogen blowers (intermittent), crane, etc.

9) For circulating cooling water pumps, 30 kW per ton total liquid products are required, this figure including also products pumps for synthesis at atmospheric pressure.

10) If the activated charcoal plant has to recover a large percentage of gasol (80-90%), a circulating blower is used with a consumption of 25 - 30 kW/ton total liquid products.

11) For the compression of propane-butane produced as gas (containing about 60% C_3C_4) from the activated charcoal plant, the electrical consumption is 150 kWh per ton C_3C_4.

12) When working at atmospheric pressure, the gas

transport blowers working at a pressure of 3000 mm have a power consumption of 15 kW for 1000 m3 gas.

When working at 10 atm., the turbo compressors consume 128 kWh/1000 m3, thus for 80,000 m3/hour, 10,000 kW !

13) The cooling water is circulated; for losses and evaporation a quantity of 50 m3 per ton of liquid primary products has to be supplied, including boiler feed water. This figure relates to the synthesis alone; for the complete process, incl. cracking and distillation, this figure would be about 70 m3.

~~The circulation at Holten amounts to about 500 m3/~~ hour.

56. Weight of steel required for each step ?

The weights of steel required for each step are given as follows:

H_2S removal	1000	tons per 40,000 m3 gas/h
organic S removal	475	" " " " "
synthesis: one l.p. converter	45	"
one h.p. converter	43	"
pipe connections, fittings, boiler drum, etc. about	8	" " converter
synthesis bldg, supports & roof, incl. crane, excl. walls	3-4	" " "
Condensing plant for l.p.	250	" " 40,000 m3 gas/h

Activated charcoal plant, incl. absorbers, steel structure, blower pipe connections, fittings, etc.	800 tons per 40,000 m3 gas/h
Condensing plant for h.p. incl. coolers	130-150 " " " "
Oil absorption plant with heaters, etc.	about 300 " " " "

57. Capital cost of each step ?

The capital cost for a plant of 40,000 m3 synthesis gas/h is given as follows:

H_2S removal (all erected, incl. foundations)	RM	500,000
Organic S removal (all erected, incl. foundations)	"	350,000
Synthesis per oven, incl. pipe connections, instruments, foundations, building and erection costs, excl. catalyst :		
for low pressure : RM 63,000/	"	65,000
" high " :	"	72,000
Condensing plant, two stages for low pressure :	"	300,000
Activated charcoal plant, excl. charcoal : (charcoal 100 tons at RM 2000/ton)	"	800,000
Condensing plant, high pressure	"	250,000
Oil absorption plant (incl. pipe stills, etc.)	"	600,000

[illegible] 1000 per Ofen = 40 Öfen
40 x 65 = 2 500 000

Blowers, 40,000 m3/h , 3000 mm H_2O each, incl. motor	RM 30,000
Ditto, 80,000 m3/h , incl. motor	" 37,000
H.p. compressor unit for 10,000 m3/h incl. bldg, foundations, pipe connections and erection	" 220,000
Complete compressor plant at Holten with 4 units :	RM 1,600,000

58. Labour cost for each step ?

The labour costs are stated to be as follows:

For purifiers and blowers up to 80,000 m3/h	3 men per shift
For purifiers and blowers 160,000 m3/h	4 " " "
For synthesis plant of 100 ovens	6 " " "
Moreover, for small repairs, cleaning, changing of catalyst (working mostly in daytime only)	10-15 " " "
Condensation (for pumps)	1 man " "
Activated charcoal plant and gasol production	2 men " "
For control laboratory	5 chemists 6 helpers } per day

Repair shop	15-20 men per shift
working two shifts, making in all	30-40 "
Repairs in the plant are executed by these men; part of these are also employed for the ammonia plant.	

59. Cost of maintenance and repairs for each step ?

The costs of maintenance and repairs are 1-1½% of the total capital cost. This figure includes labour and material, but only for the synthesis plant without watergas production.

60. Cost of miscellaneous supplies ?

The costs of miscellaneous supplies are difficult to estimate; for the plant at Holten a rough figure of RM 100,000/year was given.

61. We should like a flow diagram showing type, material and cost for each piece of apparatus.

Regarding the materials used for the various apparatuses, etc., it is answered that only normal iron has been used in the plant, with the exception of the valves, for which a chrome-nickel alloy (2% and 1% respectively) is applied. Moreover, for the h.p. condensers special tubes are used, containing chrome and nickel (probably 5-6% chrome and 3% nickel); the tubes of the heaters in the organic sulphur removal contain 4-6% Cr.