(19) Europäisches Patentamt
European Patent Office
Office européen des brevets

(11) Publication number: **0 580 941 A1**

(12) EUROPEAN PATENT APPLICATION

(21) Application number: **93103451.6**

(51) Int. Cl.5: **C10G 65/12**

(22) Date of filing: **04.03.93**

(30) Priority: **24.07.92 IT GE920078**

(43) Date of publication of application:
02.02.94 Bulletin 94/05

(84) Designated Contracting States:
CH DE ES GR LI

(71) Applicant: **PROMETHEUS S.r.l.**
Piazza Borgo Pila n. 40
I-16129 Genova(IT)

(72) Inventor: **Ferrucci, Alberto**
Via Orsini n. 24/6
I-16146 Genova(IT)

(74) Representative: **Ferrarotti, Giovanni**
Studio di Consulenza Tecnica
Via Alla Porta Degli Archi 1/7
I-16121 Genova (IT)

(54) Hydrotreating single unit for processing different combined petroleum fractions coming from atmospheric and vacuum distillation.

(57) Process for the contemporary treatment within a single unit for desulphurization, denitrification and aromatic saturation of both petroleum fractions obtained from atmospheric and vacuum distillation, based on two separate reaction section (22, 28) and a single heating furnace (37) ubicated between the two reaction sections (22, 28) operating in parallel on both sections.

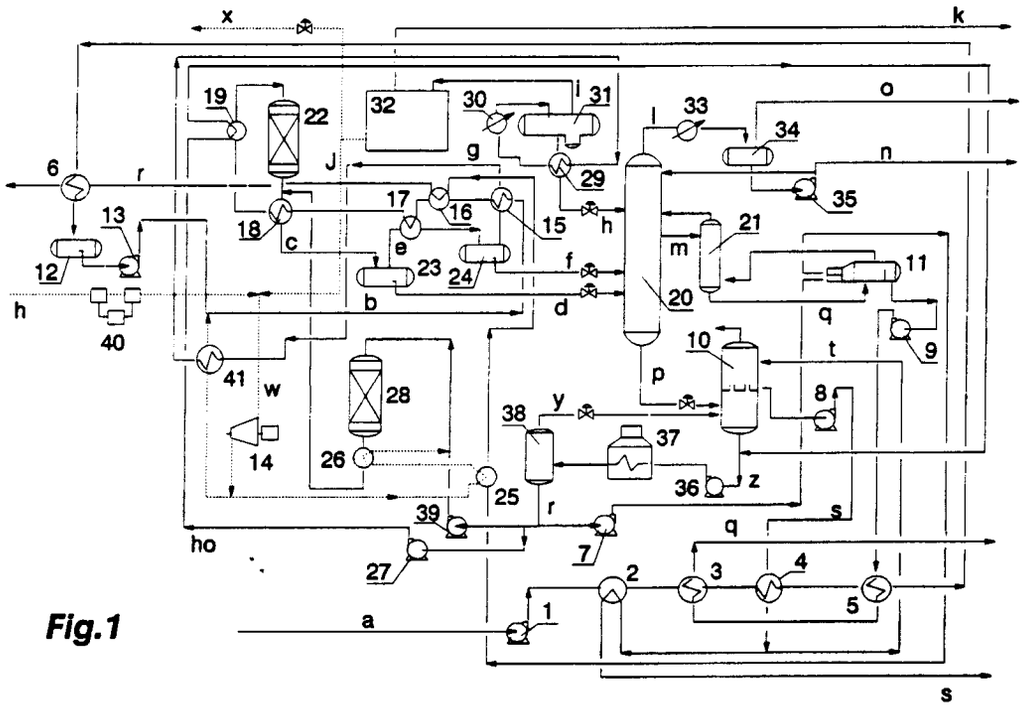

Fig.1

EP 0 580 941 A1

The technology of low severity hydrocracking or "Mild Hydrocracking" (MHC) grew in the early 1980's as application aimed to meet the increasing demand, coming from the refiners, for a deeper vacuum distillation fractions (VGO) cleanup and its molecular-weight reduction, or conversion.

Over the years, the MHC term has been applied to a wide range of processing conditions and is now used indiscriminately to describe either low (compared with conventional hydrocrackers) pressure units, which can be quite severe operations, or low-conversion units, which can be quite mild operations.

In this description MHC is used to mean a moderately low pressure (55-85 Kg/cm2) and moderate conversion (20-45%) operation in which the the feedstock is mildly upgraded, by catalytic hydrogenation and molecular weight reduction, to lighter fractions.

Tipical characteristic of MHC processes is to perform simultaneously desulphurization, denitrification and moderate hydrocracking within a single reaction section.

Normal VGO hydrotreating units (HT), equipped with standard catalyst (Nikel Molibdenum or Cobalt Molibdenum) working constantly at their optimal operating conditions, are not capable to achieve comparable conversion performances.

On the other hand, running a unit within the range of said optimal conditions, hydrogenation capabilities of MHC catalyst become more difficult to be mantained as VGO conversion increases, being not possible to compensate the deactivation due to coke deposits on the catalyst.

Bearing in mind that the MHC technology has been expressely developed for achieving a moderate hydroconversion in existing VGO hydrotreating units, once properly revamped (simply by changing the catalyst and the associated operating conditions or, more heavily, by changing and/or adding reactors), it becomes clear what are the technical and economical problems which arise from the coercive coexistence of the above different needs partially in conflict.

If it is privileged the hydrodesulphurization (HDS) activity then it becomes convenient to run the unit steadily at low severity conditions, and to compensate catalyst deactivation by moderate increases of the temperature (tipically between 0.5 and 1.5 °C/month).

As previously noted, the optimal conditions of this asset do not allow to achieve appreciable levels of conversion (MHC catalytic activity requires, in fact, quite different levels in temperature and space velocity). By the way, the recent evolutional trend in commercially available HDS catalysts shows an increasing activy at lower temperature which is further widening said gap.

On the contrary, if it is privileged the conversion, then the more severe conditions, under which the unit has to be run, will determine an excess of HDS activity and a subsequent faster catalyst deactivation with the result of a shorter operating cycle of the catalyst, more frequent turn down of the unit and increased operating costs of unit.

Besides, to mantain an appreciable conversion level it is necessary, on its turn, to provide for a heavier temperature compensation of catalyst deactivation (tipically, for MHC catalysts, between 1.5 and 3.0 °C/month).

Normally the problem to evaluate the best compromise between the above diverging needs and requirements is solved by refiners on the basis of the critical factors associated with the plant and operative configuration, the characteristics and quantities of processed feedstocks, product specification and market requirements.

A further element of compromise is due to the fact that the conversion catalyst, which requires a higher level of temperature, usually is ubicated downstream the HDS catalyst in order to utilize the associated exothermicity.

This makes inevitable to send the sulfidric acid and the ammonia, resulting as HDS by-products, on the MHC catalyst. The resulting effect is to contribute, in addition to the coke formation, to the progressive deactivation of the MHC catalyst, offsetting conversion and stability performance which become dependent of the quality of the feedstock.

Hence the foreclosure, for the most of the MHC units actually in operation, to the utilization of zeolite catalysts (recently developed and presenting a greater activity at lower temperatures associated with a greater stability). It is known, in fact, that the different behavior between traditional amorphous catalysts (silica-alumina, silica-magnesium) and zeolite ones is due to the increased acidity of the latest and thus their greater vulnerability to the effect of the ammonia.

At present time, notwithstanding that in order to counteract this problem a "second generation" of MHC catalyst with both HDS and MHC functions on the same catalyst has been developed, the above mentioned conflict remains practically almost unchanged as remains unchanged the sensitivity of conversion and stability performances to the nitrogen content of the feedstock.

It is possible then to affirm that MHC technology, although making possible a considerable number of applications with reduced capital investment, underlies to a sort of "operating dualism" arising from the overlap of different targets (conversion and desulphurization) each other partially in conflict for what concerns their optimal operating conditions.

On the other hand, the costs associated with the interventions to be carried out, within and existing HT unit to be converted to MHC, to overcome such operating dualism would be so hight to make the overall project being reconsidered and its cost/benefits ratio compared with those resulting from more complex alternatives (as f.i. installation of a thermal or catalitic conversion unit downstream the existing VGO HT unit).

Obviously, it is clear that such kind of interventions are economically unapplicable in those cases where the MHC technology is applied to pretreat the feedstock of an existing conversion unit (for improving yields and product quality).

As a consequence of the above, the more economically attractive fields of application for current MHC technology can be therefore summarized in the following two tipical motivations:

- moderate capital investments, or,
- performance improvement of existing Fluid Catalitic Cracking or Thermal Cracking units.

Nevertheless, more recently, new elements able to modify the above picture are foreseeable for the next future.

Law trends for tougher standards on petroleum products in general, and diesel fuels in particular, which are pioneered in some contries and will be sooner or later certainly introduced, under the public's opinion enviromental pressure, in a more or less severe measure in all western countries; those toughter standards will rappresent the innovative aspect that will characterize the 1990's and determine new needs and priorities within the petroleum refining industry, upsetting current traditional processing schemes.

More in detail, for what concerning diesel fuels, quality improvement trends can be mainly summarized as follows:

- progressive sulphur content reduction to 0.1-0.05%;
- progressive aromatics content reduction to 20% (and even to 10% in some areas):
  Above targets has as common denominator the consciousness that, in the next future, it will be mandatory to process light gasoils, coming from topping units and/or from conversion units, under more severe conditions than those actually performed.

In fact, as revealed by several specific studies on the matter, the presence of sulphur compounds in middle distillates (240-400 °C) is characterized by a wide variety of degrees of reactivity.

The major part of the refractory sulphur compounds is identified in the 330-380 °C boiling range as alkylated dibenzothiophenes.

These compounds remains practically untouched when operating under tipical process conditions of conventional hydrotreaters.

Practically this means that, to meet the above mentioned sulphur levels, gasoil boiling enpoint has to be lowered, in the most of cases, around to 340 °C or, alternatively, specific HT unit designed for operating under more severe pressure has to be provided.

This same conclusion is reached facing the problem of the reduction of the aromatics content in the gasoils.

Also in this case the suitability of existing gasoil HT unit appears, in the most of situations, not possible (unless the use, for achieving low pressure hydrogenation, of noble metal catalyst in a 2-stage process) so it becomes obligatory, also for this case, to process light gasoils under more severe conditions.

The resulting limitation of the range of gasoils processable through the existing HT units rappresents, in the most of cases, a new element upsetting traditional consolidated costs/benefits ratios of the current petroleum processing industry.

To give a rought indication, in practical terms, of the above it is possible to say that for a modern and well equipped refinery operating competitively on the international market to lower sulphur content in gasoil from 0,30 wt% to 0.15 wt% means additional costs of about 1 cent/liter, while the cost for reducing the maximum aromatics content from 30 vol% to 20 vol% is estimated around 3 cent/liter which is near to quadruple for contents near to 10 vol%.

This situation outlines economical margins for the development of a new generation of gasoil HT units designed with the aim to recover large fractions of virgin or conversion distillates, making them suitable for the formulation of the high quality "enviromental" gasoil and to compesate, in such way, refinery flexibility which otherwise would be compromized.

Basically, the hydrotreament processes today existing in the petroleum industry can be summarized as follows:

- units designed for the desulphurization, denitrification and moderate catalytic conversion of vacuum distillates, not able to process also atmospheric distillates;
- units designed for the desuphurization and denitrification of atmospheric distillates, not able to process vacuum distillates;
- units designed or adapted to desulphurize and denitrificate atmospheric or vacuum distillates by processing them alternatively in processing campaigns.

In accordance with FR-A-2 433 944, it is also known a process for the treatment, in a single unit, of heavier petroleum fractions for achieving a previuosly determined level of conversion. Said process, which employs two reaction sections

served by two separate preheating furnaces each having a specific purpose, has the first furnace operating on the untreated feedstock.

The process, according to this invention, has the aim not only to realize, within a single unit, a VGO MHC process and the deep desulphurization of light gasoils but also to overcome the above mentioned operating dualism which so heavily is conditioning plant operation management on existing MHC units.

The process in fact, according to this invention, make it possible:

- to desulphurize, denitrificate and dearomatize a wide variety of ligh gasoils (straigh-run and/or from conversion units) making them suitable to the most severe standars actually foreseen for sulphur and aromatics content in diesel fuels;
- to desulphurize and hydrogenate fractions of vacuum distillates giving them not only good "crackability" characteristic, useful for a possible further stage of conversion, but also providing an efficient denitrification of the same fractions.
- to fully utilize the newer HDS catalysts, more active and stable at lower temperature, without affecting the conversion;
- to be able to utilize zeolite conversion catalyst, that are more active than the traditional MHC ones; this because the poisoning by-products (NH3 add H2S), that would cause an inacceptable deactivation rate on such catalysts, are previously removed from the feedstock;
- to operate continuously in segregated and optimized mode the HT and MHD functions in parallel, free from the above mentioned inconsistence of diverging needs;
- to mantain a good selectivity toward the products also compensating possible low efficiency fractionation performances of the upstream distillation units, and/or possible cut point fluctuations of the feedstock;

At last, it has to be noted that the presented process, for its innovative characteristics, rappresents a profitable opportunity for those small hydroskimming refineries runned with high operative flexibility, to become, in short time and with moderate capital investments, more competitive if compared with larger and more sophisticated industrial complexes with whom, otherwise, in view of the additional costs required to meet the above more severe gasoil specifications, the competion would be practically foreclosed.

Basically, according to this invention, the process consist of a single unit which is able to perform desulphurization and denitrification treatment of both atmospheric and vacuum distillates and, at the same time, a conversion treatment of the heavier fractions with a separate and indipendent control and management of the operating parameter of each treatment.

Such process consist of:

- two separate reaction sections, one processing all the feedstock coming from either atmospheric distillation or vacuum distillation the other processing the bottom of the vacuum distillation unit,
- a single furnace installed between the two reaction section and operating on both in parallel,
- a single section for the product fractionation cosisting of two parts: atmospheric and vacuum distillation,
- a phase separator in which the vapour, obtained by heating in the furnace stream coming from the vacuum bottom, is returned to the vacuum distillation column, after the separation from the liquid fraction which is partly send to the conversion reactor and partly to preheat the desulphurization reactor,
- equipments for the removal of the gaseous contaminants (NH3 and H2S) produced in the desulphurization reactor, before feeding the conversion reactor.

The technical and economical advantages of the process, according to this invention, are below explained.

While the worldwide demand for petroleum products is expected to increase, as the need for their better quality, now the petroleum refining industry is facing, for enviromental reasons, increasing difficulties to add the refining capacity necessary to meet the above mentioned growing demand.

The obvious consequence of the above is the convenience to make the best possible use of existing refining capacities, realizing more appropriate less-expansive technological upgradings instead of draw efforts to costly and more enviromental concerned erection of new refineries.

It is just within such scenario the presented process can achieve substancial opportunities.

As previously stated the process rappresents a significant low-cost tool for the technological upgrading of tipical small hydroskimming refineries allowing to meet the most stringent product-quality standards today requested by the international market and assuring, at the same time, process profitability by enhancing the yields of more valuable products, gasoil in particular, on processed crude.

According to this invention, in fact, it becomes possible to perform desulphurization-denitrification treatment of both atmospheric and vacuum distillates and, at the same time, conversion treatment

of the heavier fractions within a single process unit, allowing their separate and indipendent operating control.

Moreover, the process has been also developed in such a way as to minimize liquid or gaseous pollutant effluents.

The resulting effect is to realize, in a single step, some of the most important processing stages, which are mandatory for a modern refinery, extending the possibility of their implementation also to those small refineries where, due to their low-scale processing capacity, would not be achievable enough cash flow to balance the capital costs involved in technical solution considering separate process units .

In addition, due to its reduced area requirements, the process has the characteristic of being attractive expecially for those refineries presenting space availability limitation for new plants as usually happens in those ones ubicated close to urban areas.

In this application we refer for heat exchanging, catalytic and fractionation sections to known technical solutions. What is the innovative subject matter of this invention is the original combination of said known technical elements and the process conditions which have been conceived to perform in a flexible way all process functions concentrating, in case of plant of limited capacity, all process heat requirements in a single heating fournace.

Furthermore, by this process, it becomes possible to achive good performances and competitive plant arrangements also in those refineries presenting low efficiency crude fractionators, once it has been improved the efficiency of vacuum section only.

Summarizing what has been already explained, the process, according to this invention, allows:

- to realize, in a separate continuous asset and balancing hydrogen consumption with the quantity available within the refinery, a deep desulphurization-denitrification treatment of the heavier fractions (VGO) followed by their mild hydrocracking conversion;
- to minimize the production of pollutant effluents to be processed in auxiliary service units (off-sites);
- a noticeable reduction of the aromatics content in gasoil fractions which otherwise would not be suitable to meet the expected diesel future requirements;
- to overcome problems due to space limitation or situations presenting tight lay-out constraints for new plants installation (f.e. in sites close to urban areas);
- the utilization of the newer, more active and/or selective catalysts commercially available, each at its optimal operating conditions (so reducing deactivation rate and operating costs) due to the segregation, realized by the process, of the desulphurization-denitrogenation treatment from the catalytical mild hydroconversion section;
- the preservation of optimal operating condition in each reactor separately not only during the start-up conditions, but also through the operating cycle being possible the variation of process conditions (f.e. as to compensate catalyst aging) irrespective from feedstock quantity/quality and process thermal integrations;
- an attractive opportunity also for more complex refineries: the process, in fact, producing a hydrotreated vacuum distillate with good paraffinic characteristics, can be succesfully implemented upstream an existing conversion unit (f.i. catalytic or thermal cracking) giving both yield and product quality enhancement. In particular, in case of a thermal cracking unit, said distillate would eliminate pollutant emissions and unpleasant smells normally produced in such processes.

The process, according to this invention, is illustrated in principle in the flow diagram of Fig.1 in which appear:

- 1 : feed pump (1st stage) ,
- 2 : heat exchanger feed/light gasoil from 4,
- 3 : heat exchanger feed/light gasoil from 5,
- 4 : heat exchanger feed/light gasoil from 8,
- 5 : heat exchanger feed/gasoil from 9,
- 6 : heat exchanger feed/residue to battery limits,
- 7 : pump for residue from 38 to 11,
- 8 : pump for light gasoil from 10,
- 9 : pump for light gasoil from 11,
- 10 : vacuum distillation column,
- 11 : gasoil side stripper reboiler,
- 12 : preheated feed surge drum,
- 13 : feed pump (2nd stage),
- 14 : hydrogen recycling compressor,
- 15 : heat exchanger feed/off-gases from 24,
- 16 : heat exchanger feed/residue from 25,
- 17 : heat exchanger feed/off-gases from 23,
- 18 : heat exchanger feed/effluent 22 and 28,
- 19 : heat exchanger feed/residue from 27,
- 20 : product fractionator,
- 21 : gasoil side stripper,
- 22 : hydrodesulphurization reactor,
- 23 : 1st hot separator,
- 24 : 2nd hot separator,
- 25 : heat exchanger residue from 11/hydrogen from 14,
- 26 : heat exchanger effluent from 28/hydrogen from 25,
- 27 : pump for liquid effluent from 38 to 19 ("hot oil" circuit)

- 28 : mild hydrocracking reactor,
- 29 : heat exchanger gaseous stream from 41/liquid effluent from 31,
- 30 : trim cooler gaseous stream from 29 to 31,
- 31 : cool separator,
- 32 : hydrogen recovery and purification section adopting the known technology of water and subsequent amminic gas scrubbing,
- 33 : off gas condenser,
- 34 : reflux drum,
- 35 : pump for naphtha product to battery limits and reflux to 20,
- 36 : pump for residue from 10 to 37,
- 37 : heating furnace,
- 38 : phase separator of the residue from 37,
- 39 : surge pump to 28,
- 40 : hydrogen make-up booster compressor,
- 41 : heat exchanger gaseous stream from 15/hydrogen from 14.

According to the above, the conceptual operating scheme of the process is as follows.

The process feedstock a, a mix of variable ratios of petroleum atmospheric and vacuum distillates, is sent by the first stage feed pump through the first series heat exchangers 2, 3, 4, 5, and 6 operating at a lower pressure for a first preheating.

The preheated feedstock, recovered in the surge drum 12, is therefore pumped and mixed with the hydrogen coming from the recycle compressor 14, previously preheated in the heat exchanger 24.

The feed mixed with hydrogen b is further heated through the second series of heat exchangers 15, 16, 17, 18, and 19 which are under higher pressure.

The feedstock is therefore sent to the first reactor 22 where hydrodesulphuration, denitrification and dearomatization reactions of the feedstock take place.

The effluent from the reactor 22, once mixed with the effluent from the reactor 28 and cooled in the heat exchanger 18, is sent to the first hot separator 23.

Here a phase separation takes place, the liquid phase d is sent, under pressure control, to the product fractionator 20, while the off gases e, partially condensed in the heat exchanger 17, are sent to the further hot separator 24.

The liquid phase obtained from the hot separator 24 f is also sent to the product fractionator 20 while the off gases g are cooled and condensed through the heat exchangers 15, 41, 29 and 30 and sent to the cool separator 31.

The liquid h obtained from said cool separator 31, after contributing to cool off-gases g in the heat exchanger 29, is sent to the top section of the product fractionator 20. The gaseous phase obtained from the cool separator 31 is fed to a known system of gas scrubbing for the recovery of the hydrogen to be recycled and the concentration of the sulphidric acid and ammonia coming from the reactor 22 (to be sent to a further sulphur recovery plant).

The recovered hydrogen j, once purified from sulphur and ammonia, is send to the recycle compressor 14 together with the high purity hydrogen of make up.

Hydrogen purity level in the overall treat gas w sent to the recycle compressor 14 is mantained by controlling the purge gas x volume.

The off gases l from the fractionator 20 are condensed in the cooler 33, collected in the reflux drum 34 from where uncondesable gases are sent to battery limits (for their purification before their use as refinery fuel gas) while the condensate naphtha is partly recycled and partly sent to battery limits to a known stabilization debutanizing treatment, to be ubicated downstream pump 35.

From the product fractionator 20 a gasoil m is also taken. This gasoil is further stabilized in the side stripper 21, utilizing the heat obtained by the reboiler 11 from hot vacuum residue, and sent to the battery limits as stabilized product q by pump 9 through the heat exchangers 5 and 3.

The bottom stream p of the main fractionator is sent direcly to the vacuum distillation column 10 for a further light gasoil recovery. The recovered gasoil s is partly used to wash the upper section of the vacuum column itself and partly sent to the battery limit after its quenching through the heat exchanger 2.

The residue z obtained from the vacuum distillation column 10 is sent by the pump 36 to the heating furnace 37 and then, under pressure control, to the separator 38. Here the vapours y obtained are returned, under temperature and pressure cascade regulation, back to the vacuum distillation column where they condense giving the heat required for the vacuum distillation.

The liquid effluent r coming from the phase separator 38 is sent:

- partly, by pump 7, to the battery limits for tankage, previuos its quenching through the reboiler 11 and the heat exchangers 11, 25, 16 and 6;
- partly , by pump 27, into the "hot oil" circuit for controlling, by the heat exchanger 19, the inlet temperature of the desulphurization reactor 22 and then, by pump 36, to the heating furnace 37 inlet again;
- partly, by pump 39, to the reactor 28 for its conversion to lighter products previous the addition with the treat gas w coming from recycle compressor 14 and preheated through the heat exchangers 25 and 26.

The special mechanical design of said phase separator 38, on the immediate downstream of the heating furnace 37, will allow to pump the effluent r notwithstanding its closeness to boiling conditions avoiding the need of a further quench, to prevent pump cavitation problems, and subsequent further heating to meet reactor 28 inlet temperature conditions.

The effluent from the reactor 28 is then mixed to the effluent from the reactor 22, as previously said, sent to preheat the feedstock in the heat exchanger 18 and then to the sequence of separators 23, 24 and 31 for the hydrogen recovery and products separation.

A partial recycle for a further conversion of the heavier fractions is thus obtained in the following way: heating furnace 37 - flash separator 38 - pump 39 - mildhydrocracking reactor 28 - heat exchanger 18 - series of separators in cascade 23, 24 and 31 - fractionator 20 - vacuum distillation column 10 - pump 36. Such partial recycle is directly conditioning overall conversion, products characteristics and related parameters (as hydrogen availability, space velocity, and so on).

As alternative to the above it is also possible to provide the effluent from the reactor 28 to be mixed with the charge feed of the reactor 22 instead of to be mixed with its effluent. This would allow to further desulphurize and hydrogenate the heavier fraction coming from the reactor 28.

Claims

1. Process for the hydrotreatment of both petroleum fractions obtained from atmospheric and vacuum distillation, which at present time are processed respectively in:
   - units for desulphurization, denitrification and moderate catalytic conversion and aromatic saturation of vacuum distillates only (with the exclusion of atmospherical ones),
   - units for desulphurization and denitrification of atmospheric distillates only (with the exclusion of vacuum ones),
   - units for desulphurization and denitrification processing alternatively vacuum or atmospheric distillates,

   provided with two separate reaction sections (22, 28) and tools for the segregation, down stream the first reactor (22) of the catalyst contaminant by-products (NH3, H2S) before entering the second reactor (28).
   characterized by that, within a single unit, desulphurization, denitrification and aromatic saturation are provided for both atmospheric and vacuum distillates, which are coprocessed together, as the catalytic conversion of the heavier fractions providing also a separate and indipendent control of the operating parameter of each treatment and consisting of:
   - two separate catalytic reaction sections (22, 28), one (22) treating all the atmospheric and vacuum distillates in the feedstock, the other (28) treating the residue (z) coming from the the vacuum distillation column (10),
   - a single heating furnace (37) ubicated between the two reaction sections (22, 28) operating in parallel on both sections,
   - a single product fractionation section consisting of atmospheric and vacuum distillation sub-sections,
   - a phase separator (38) to split the hot residue (z) coming from the furnace (37) into a gaseous stream (y), to be sent back to the vacuum distillation column, and a liquid heavier part to be charged to the hydroconversion reactor (28) and to be used to preheat the feedstock for the hydrodesulphurization reactor (22),

2. Process according to claim 1, *characterized* by that the conversion section (28) is charged with feedstock which has been previously desulphurized, denitrogenized and stabilized by vacuum distillation (22, 20, 10) for improving catalytic selectivity, conversion and performance stability.

3. Process according to claim 1, *characterized* by that both catalytic sections (22, 28) are integrated with a single two-staged product fractionation section, where the first stage is operating under pressure, for a further product purification from contaminants while the second stage is under vacuum for a deeper recovery of gasoil fractions.

4. Process according to claim 1, *characterized* by that a single furnace (37) provides all the necessary heat for the vacuum distillation column (20), for the hot-oil circuit which controls the inlet temperature at the desulphurization reactor (22) and for preheating the inlet stream to the conversion reactor (28).

5. Process according to claim 1, *characterized* by fact that the sole heating furnace (37) is fed with a hydrotreated feedstock.

6. Process according to claim 1, *characterized* by fact that it is provided the installation of a phase separator (38) arranged in such a way that the withdrawal of its liquid effluent r is made possible notwithstanding its closeness to

boiling state conditions.

7. Process according to claim 1, *characterized* by that it allows the utilization of a wide range of different catalysts including those, more selective, presenting tight feedstock quality constraints (as zeolites).

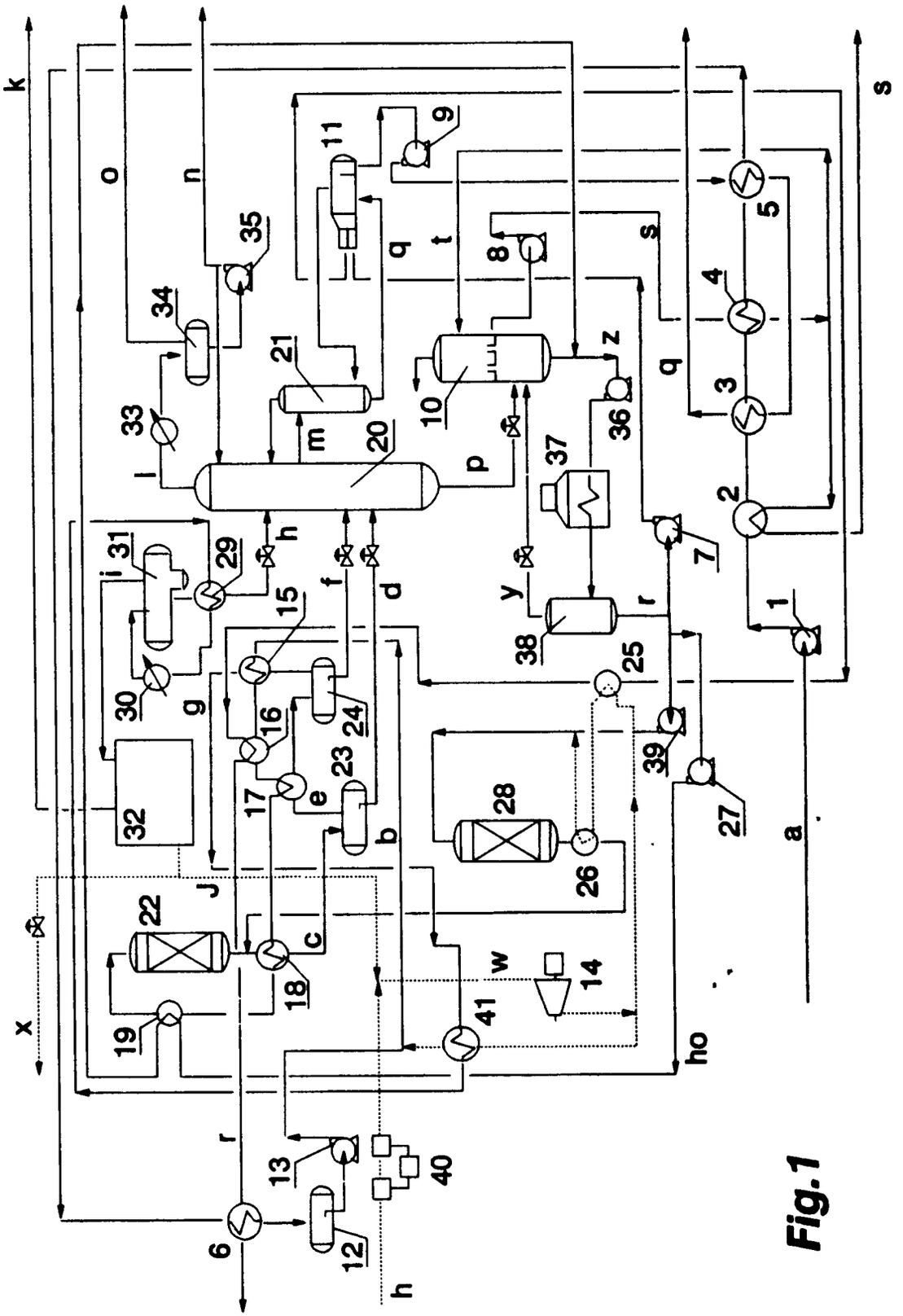

Fig.1

European Patent Office

EUROPEAN SEARCH REPORT

Application Number
EP 93 10 3451

DOCUMENTS CONSIDERED TO BE RELEVANT

Category	Citation of document with indication, where appropriate, of relevant passages	Relevant to claim	CLASSIFICATION OF THE APPLICATION (Int.Cl.5)
A	FR-A-2 433 044 (UOP) * claim 1 * * figure 1 * -----	1-3,7	C10G65/12
			TECHNICAL FIELDS SEARCHED (Int.Cl.5)
			C10G

The present search report has been drawn up for all claims

Place of search	Date of completion of the search	Examiner
THE HAGUE	29 October 1993	DE HERDT, O

CATEGORY OF CITED DOCUMENTS

X : particularly relevant if taken alone
Y : particularly relevant if combined with another document of the same category
A : technological background
O : non-written disclosure
P : intermediate document

T : theory or principle underlying the invention
E : earlier patent document, but published on, or after the filing date
D : document cited in the application
L : document cited for other reasons

& : member of the same patent family, corresponding document

EPO FORM 1503 03.82 (P04C01)