EXPERIENCE WITH FLUIDIZED BED FISCHER-TROPSCH REACTORS AT SASOL

Mark E DRY
SASOL R&D, P O Box 1, SASOLBURG 9570 SOUTH AFRICA

ABSTRACT

In the mid-seventies the Sasol circulating fluidized bed reactors were re-designed. The internal heat exchangers were improved and the capacity per reactor was increased three fold. Further scale-up appeared impractical and so, looking to the future, Sasol R&D decided to re-investigate the feasibility of using the classical fluidized bed system.

Extensive cold model fluidization studies was followed by the construction in 1983 of a 1 metre ID demonstration reactor. In both studies the main concern was the development of a reliable gas distributor. The success of these developments led to the construction of a commercial scale reactor which came on stream in mid-1989. The reactor has met all expectations. The capital cost is half of that of a circulating bed reactor. Operation is simpler and much more flexible. Running costs are also lower because of the lower differential pressure. Maintenance costs due to erosion are lower because of lower linear velocities. Higher heat exchange area has made it feasible to utilize the benefits of higher operating pressure and thus removed a major limitation on further scale-up.

On the demo unit, initial tests with metal filters have been promising. Implementation would further lower capital cost and improve the thermal efficiency.

HISTORY UP TO 1974

Sasol uses both fixed and fluidized bed Fischer-Tropsch (F-T) synthesis reactors. When the first plant was designed in the early fifties, Sasol decided to use the Kellogg circulating fluidized bed (CFB) concept (see Figure 1). This involved the risk of scale-up from a four inch ID pilot unit directly to the Sasol One commercial reactors. In hind-sight it is not surprising that many teething problems were experienced and several reactor design and catalyst formulation changes were made before the commercial operation became technically successful. These altered reactors subsequently became known as the Sasol Synthol reactors.

Because on a small scale the "classical" fluidized system (see figure 2) is much simpler to build and to operate than the circulating system, the pilot plant reactors used at Sasol always were the "classical" types. Since, under apparently similar conditions, these pilot units out-performed the commercial CFB units, there was a longstanding desire in the Sasol Research department to investigate the simpler "classical" option.

As a result of the crude oil crises during the early seventies, Sasol decided to build a much bigger automotive fuels producing plant using the fluidized bed technology. The USA Badger company put forward two design proposals, firstly a scaled up version of the Sasol CFB units, and secondly, the "classical" option (Badger had considerable experience with the latter system in other applications). Because of the urgency of the overall project, Sasol decided to stay with the proven CFB process, but the second Badger option at least spurred Sasol Research into actively investigating the classical option.

OPERATION OF THE SYNTHOL CFB REACTORS

The circulating fluidized bed reactor is depicted in Figure 1. The gas (fresh feed plus recycle) is preheated to about 200°C and introduced into the bottom of the reactor where it meets a stream of hot (about 340°C) catalyst flowing down the standpipe. Gas plus catalyst flow up through the reaction zone where two banks of heat exchangers remove a large portion of the heat of reaction, the balance being absorbed by the recycle and product gases. In the wide hopper section, the catalyst disengages from the gas, because of the lower linear velocity there, and the catalyst flows down the standpipe to continue the cycle. The rate of flow of catalyst is controlled by the slide-valve at the bottom of the standpipe. The gas leaves the reactor via highly efficient (greater than 99%) cyclones which remove entrained fine catalyst particles and returns them to the hopper. A small amount of catalyst fines do nevertheless get through the cyclones and this necessitates the use of a heavy oil quench scrubber which removes the last traces of catalyst. The oil scrubber is a relatively large unit and so adds to both the capital and running costs.

The catalyst flowing down the standpipe is in a "dense phase fluidized" state and it is important to keep it in this state in order to maintain the required differential pressure down the standpipe and also to ensure smooth flow. If the catalyst defluidizes, unstable "slip-stick" flow results and could lead to bridging of the standpipe which would stop catalyst flow altogether. The catalyst flowing up the reaction zone is in the "lean phase" state and the slip velocity is relatively high. Because the iron catalyst used has a high density it is intrinsically more difficult to fluidize. Hence a high linear gas velocity is required. The high velocity results in a high differential pressure over the reactor, ie, increases the operating costs.

The differential pressure across the standpipe/hopper must be higher than across the "reaction" zone, otherwise the gas would flow up the standpipe and this could result in catalyst "puking" and temperature runaways. This means that the amount of catalyst flowing out from the standpipe, must be kept below a certain value.

To achieve satisfactory catalyst flow up the "reaction" zone it is essential to maintain a minimum gas linear velocity. Hence if the fresh feed flow decreases, the recycle flow has to be increased. Increasing the recycle ratio results in a change in the gas composition inside the reactor which results in changes in product selectivity and increases the rate of carbon deposition on the catalyst. The turn-down ratio is therefore limited.

Under normal commercial operating conditions, carbon deposition on the iron based catalyst is inevitable. This does not have a strong negative effect on the intrinsic activity of the catalyst but it does result in a lower bulk density of the particles. While this means that the catalyst becomes easier to fluidize, it unfortunately also results in a decrease of the differential pressure across the standpipe. As the catalyst's activity declines (due to sintering, poisoning and fouling) one would like to compensate for this by increasing the catalyst loading in the reaction zone. However, because of the decline in the catalyst's density, this is not possible (for the reasons already discussed).

The high gas linear velocity coupled with the erosive nature of the catalyst, does result in erosion of the reactor lining material in certain areas and this means that run lengths have to be limited in order to allow regular maintenance inspections, and if required, repairs.

CFB REACTOR AND PROCESS IMPROVEMENTS

Having decided in 1974 rather to stay with the known and proven CFB Synthol units, it was decided to incorporate several improvements into the next generation reactors. The old Synthol internal heat exchanger banks were multi-tubular and oil cooled. Selective catalyst flow through the tubes resulted in poor heat exchange and also in tube blockages (which increased maintenance costs). In the new reactors direct steam generating coils were installed which were thermally more efficient and also had much lower maintenance costs. The slide valve design was altered as was the design of the catalyst settling hopper which resulted in better catalyst flow down the standpipe. The physical size of the reactor was also enlarged.

Studies previously carried out in the Sasol pilot reactors showed that under commercial conditions the kinetics of the F-T reaction with iron catalysts could be described by the rate equation

$$k \ p_{CO} \ p_{H_2} / (p_{CO} + a \ p_{H_2O})$$

and that the rate of carbon deposition was proportional to

$$p_{CO} \ / \ p_{H_2}^{\ 2}.$$

Both these equations predicted that there were incentives to operate at higher pressures. The reaction rate increases with pressure which means that the fresh feed can be increased in proportion to the increase in pressure and the percent conversion would remain unchanged. The rate of carbon deposition decreases at higher pressure which means that the rate of decline in catalyst bulk density will be lower which translates to a higher differential pressure over the standpipe. Consequently higher catalyst loadings can be maintained in the reaction zone which in turn should result in higher conversions.

The combination of physically larger units operating at higher pressure resulted in a three-fold increase in capacity per Synthol reactor. There are sixteen of these new Synthol units in operation at Sasol and their performances have been in line with predictions.

DEVELOPMENT OF THE ALTERNATIVE FLUIDIZED BED REACTOR

The already large size of the new Synthol CFB reactors makes it impracticle to physically increase their size further. Due to the difficulty of installing more internal heat exchanger coils in the reactor, the utilization of even higher pressure, to increase production capacity, is also limited. Because of these limitations and also since it was estimated that the classical fluidized bed system should be cheaper to build and to operate, Sasol decided to investigate this alternative.

The "quality" of fluidization, which reflects the effectiveness of contact between the gas and catalyst, was perceived to be the main area of concern. The design of the gas distribution system and the influence of the reactor internals (eg the cooling coils) on the fluidization were key items to investigate. Several cold models were built, varying in size from 2 to about 30 inches diameter. The catalyst particle size distribution was shown to be a key factor in the quality of fluidization. It was also found that the linear gas velocity could be varied over a wide range without adversely affecting fluidization quality.

These studies gave Sasol and Badger the confidence to design a one meter diameter demonstration reactor operating under normal F-T synthesis conditions. This unit came on line in 1983 and was operated under various conditions to test several design options.

The success of this demonstration unit led to the design of a full scale commercial unit which came on stream in mid-1989. The performance has met all expectations. Higher conversions were achieved and the selectivities were similar, and in some aspects better, when compared to the CFB Synthol units.

ADVANTAGES OF THE NEW REACTOR

The cost of the new reactor will be half of that of an equal capacity CFB unit. Comparison of Figures 1 and 2 which are drawn roughly to the same scale, make it obvious that the new reactor is much cheaper. The size of the reactor is not very different from that of the catalyst settling hopper of the CFB unit. Because of the complexity of the CFB unit, it requires an expensive support structure whereas the new unit is simply supported on a skirt. The required operating platforms could be supported from the reactor itself. The structure of the new reactor costs about 5% of that of an equal capacity CFB unit.

Because of the lower gas linear velocity the pressure differential over the new reactor is about half of that over the CFB. This results in a saving of about 20% on total feed compression costs. Also because of the lower linear velocities inside the reactor, the maintenance costs related to erosion by the catalyst are clearly lower although more running time is required to quantify this saving.

Whereas at any instant only a part of the catalyst in a CFB reactor is in the reaction zone, all of the catalyst in the new reactor is in contact with the feed gas and this partially accounts for the higher conversion observed for the latter case. As explained previously, carbon deposition results in expansion of the catalyst bed which results in a lower pressure drop over the CFB's standpipe and consequently makes it impossible to increase catalyst loading in the reaction zone when it is needed most (when the intrinsic activity of the catalyst has declined). In the new reactor carbon deposition also results in expansion of the catalyst bed but in this case it is beneficial as it increases the residence time of the gas in the catalyst bed and so tends to compensate for the normal loss of catalyst activity. This factor also contributes to the higher overall conversion achieved in the new reactor.

Because of the relative simplicity of the reactor, it is easy to operate. There is no pressure balance to be concerned about as is the case for the CFB unit. Re-starting the unit after prolonged feed gas interruptions (because of external reasons) has been found to be a much smoother and simpler operation. This could mean that less skilled, or fewer operators are needed to run the units. It has also been demonstrated that lowering the linear velocity of the total feed gas has no adverse effect on the operability of the reactor. As was previously described, this is not easily achieved with the CFB reators. Thus in the new reactor the turn down ratio is much larger.

The maximum coiling coil area that can be physically installed in the new reactor, is considerably more than is feasible in a CFB reactor of the equivalent capacity. This means that production capacity could be increased since more reaction heat can be coped with. Production can be increased by increasing the fresh feed flow and simultaneously decreasing the recycle flow. This could also improve the thermal efficiency (less gas to cool down, recycle and then re-heat again). This mode of operation would, however, result in a lower percent conversion. Alternatively the total operating pressure could be increased with a proportional increase in the fresh feed flow. In this operating mode the percent conversion would remain high.

An unexpected finding has been that the cyclones in the new reactor are much more efficient than similar cyclones in the CFB reactor. The consequence is that much less catalyst dust is carried out of the reactor to the oil scrubber system downstream. The reason for this improvement is due to the difference in flow patterns of catalyst to the cyclones.

FILTERS INSTEAD OF CYCLONES FOR CATALYST FINES REMOVAL

As mentioned, the cyclones are very efficient but they will never be perfect, which means that there will always be a need for the downstream scrubber unit. Now if the cyclones could be replaced by filters, the scrubber tower would not be required which would further lower the capital outlay and also improve the overall efficiency of the cooling train.

Some time ago, a set of porous metal filters was installed in the demonstration reactor and the process appeared to operate satisfactorily, although a much longer operating time would be needed to prove the mechanical reliability of the system. Further testing of the filters is being planned. When cyclones are used, the catalyst ultra fines produced during synthesis pass through the cyclones and so leave the reactor. When filters are used, these fines will build up in the catalyst bed and the influence of these fines on the quality of fluidization and hence on conversion, still needs to be quantified thoroughly.

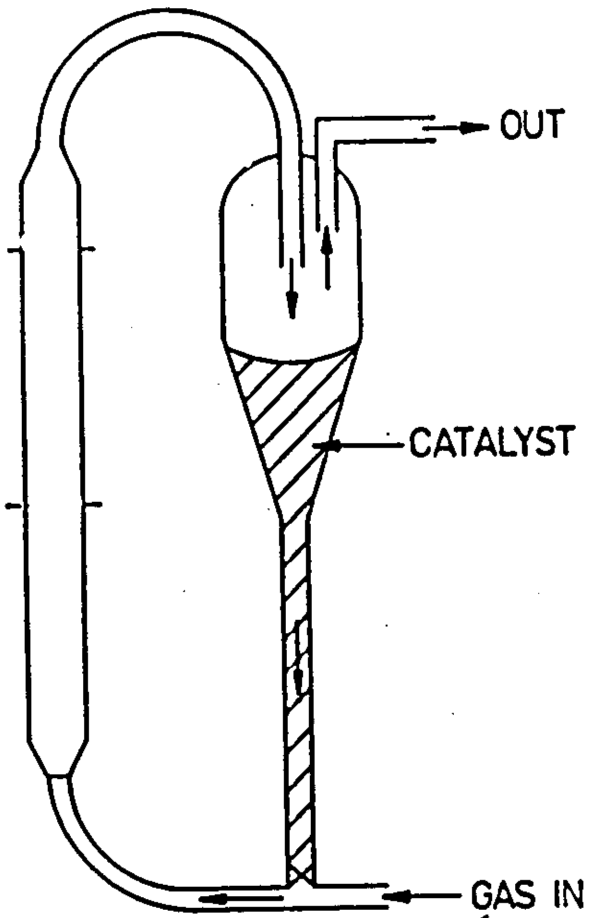

FIG. 1

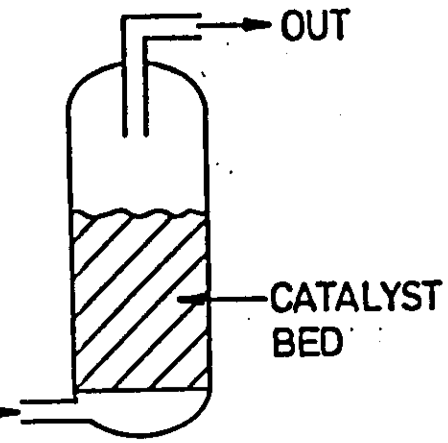

FIG 2