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REDUCTION OF FUSED IRON CATALYST FOR A 10,000 to/yr INSTALLATION

The total amount of catalyst to be reduced for 5 reactors of 12 cbm capacity and 1 reactor of 6 cbm capacity, with a 4 months life of the catalyst amounts to 3 x 66 = 198 cbm catalyst per year. With a bulk weight of 1.7, this represents 337 to iron per year corresponding to 465 to of unreduced iron, or with a bulk weight of 2.0 it would represent 232 cbm of unreduced catalyst (assumed to be Fe_3O_4). The reduction would require [illegible] cbm H_2 per year, assuming the efficiency of hydrogen utilization to be 80%, and this will represent 23 cbm hydrogen per hour. (1 yr. assumed to be equal to 8000 hours). The hydrogen must be very pure, in particular very free from sulfur. We will first describe the reduction installation in Me458 (fig 1). The hydrogen enters at E, is heated to 120° in the preheater V1 (using high pressure steam) and then enters the circuit through the towers A1, A2 filled with LTC coke. The circuit consists of the blowers G1 G2, the preheater V2, the reducing chamber R, the cooler K, the ammonia cooler A and the two silica gel towers Ks1 and Ks2 installed in parallel. Part of the hydrogen is bled off through the blow-off valve U. At the reduction temperature of about 440° almost all the water produced is absorbed in the silica gel towers, because under the existing conditions the dew point of water is just reached behind the NH_3 cooler, and the towers are dried by the circuit consisting of the high pressure steam preheater V3, cooler K2 and blower G5.

Sizes and capacities of the present installation:

Space utilization 1:2000. Time of reduction: 56 hours. Pressure on the suction side of the blower–500 to 500 mm of water. Nominal capacity of the blowers is 600 cbm each. The 2 blowers produce however together only [illegible] cbm per hour of the heated gas against 100 mm Hg. The diameter of the reduction chamber is 1/[illegible] m. The resistance of the catalyst layer 1 m thick with the smallest particle size that can be considered (0.5-1 mm) is 150 mm Hg. The capacity of each silica gel tower is [illegible] of gel. Hydrogen leaves the ammonia cooler at 4° and the silica gel tower at about [illegible]° because of the heat of adsorption which has been set free. The whole system operates discontinuously and must therefore be preheated, cooled, purged with inert gas, and as a result only some 50 cbm of catalyst can be produced in the present installation per year.

The following installation is recommended for the above mentioned 250 cbm of catalyst (see sketch 2).

(Equipment used for the same purposes as in sketch 1, is marked in the same way). The introduction of a continuous method of operation does not seem to be desirable in view of the long time required for the reduction and the introduction of rotating parts (catalyst [illegible]). The installation is therefore made with 2 parallel connected alternatingly operated reduction vessels (R1 and R2) with 8000 hours per year and reduction time of 56 hours, 1.67 cbm should be reduced in one charge. The resistance of the beds limits the catalyst height to 1 m, and the reducer must have a diameter of 1.5-2 m. The blower capacity (G1) should be 3240 cbm per hour. The grate in this installation, which supports the catalyst is composed of 2 parts which can be rotated on a shaft located in the middle of the reducer. This arrangement is desirable primarily because of the large weight of the catalyst, but also because of the poor packing properties of the latter. A proper kind of grate remains yet to be constructed. The second reducer must be cooled during the 56 hours required for the reduction of one charge, flushed with CO_2, emptied, filled, and again flushed with H_2. The cooling in the sketch is suggested by using some of the circulation hydrogen after branching it off at Z. The hydrogen is cooled in an additional cooler K3 and reenters the circuit behind Z. The flushing with CO_2 can be done after the reducer temperature has dropped to 80°C, and the CO_2 required in the flushing as well as during the filling of the synthesis reactor must be free from sulfur, and dry (longer storage of the catalyst must be done in an atmosphere of H_2). The CO_2 coming from the CO_2 scrubbing is used after being desulfurized with lumps of the Lux mass (red mud from alumina production) mixed with clay. The Lux mass is regenerated with air. Up to 60 cbm per hour of CO_2 is required, but it is only used occasionally (for the flushing of the reducers). The desulfurizing requires 2 containers with about

1 cm cross section and 1 ohm content for the pea size Lauta mass.

To desulfurize CO_2 it must be heated to 30-35°. In connection with desulfurizing 2 silica gel towers are required, each holding 100 li. of the gel, which must be alternately dried, preferably with CO_2.

In the filling of the reducer a manhole is needed of sufficient size and best located in front, through which the catalyst may be readily introduced and uniformly distributed. In this case the H_2 should best be introduced through the circuit pipeline similar to that of the H_2 outlet. For safety sake the H_2 and CO_2 supply and removal must be closed off with blinds or some similar arrangements. After the charge has been introduced, the CO_2 is driven out by H_2 into the open and the container started.

The following changes have been made against the older installations: A reduction temperature of 500° must be reached. A heat exchanger is provided. The H_2 outlet is preferably arranged in front of the KG towers. The preheater for the drying of the KG towers is provided with a by-pass, because the towers must be cooled after drying. The KG towers in the present installation have been designed somewhat too small in size and should be increased about 1.5 times. Otherwise their conditions are properly designed and can be carried over into the new installation with the same proportion.

The present heating gas and air blowers must be operated to capacity.

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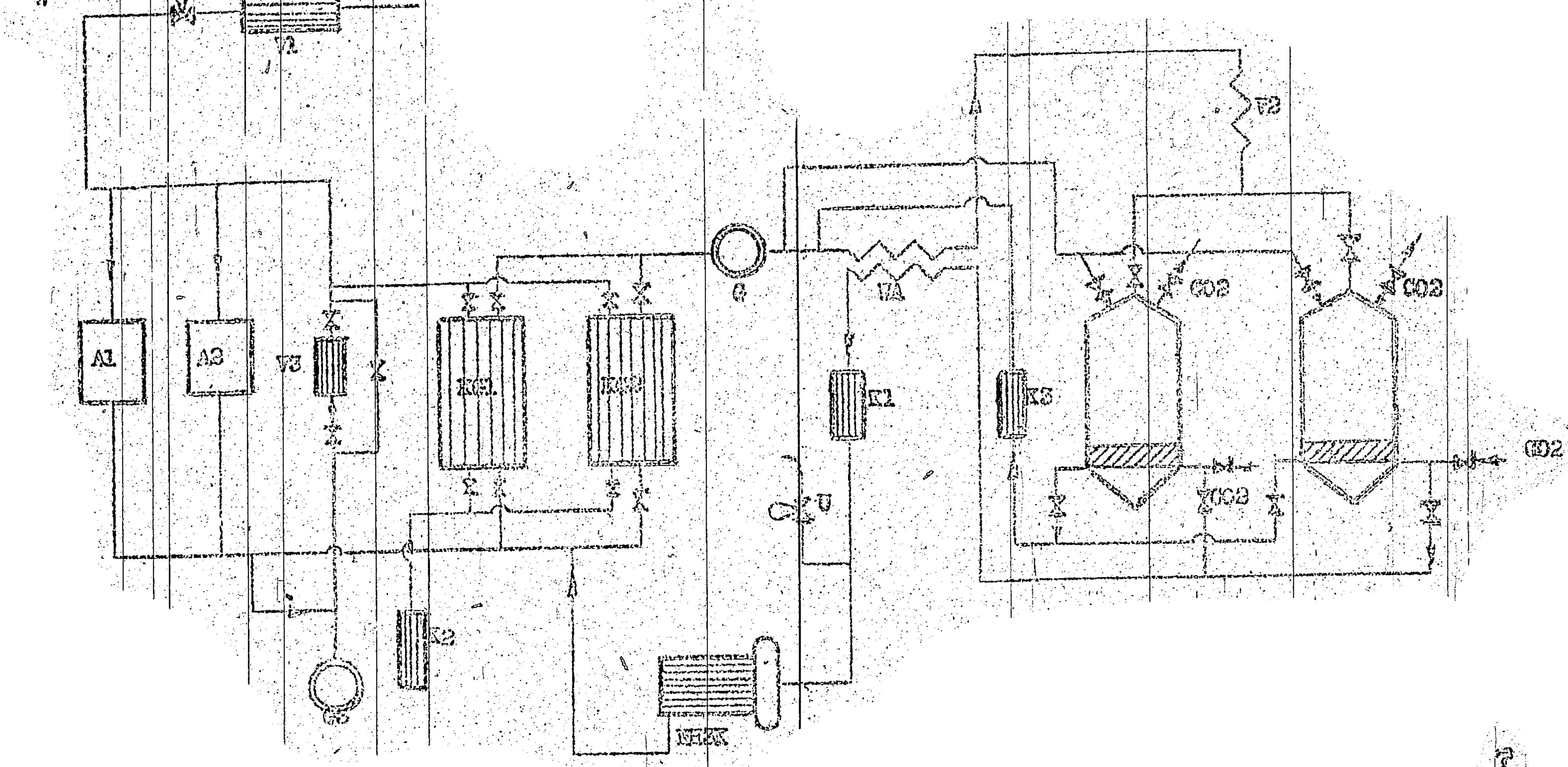

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