8.2.1 Base oil production technologies

As illustrated in Fig. 1 in Chapter 8.1, the base oils processing consists of vacuum distillation, deasphalting, aromatics extraction, dewaxing and finishing. It should be noted that during the following discussion reference will also be made to processes no longer used.

**Vacuum distillation**

In this case, the purpose of the vacuum unit is to prepare the feedstocks for subsequent treatments by removing heavier hydrocarbons, resins and asphaltenes, which are incompatible with the production of base oils and the performance of lubricating oils. It is thus designed in a different way from the fuels type vacuum unit, which manufactures vacuum gas oil for use as a feedstock in catalytic cracking, for the following reasons:

- It is important to avoid residua being entrained in the distillates, as they would lower yields from aromatics extraction and the filtration rate in dewaxing. Furthermore, they would shorten the life of catalysts (due to the increase in coke deposits) and decrease their selectivity (Soudek, 1974).
- It is desirable to obtain a selectivity as high as possible in the solvent aromatics extraction process. The lube vacuum unit produces gas oil overhead and 4 or 5 sidestream products. The heavier three distillates, together with the residuum, are generally processed to base oils. Basically, the vacuum gas oil is fractionated in the vacuum unit before and not after the aromatics extraction (Kosters, 1977).

Furthermore, unlike the fuels type vacuum unit, in the lube vacuum unit each sidestream product is steam stripped in a vacuum tower (stripper). These are located one above the other in a column placed next to the vacuum tower. This further improves the selectivity of aromatics extraction by solvents; the flammability of the base oils is controlled by removing their more volatile components.

**Deasphalting**

The use of a deasphalting plant for the precipitation of asphaltenes in base oils processing dates back to the 1930s and is needed to recover the oil in the vacuum residuum, as well as in the asphalt. This oil cannot be vacuum distilled as thermal cracking phenomena arise due to the high temperatures required.

However, this process is not always used: the vacuum residuum is not treated in a deasphalting plant if the asphaltenes content in the crude oil is very low, or if the amount of oil in the vacuum residuum is too low (with respect to the amount of asphalt) in order to be recovered economically. This is true of the Venezuelan naphthenic crudes used to produce naphthenic base oils from vacuum distillates and asphalts from the vacuum residuum. Furthermore, in the case of paraffinic-naphthenic crudes, the processing can be limited to the production of an extremely heavy fraction from the atmospheric residuum, which is nevertheless lighter than bright stock.

Liquid propane is the most commonly used solvent to deasphalt the vacuum residuum in lube processing, due to its selectivity in precipitating resins and asphaltenes because of its low molecular weight. Paraffins heavier than propane, such as butanes, normal pentane and normal hexane, are not used because they are less selective. As a result, the oil
yields are higher because they do not precipitate a part of the resins which must neither be present during subsequent treatments nor in the base oils.

Propane boils at −42.1°C and has a critical temperature and pressure of 96.8°C and 4.26 MPa. Between −40°C and +20°C liquid propane does not dissolve saturated hydrocarbons whereas between +40°C and +60°C saturated hydrocarbons, especially paraffins, are extremely soluble in propane. This leads to precipitation of the resins and asphaltenes, the efficiency of which increasing on raising the temperature. Above the critical temperature, propane causes all hydrocarbons to precipitate (Soudek, 1974).

The applications of propane for residuum deasphalting, aromatics extraction and dewaxing are based on these properties. Furthermore, towards the end of the 1970s, the recovery of propane from the extract under supercritical conditions was introduced in the deasphalting process. Instead of multiple effect flash evaporation, heat exchangers are used, where the extract is brought above the solvent critical temperature. The deasphalted oil separates out and is then fed to the stripper. This results in considerable savings, leading to the modification of numerous plants (Northrup and Sloan, 1996). Fig. 1 shows a simplified flow diagram of a deasphalting plant with solvent recovery under supercritical conditions.

The vacuum residuum is blended with the liquid propane and sent into the extraction tower at a temperature, e.g. of about 50-55°C; the internal of the tower consist of perforated trays or of a mixer with various disks rotating around a shaft in the centre of the tower (Rotating Disk Contactor, RDC). The RDC tower is also used for aromatics extraction (see below). In the earliest low capacity plants this operation was counter-currently carried out by means of mixer-settlers.

Inside the tower deasphalting is operated counter-currently: the lighter liquid propane has a density of 508 kg/m³ at 15°C and is fed to the bottom of the tower, just above the asphalt outlet. The propane rises dissolving the hydrocarbons and causing the resins and the asphaltenes to precipitate towards the bottom, in turn entraining some of the propane. The feedstock may be blended with liquid propane to regulate viscosity and its inlet is between half and two-thirds up the tower, depending on the resins and asphaltenes content.

In the upper part of the tower the temperature is regulated by steam condensation in internal coils, and in the lower part by injecting liquid propane at a level below the feedstock inlet and the bottom. For example, about 70-75°C top and 40-45°C bottom temperatures are attained: in addition to the difference in density,
the liquid propane also rises due to the effects of the temperature gradient between the bottom and the top of the tower. Increasing the top temperature, however, tends to decrease oil yields because the propane’s tendency to precipitate all hydrocarbons is increased. The downward flow of propane together with resins and asphaltenes can be considered an internal reflux within the tower, as in distillation. The operating pressure is such as to maintain the liquid phase.

The extract, i.e. the liquid propane containing the dissolved deasphalted oil, is withdrawn from the top of the tower; the extract may contain, e.g. about 80-85 wt% of propane and 15-20 wt% of oil. The bottom product, i.e. the raffinate, contains about 65-70 wt% of asphalt and 30-35 wt% of propane, making it important to operate with a fairly high propane-feedstock ratio.

The propane must be recovered and recycled: the raffinate is first flash evaporated and then vacuum steam stripped to remove any traces of propane. After the water has been removed by cooling and flushing the vapours, the gaseous propane is liquefied by compression and cooling. The same procedure is used for the extract; in old plants, due to the larger amounts of propane, the multiple effect flash evaporation may entail two or three stages at decreasing pressure, not shown in Fig. 1. Due to the selectivity of this process, the asphalt produced is very hard, and represents a base for the manufacture of other asphalts.

The operating temperature, the propane-feedstock ratio and the yields vary depending on the type of crude oil. For example, DeAsphalted Oil (DAO) yields of about 30-40 wt% of residuum can be obtained (Normand, 1973; Soudek, 1974).

The DuoSol process, which was also used for aromatics extraction, was a widely employed deasphalting process in the past. In this case, liquid propane is used for deasphalting in combination with selecto, i.e. a blend of phenol and cresylic acid, in turn a mixture of ortho-, meta- and para-cresol, which simultaneously deasphaltizes the oil. Extraction is operated counter-currently, with the propane being injected into the bottom of the tower, the selecto at the top and the feedstock at a point in between. Solvent recovery is carried out in a similar way the propane recovery described above, but in this case supercritical conditions are not adopted for the propane, whereas for the selecto a more complex system is needed, since it is in turn a mixture of solvents.

The DuoSol process is more complex to operate given the use of a solvent system, and is less easy to convert to large dimensions than the propane process; the latter is preferred also because the phenol and the cresols represent a potential hazard due to their toxicity (Normand, 1973).

Aromatics extraction

Aromatics have a negative impact on the properties of base oils and their removal improves the quality (Normand, 1973; Billon et al., 1975; Kosters, 1977; Kramer et al., 1999a; Kramer et al., 1999b; Werner et al., 2001). More specifically, aromatics extraction:

- Increases the viscosity index (the higher the viscosity index, the lower the variation of viscosity with temperature).
- Increases the resistance to the oxidation from dissolved air in the lubricating oil and contact with its surface. Oxidation leads to the formation of acid compounds, which are corrosive as well as lighter and more volatile; as oxidation progresses, it leads to the formation of solid particles in suspension, of deposits on the bottom and carbonaceous resida on the surfaces. Oxidation is accelerated by increasing temperature and catalysed by metal compounds, generally in the form of soluble complexes in the oil, albeit in very small quantities.
- Increases the thermal stability. Over time, under the action of heat, the molecules must not crack or condense, leading to the formation of volatile materials, suspensions, deposits and carbonaceous resida on the surfaces. The effects of lowered stability may compound the effects of oxidation.
- Increases the surface tension, leading to a reduction in the foaming tendency.
- Decreases the volatility (aromatic molecules are more volatile than saturated hydrocarbons). In the engines of motor vehicles the quantity of oil vapours burnt together with the fuel in the combustion chamber thus decreases. This lowers the consumption of lubricating oils and the formation of toxic emissions, especially particulate, lowering environmental pollution and increasing the life and efficiency of catalytic converters.

Sulphuric acid treatment was the first process used for aromatics removal and later abandoned since the acid preferentially reacts with the most oxidizable compounds, in other words the aromatics, leading to low selectivity and the formation of acid sludges requiring disposal (Nelson, 1958). Today, aromatics removal is performed by solvent extraction processes (see again Fig. 3 in Chapter 8.1), which are the most widespread, by hydrotreating (see again Figs. 4 and 5 in Chapter 8.1) and by hydrotreating.

In this case, the hydrotreating process (see Chapter 3.1) can be placed before or even after
aromatics solvent extraction and saturates the aromatics in the vacuum distillates and the deasphalted oil for the production of group I oils and group II medium and heavy base oils. Operating conditions are milder than in hydrocracking, but more severe than in hydrofinishing. The saturation of the aromatic hydrocarbons removes some of the sulphur and nitrogen in the molecules (in the form of hydrogen sulphide and ammonia) and produces paraffinic base oils with a higher paraffinic content than solvent extracted oils. The base oil yield and also the crude oil flexibility increase (Arnold, 1997). Hydrotreating has largely replaced aromatics extraction in the case of naphthenic oils processing. The saturation of aromatics by hydrogenation is also the current process for producing technical and medicinal grade white oils from naphthenic crudes (see below).

With respect to hydrotreating, hydrocracking (see Chapter 6.2) entails greater crude oil flexibility and produces higher quality base oils including solvent dewaxed group II oils and group III oils from catalytic dewaxing. In comparison to solvent extraction and hydrotreating processes, hydrocracking produces more base oils with a higher paraffinic oil content, because the aromatic hydrocarbon content is lowered more significantly, resulting in both higher quality levels and improved performance of lubricating oils. Furthermore, hydrocracking can be better integrated with fuel production since it allows an increase in refinery yields of gasoline, kerosene, jet fuel and gas oil. The discussion below will thus be restricted to a description of solvent extraction and hydrocracking technologies.

**Solvent extraction**

Solvents with strong polarity (due to the asymmetry of their molecular structure and the presence of a heteroatom, such as oxygen or nitrogen) easily dissolve polar substances (such as aromatic hydrocarbons) but only very partially dissolve weakly polar substances (such as saturated hydrocarbons).

When the oil to be dearomatized is mixed with the solvent at constant temperature and a sufficient pressure to maintain the system in the liquid phase, at equilibrium the separation into two liquid phases is observed. The heavier phase (extract) consists mainly of aromatic hydrocarbons and most of the solvent; the lighter phase (raffinate) consists largely of the saturated hydrocarbons and a smaller proportion of the solvent. As the temperature increases, the two-phase region gradually disappears (Treybal, 1955). Extraction is a continuous counter-current operation; the feedstock is injected into the middle of the tower and may already be partially blended with the solvent. The solvent enters at the top of the tower, and flows towards the bottom due to the action of gravity, dissolving the aromatic components, while the lighter oil rises from bottom to top together with a small quantity of the solvent. In extraction a temperature gradient is maintained between the bottom and the top of the column. Temperature is kept as high as possible in the top of the tower, since reducing the two-phase region makes it possible to lower the solvent-oil feed ratio, while a lower temperature in the bottom of the tower facilitates the separation of the extract and raffinate phases.

The physical properties of the solvents have a well-defined technological purpose: the high density facilitates the separation of the raffinate and extract phases, the high boiling point makes it possible to use lower pressures to prevent vaporization phenomena inside the extraction tower, and the low latent heat facilitates the recovery of the solvent by distillation. By contrast, as in the case of furfural and phenol, the presence of azeotropes with water does not facilitate recovery.

The highest solvent-feedstock ratios and the highest top temperatures are reached in the treatment of the heaviest or most aromatic feedstocks, or when higher viscosity indexes are required. On the other hand, a low solvent-feedstock ratio reduces the size of all of the equipment and the operating costs.

A single solvent (this category includes mixtures which behave as a single solvent in extraction), or a binary system (in which two solvents carry out different actions) may be used. Binary solvents and more complex mixtures are considered obsolete, since they entail more complex and thus more expensive processing to recover the raffinate and the extract, especially when production capacity is high. Currently, furfural, N-methyl-2-pyrrolidone (NMP) and phenol are the most widely used solvents:

```
CH3 O
H
C ≈ O
```

furfural

```
N-methyl-2-pyrrolidone
```

phenol

New plants, however, use only furfural and N-methyl-2-pyrrolidone, with the latter usually being preferred. NMP is increasingly used due to its higher selectivity and lower toxicity with respect to phenol; in the United States and Canada all existing phenol plants have been converted to N-methyl-2-pyrrolidone. It should be remembered that replacing one solvent with another requires verification of the performance of the lubricating oils. **Fig. 2** shows a simplified flow diagram of the furfural process.
Furfural is a cyclic aldehyde, which is unstable and tends to decompose above 220°C. Below this temperature, it is oxidized by air to form corrosive acids and tends to polymerize; this in turn leads to the formation of deposits, which become concentrated in the extract. The feedstock is therefore vacuum deaerated and the furfural is kept under nitrogen blanketing in storage tanks.

The first furfural extraction columns were packed towers; later the increase in capacity led to the use of far more efficient mixing methods, which were thus more suitable for increasing the capacity of plants. To this end, as in propane deasphalting, the RDC tower is adopted. This consists of a series of compartments between stator rings, each containing a disk rotating around a shaft in the centre of the tower (see again Fig. 2).

The top temperature is regulated by the flow and the temperature of the solvent, while the bottom temperature, where the separation of the extract occurs, is controlled by the flow and the temperature of an external reflux, which is withdrawn from the distributor between the extraction zone and the settler of the extract phase and injected into the bottom of the tower.

To recover the furfural, the raffinate is stream stripped in a vacuum tower, in the lower part of which (the stripper) the last traces of furfural are removed from the paraffinic oil; at the top the azeotrope is distilled off and the furfural reflux is injected. After the steam stripping of the furfural in medium and high pressure towers, the extract passes into a final vacuum tower, similar in design to the column for the raffinate treating, where the azeotrope is distilled off overhead. The azeotrope is also distilled off in the medium pressure tower, whereas furfural is obtained overhead from the high pressure tower and recycled after condensation.

By settling at room temperature and pressure all the combined azeotropes results in the 96% furfural-rich phase and the 92% water containing phase. The latter is distilled in a tower where an azeotrope is produced overhead and recombined with the others in the settler while waste water is drained from the bottom. The furfural-rich phase feeds medium pressure distillation, from which the dried furfural is extracted for recycling (Nelson, 1958; Normand, 1973; Soudek, 1974; Kosters, 1977).

As far as phenol extraction is concerned, it should be remembered that this solvent does not oxidize and is stable. The heated feedstock absorbs the phenol in the steam from the vacuum stripping towers where the solvent is recovered from the raffinate and the extract, and where the azeotrope is distilled off overhead. After absorption, the waste water is subjected to specific treatments to remove the remaining phenol.

The cooled feedstock contacts the phenol in a counter-current operation. Perforated tray towers, RDC and Podbielniak extractors are used: the latter are small horizontal centrifuges, in which the counter-current flow of the light and heavy phases is ensured by centrifugal force.

**Fig. 2.** Furfural extraction process (Normand, 1973).
The phenol also dissolves hydrocarbons usually present in the raffinate and entrains them in the extract; to avoid this, water is injected into the extract at the bottom before its outlet from the extraction tower. A pseudo-raffinate phase is separated, recycled and reextracted. A similar effect is obtained by lowering the bottom temperature.

The raffinate passes first into an evaporator, where it releases some of the phenol, which is condensed and recycled, and then into a vacuum steam stripper, from which an azoetrop is distilled off overhead, whereas the waxy base oil is withdrawn from the bottom. The extract is treated first in a two-stage flash evaporator, where the phenol is removed, and then in a vacuum steam stripper, where an azoetrop is distilled off overhead, and the extract is obtained as the bottom product (Nelson, 1958; Soudek, 1974; Kosters, 1977).

As far as N-methyl-2-pyrrolidone extraction is concerned, it should be remembered that this solvent is extremely stable, does not form azoetropes with water, and is more selective for aromatics than the other two solvents; as a consequence, lower solvent-feedstock ratios are needed. When NMP replaces furfural or phenol and an equal quantity of feedstock is treated, much less solvent thus needs to be regenerated, hence lowering operating costs.

The NMP extraction is counter-currently operated in perforated tray columns. The solvent recovery from the raffinate consists of an evaporator followed by the vacuum steam stripping of the raffinate. The vapours from the evaporator and the stripper are condensed by cooling and recombined to separate out the water.

For the extract a multiple effect flash evaporation is required, at high and medium pressure, and finally under vacuum; the extract is then sent to another steam stripper where the last traces of solvent are removed. The vapours are condensed and the water removed. Traces of water remain in the solvent, which must therefore be dried by vacuum distillation. Another possibility, however, is to inject wet NMP, or even water, into the feed to control the solubility of the hydrocarbons (Refining [...], 2002).

It is thus easier to convert a phenol plant than a furfural plant for use with NMP due to the differences in the solvent recovery systems.

As far as the nowadays obsolete solvents are concerned, only the binary systems most widespread in the past will be mentioned (Nelson, 1958; Normand, 1973; Kosters, 1977):

- Liquid sulphur dioxide, with the addition of benzene because the dioxide is overly selective for aromatics, and barely dissolves naphthenes, whereas benzene dissolves all types of hydrocarbons. This process is highly suited to the treatment of naphthenic oils, for which benzene is not used, and was the first to be used for aromatics extraction.
- Selecto and propane (DuoSol process, already mentioned for deasphalting); the selecto behaves as a single solvent, descending through the column and extracting the aromatics whereas the liquid propane rises through the tower, prevalently dissolving saturated hydrocarbons.

**Hydrocracking**

Hydrocracked base oils are manufactured by means of three basic methods of processing:

- The hydrocracked atmospheric residuum from a high conversion unit for fuel production is refined for base oils production.
- The same feedstock for catalytic cracking (i.e. a blend of vacuum gas oil and deasphalted oil) is sent to low once-through hydrocracking in order to maximize base oil yield (bulk-feed operation). The resulting hydrocracked atmospheric residuum is fractionated into light, medium and heavy waxy distillates and a vacuum waxy residuum, which are sent to further processing.
- The individual vacuum distillates and deasphalted oil are sent to once-through hydrocracking under the most appropriate operating conditions for each feedstock in order to maximize the yield and the quality of each base oil (blocked operation).

The selection within these possible uses of hydrocracking depends on numerous factors, of which the most important are: a) the availability of hydrocracked residua (also from other refineries); b) the material and economic balance of the treatments; c) the quality and the yields of base oils; d) the market demand for fuels and base oils.

The quality of the base oils, their yields and the crude oil flexibility of the plants differ, depending on the catalysts, the reaction mechanism and whether the processing mode is a bulk-feed operation or a blocked operation.

The catalysts consist of metal oxides with a hydrogenating-dehydrogenating activity, belonging to groups VIB (usually molybdenum and tungsten) and VIII (including cobalt, nickel and the noble metals, such as platinum and palladium) on an acid amorphous or zeolyte-based support. The zeolyte-based catalysts have a clearly defined mean pore structure and size, and a high density of acid sites; as a result, less coke is formed and their cracking activity is greater when compared with the amorphous catalysts (Corma et al., 1997).

If hydrocracking is intended to maximize gasoline, kerosene, jet fuel and gas oil, the conversion is to be maximized by recycling the hydrocracked atmospheric...
residuum at more severe operating conditions. This entails the need to use catalysts which favour the production of light and medium distillates and simultaneously exhibit low coke formation, such as zeolite catalysts.

Noble metals are used to maximize gasoline yields because the increase in cracking activity of the matrix must be matched by the increase in the hydrogenating-dehydrogenating activity of the metals.

On the other hand, if the production of gas oil and especially manufacture of lube vacuum distillates are to be maximized, the operating conditions must be milder and the conversion must be low: in the lube production it is not necessary to recycle the hydrocracked atmospheric residuum.

Cracking reactions must be limited because the hydrocracked atmospheric residuum yield must be high; efforts are also made to keep the paraffins within the distillation interval of the base oils as their most valuable components. Both amorphous matrices consisting of alumina or silica-alumina, and metals with a lower hydrogenating activity are thus suited to the production of base oils; catalysts are used containing cobalt-molybdenum on alumina or nickel-molybdenum on alumina, and nickel-tungsten on silica-alumina or on alumina.

The metal oxides are activated in the reactor by a hydrogen sulphide treatment which converts them into sulphides; in the plant operation an equilibrium is established between the metal sulphides and the hydrogen sulphide in the gas from the desulphurization of the feedstock. Fig. 3 shows a simplified diagram of the reactions in the order in which they occur, the corresponding variations in structure of typical molecules and the effect of the chemical transformations on the direction of the change of the desired properties in the base oils.

The metal sulphides catalyse the fastest reactions, i.e. the hydrogenation of the aromatic rings and of the olefinic intermediates, from dehydrogenation (the products contain only traces of olefins). As a consequence, the remaining heteroatom (i.e. sulphur, nitrogen and oxygen) of the hydrotreated feedstock is hydrogenated, leading to formation of hydrogen sulphide, ammonia, and steam. Dehydrogenation reactions also lead to the formation of coke deposits on the catalyst.

The acid matrix catalyses all of the other reactions, of which only the most important for the production of base oils are listed, in their hypothetical order of occurrence. In polycyclic aromatic hydrocarbons the aromatic rings are hydrogenated one after the other to naphthenic rings. The naphthenic rings then open (decyclization) and the alkylic free radicals become detached from the rings (dealkylation), forming isoparaffins and normal paraffins which undergo isomerization reactions and cracking to products with a low molecular weight. The acid matrix also catalyses condensation reactions in the aromatic rings, leading
to the formation of coke and releasing hydrogen. Further coke is formed by the thermal route, because the operating temperatures promote dehydrogenation and condensation reactions.

The desired properties of the base oils improve both with the increase in hydrogenation and ring opening (up to the formation of monoalkyl naphthenes) and with the increase in the quantity of paraffins (especially normal paraffins). The concentrations of the various hydrocarbon classes, however, are not uniform as the boiling point varies: specifically, aromatics are more concentrated in light and medium hydrocracked distillates, which therefore get to lower viscosity index, oxidation and thermal stability than heavy hydrocracked distillates and bright stocks, and to potentially excessive volatility.

Table 1 shows the yields and the properties of dewaxed base oils produced by bulk-feed hydrocracking in the refining of good Middle Eastern lube crude.

<table>
<thead>
<tr>
<th></th>
<th>Vacuum gas oil</th>
<th>Dewaxed base oils from hydrocracked vacuum gas oil</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Feed Hydrocracked 150 N 500 N BS 150</td>
<td></td>
</tr>
<tr>
<td>Yields after dewaxing (% by weight)</td>
<td>– 66.0</td>
<td>33.0 28.0 5.0</td>
</tr>
<tr>
<td>Viscosity index</td>
<td>62 111</td>
<td>102 111 111</td>
</tr>
<tr>
<td>Density at 15°C (kg/m³)</td>
<td>931 870</td>
<td>869 870 870</td>
</tr>
<tr>
<td>Viscosity at 98.9°C (cSt)</td>
<td>20.5 8.4</td>
<td>5.2 12.0 32</td>
</tr>
<tr>
<td>Pour point (°C)</td>
<td>+21 –21</td>
<td>–21 –21 –18</td>
</tr>
<tr>
<td>Composition (%):</td>
<td></td>
<td></td>
</tr>
<tr>
<td>– carbon in aromatic rings</td>
<td>21.4 3.5</td>
<td>5.2 1.2 0.5</td>
</tr>
<tr>
<td>– carbon in paraffinic chains</td>
<td>51.9 69.7</td>
<td>64.6 72.7 74.0</td>
</tr>
<tr>
<td>– carbon in naphthenic rings</td>
<td>26.7 26.8</td>
<td>30.2 26.1 25.5</td>
</tr>
</tbody>
</table>

The hydrocracking reaction is exothermic because the effects of hydrogenation prevail over the endothermic cracking process. An increase in temperature promotes dehydrogenation and cracking reactions, increases conversion to light products, but not hydrogenation reactions and favours the coke formation on the catalyst which decreases the life of the catalyst. A decrease in space velocity increases conversion, promoting the formation of coke to a lesser extent. Due to the greater complexity of the reactions, the operating pressures and temperatures are higher than in subsequent catalytic hydrogenation processes. The hydrocracking process can follow various possible flow diagrams (see Chapter 6.2).
Dewaxing

Following aromatics extraction, the paraffinic base oils are in the solid or semi-solid state at room temperature and pressure due to their normal paraffin content; this must therefore be reduced in order to allow the oil to flow up to a predetermined temperature (pour point).

Normal paraffins, like aromatics, cannot be removed by vacuum distillation because they are distributed throughout the distillation interval: solvent and catalytic hydrogenation processes must therefore be used for dewaxing. However, the necessary reduction in the normal paraffin content worsens the quality of the base oils because these paraffins are the hydrocarbons with the highest viscosity index, the highest oxidation and thermal stability and the lowest volatility. Therefore, the preferred processes do not remove the normal paraffins, but convert them into isoparaffins by hydroisomerization, maximizing their concentration in order to increase the oil yield and raise the quality of the base oils.

Isoparaffins improve fluidity due to their low pour point, also facilitating the cold starting of engines. They also have the best stability and volatility properties though to a lower extent than normal paraffins: from a qualitative point of view, an ideal base oil should consist exclusively of isoparaffins. Essentially, the hydroisomerization process completes the action of the prior hydrocracking (see again Fig. 4 in Chapter 8.1) or is carried out on the deoiled waxes resulting from the increase in the viscosity index, i.e. the flow rate of the filtrate (blend of oil and solvent) per surface area unit of the filter (in m³/hm²), μ is the absolute viscosity of the filtrate at the filtration temperature, ΔP is the pressure differential between the oil to be filtered and the filtrate and R is the sum of the resistances of the filter and the layer of wax deposited on the filter (known as cake), which retains oil and solvent (Brown et al., 1966; Normand, 1973).

This equation shows that the higher the viscosity (at room temperatures) of the oil to be filtered, the larger the quantity of solvent which must be added in order to obtain a filtrate viscosity such as to allow for an acceptable filtration rate at the filtration temperature. Obviously the wax must form sufficiently large crystals in order to retain as little oil and solvent as possible and the solvent must have low viscosity even at low temperatures.

In order to define the structure of the wax crystals the slope of the chilling temperature-time curve and the agitation in the mixture of the solvent and waxy oil are extremely important.

It also appears that the filtration rate of hydrocracked distillates decreases both due to the larger quantity of isoparaffins and alkyl naphthenes resulting from the increase in the viscosity index, and due to the microcrystalline waxes from the cracking of the higher molecular weight wax in the heavy distillates and the deasphalted oil (Arnold, 1997). The capacity of a solvent dewaxing plant may thus be lowered, with a consequent decrease in the production of base oils and waxes.

In the cake the motion of the oil is laminar. Given an identical pressure differential, as filtration progresses, the cake thickens and becomes more compact, thus increasing the resistance to motion and reducing the filtration rate until it becomes almost zero.

Solvent dewaxing

A solvent dewaxing and separation stage involves diluting the waxy oil with suitable solvents, chilling the mixture to a temperature below the pour point to induce the formation of wax crystals, and separating the crystals by means of a filter. The solvent is recovered and recycled. After the recovery of the solvent, the wax is a blend of normal paraffins, isoparaffins and oil, with a colour ranging from yellow to dark brown.

It is necessary to remove only such a quantity of normal paraffins that the oil remains in the liquid state up to a given pour point. If it is intended that the oil remains liquid at even lower temperatures, the treatment must be more severe, removing a larger quantity of normal paraffins.

For all types of oil to be dewaxed, the filtration rate is described by the equation:

\[
\frac{dQ}{dt} = \frac{\Delta P}{\mu R}
\]

where \(dQ/dt\) is the mean linear velocity of the filtrate, i.e. the flow rate of the filtrate (blend of oil and solvent) per surface area unit of the filter (in m³/hm²), \(\mu\) is the absolute viscosity of the filtrate at the filtration temperature, \(\Delta P\) is the pressure differential between the oil to be filtered and the filtrate and \(R\) is the sum of the resistances of the filter and the layer of wax deposited on the filter (known as cake), which retains oil and solvent (Brown et al., 1966; Normand, 1973).

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It also appears that the filtration rate of hydrocracked distillates decreases both due to the larger quantity of isoparaffins and alkyl naphthenes resulting from the increase in the viscosity index, and due to the microcrystalline waxes from the cracking of the higher molecular weight wax in the heavy distillates and the deasphalted oil (Arnold, 1997). The capacity of a solvent dewaxing plant may thus be lowered, with a consequent decrease in the production of base oils and waxes.

In the cake the motion of the oil is laminar. Given an identical pressure differential, as filtration progresses, the cake thickens and becomes more compact, thus increasing the resistance to motion and reducing the filtration rate until it becomes almost zero.
The importance of the type of solvent and the filtration technology has been clear since the use of the first solvents, i.e. gas oil for vacuum distillates and virgin naphtha for heavier distillates, bright stock and cylinder oil. In this case, the evaporation of liquid ammonia chills the blends in order to obtain wax crystals, which must be separated in a different way.

Light and medium distillates are filtered in filter presses, formed of up to a hundred plates. In each plate a mesh retains the slack wax (so-called because it is an intermediate and not the finished commercial product) which in turn retains gas oil and is discharged onto a conveyor belt after filtration is complete. The dewaxed oil is sent for finishing after the gas oil recovery by steam stripping.

The heavier distillates and intermediate oils from residues, by contrast, cannot be treated in this way since the wax crystals are small. This wax is described as microcrystalline (ceresin) and, unlike macrocrystalline wax, hinder filtration by retaining the oil in the cake: centrifuging is required to separate the oily wax (known as petrolatum). The virgin naphtha is recovered by vacuum distillation and the dewaxed oil and the petrolatum are sent for finishing.

Furthermore, unlike slack wax, petrolatum cannot be deoiled by heating (this operation is known as sweating). For deoiling, slack wax is solidified and is heated in an oven at constant temperature to about 40-60°C. An oil is collected below (known as foots oil) up to about 40°C and is generally recycled to pressing, whereas the oil collected at higher temperatures is dewaxed. This results in a wax which has not been selectively deoiled, known as scale wax, which is sent for finishing (Nelson, 1958).

The processes described above are now obsolete because they are expensive (pressing is also a batch operation), entail low selectivity, and are unsuited for upscaling to large capacities. More recently, new types of solvent have been used. Filtration has also been rendered a continuous operation, and the thickness of the cake has been limited by means of Oliver type rotary vacuum filters (Nelson, 1958; Brown et al., 1966; Normand, 1973). Below, the solvents and processes currently of greatest importance will be described.

Benzene is not used because it is hazardous; toluene is an extremely good solvent for paraffins and an excellent solvent for other hydrocarbons, but cannot be used on its own because the oil mixture would have to be chilled to excessively low temperatures with respect to the desired pour point of the oil; the wax would also crystallize in such a way as to impede filtration.

Alkyl ketones, including methyl ethyl ketone (MEK) and methyl isobutyl ketone (MIBK), known as antisolvents, do not efficiently dissolve oil, and precipitate wax extremely well in a way which allows for filtration and the retention of only a small quantity of oil. Antisolvents attenuate the solvent properties of toluene, thus approaching the ideal solvent.

Although propane is successfully used on its own, the best results are obtained by means of a mixture of two solvents. Currently, the solvents generally used are blends of MEK and toluene, the most widespread, and MIBK with MEK, the latter being composed exclusively of ketones, but for which special crystallization technology is adopted:

\[
\text{MEK} \quad \text{MIBK} \quad \text{toluene}
\]

**Fig. 4** shows an example of phase equilibria in MEK/toluene dewaxing, using a solvent-dewaxed to −15°C oil ratio of 4/1 by volume. With 20 vol% of MEK in the solvent, the cloud point (i.e. the temperature at which wax crystals form in the liquid) is −18°C; point A corresponds to 30 vol% of MEK in the solvent and a temperature of −15°C: at A the waxes and oil precipitate together, and this is the desired pour point for the oil. For concentrations of MEK above 30 vol%, the oil is precipitated, entraining all the wax, e.g. at −4°C with 50 vol%, and the oil cannot be dewaxed.
The cloud point corresponds to the filtration temperature, which may be lowered by increasing the solvent/dewaxed oil ratio and decreasing the concentration of MEK in the solvent. This type of diagram is also valid for other binary solvents, including MIBK/MEK (Normand, 1973).

The stages in an MEK/toluene dewaxing, deoiling and fractionation plant are shown in Fig. 5. In the dewaxing stage, the solvent is usually added at the beginning and the end of chilling, when the filtration temperature is reached. In the case of bright stock, the quantity added at the beginning is larger, in part because in this way the filtration is facilitated by the structure of the crystals. By sending in the solvent at various points during chilling, the solvent-feedstock ratio is lowered, and the filtration rate is increased. The deposition of wax crystals on the exchangers is prevented by scraping their surface and returning the crystals to the fluid mass.

The blend of oil, solvent and wax crystals is filtered by means of Oliver type filters, which are partially immersed in the liquid and maintained at low pressure by flue gas blanketing. The slack wax is deposited on the filter, whose surface is subdivided into panels where a vacuum is formed to suck out the filtered oil and solvent. As the filter rotates, the slack wax cake is first washed with cold solvent, sucked up with the oil, and then detached by flue gas blowing and, at the end of the circuit, is scraped with a blade for discharge into a collecting screw channelling it outside.

The solvent is recovered from the oil by means of a multiple effect evaporation system composed of two or three flash evaporators, and a final vacuum steam stripper; after drying, it is recycled to minimize the formation of ice due to the water used in distillation, while the dewaxed oil is sent for finishing.

After the dewaxing stage, the slack wax can be sent for the removal of the solvent by distillation and a final steam stripper; alternatively it may be sent to a second stage consisting of deoiling, in which it is diluted with solvent, chilled, filtered and washed in Oliver type filters. The filtrate and the solvent are recycled to the chilling section of the first stage. The deoiled wax (known as scale wax) is freed of the solvent in the same way as slack wax, or sent to a third stage involving fractionation, if it is in operation. In this case, the scale wax is diluted with a solvent at a higher temperature, filtered and washed to collect as much as possible of the normal paraffins on the filter: the result is a cake consisting of solvent and a wax, known as hard wax due to its high normal paraffin content and very low oil content.

The flushing solvent is mixed with the filtrate (i.e. a soft wax with a high oil content and containing solvent) which is collected and washed during subsequent filtration at a lower temperature. The filtrate, not shown in Fig. 5, is recycled to the first stage. The solvent is recovered from the soft wax and hard wax in a similar way to the other stages. The waxes are then sent for finishing. In the third stage, the petrolatum may also be fractionated after deoiling (Normand, 1973).

In MIBK/MEK dewaxing, the crystallization technology is different: the formation of small crystals
which would obstruct filtration is avoided by gradually adding the cold solvent to the waxy oil in a crystallization tower, subdivided into stages where the mixture is chilled and agitated vigorously. The wax crystallization largely occurs in the tower, and continues in scraped surface exchangers.

It is not necessary to add new solvent for deoiling, but it is necessary to add solvent at a higher temperature before the separation into soft and hard wax. Also in this case Oliver type filters are used for filtration; the recovery and the drying for the two solvents are similar to the MEK/toluene process (Citarella et al., 2000; Refining [...], 2002).

A recent technological development consists in the use of polymeric membranes to recover the solvent from the oil and the soft wax.

The liquid propane dewaxing process dates to 1932 and is still operated today. Any feedstock can be dewaxed with the advantage of using the same solvent as deasphalting. The peculiarity of the process is in the evaporation of some propane in order to chill its mixture with the waxy oil; temperatures as low as about −40°C can be attained. Wax forms crystals which are filtered on an Oliver type filter, maintained at low pressure, and washed with liquid propane. The propane is then recovered from the wax and the filtered oil by evaporation and vacuum distillation and returned to the liquid state by compression and cooling of the gases (Nelson, 1958; Normand, 1973).

**Catalytic dewaxing**

The base oils are dewaxed by means of two different catalytic hydrogenation technologies.

In the first the normal paraffins are cracked in order to lower the pour point, wax is not manufactured and the viscosity index decreases by about 7-8 points. This processing produces group I, II and III base oils depending on the quality of the feedstock.

In the second, by contrast, cracking is kept to a minimum, and the normal paraffins in waxy oils and waxes are hydroisomerized to isoparaffins in order to decrease the oil’s pour point, improve cold starting, increase the viscosity index and improve all of the properties, sometimes considerably. Group I, II and III base oils can be produced from waxy oils, however this process is generally used to produce group II and III base oils.

The processes to crack normal paraffins and hydroisomerize waxy oils may be used to increase the capacity of a solvent dewaxing plant, to be operated under conditions of partial dewaxing, later completed in catalytic plants with an increase in the quality level of the base oils. The wax hydroisomerization is a process which converts wax into group III base oils with an extremely high viscosity index (up to 150 and above). In the case of crude oil refining, this technology produces the highest quality base oils, which exhibit the lowest volatility, the highest thermal and oxidation stability and the best ease of starting up engines. It is also possible to use treated Fischer-Tropsch wax as a feedstock for hydroisomerization.

The feedstocks for catalytic hydrogenation processes must be as free as possible from molecules containing sulphur and nitrogen, which reduce the life and the selectivity of the catalyst. The feedstock must also be free from aromatics and from intermediate olefinic products of the hydrocracking reactions. These molecules are adsorbed on the active sites of the catalyst, where they are hydrogenated and form coke by dehydrogenation and cyclization, as well as by thermal route, reducing the life of the catalyst and making a hydrotreating process upstream necessary.

The catalytic dewaxing is based, as hydrocracking, on the use of bifunctional catalysts, consisting of two classes of components with very different functions. The hydrogenating-dehydrogenating function is generally ensured by the noble metals of the VIII group (such as platinum and palladium), while the isomerization and cracking reactions are catalysed by the acid matrix, consisting of silica-alumina. This matrix may be either amorphous or based on zeolites and molecular sieves.

The initial dehydrogenation and skeletal hydroisomerization are fast reactions and dominate over cracking if the contact times on the catalyst are short (i.e. the space velocity is high). This increases the concentration of isoparaffins in the product; however, if the residence time on the catalyst becomes sufficiently high, cracking dominates, and the concentration of isoparaffins decreases after passing through a maximum.

Furthermore, if the feedstock consists not of a fully deoiled wax (composed essentially of normal and isoparaffins), but of an oily wax or a hydrocracked waxy effluent, a lower quantity of isoparaffins in the oil is obtained (Bellussi et al., 1997-1998). Other conditions being equal, the oil content lowers the quantity of paraffins in the feedstock and the naphthenes and the aromatics in the oil are adsorbed onto the active sites, where they undergo the typical hydrocracking reactions.

The highest viscosity index is reached if the feedstock is completely deoiled. Furthermore, the isoparaffins produced must exhibit a single long side chain in the centre of the molecule. For a given number of carbon atoms in an isoparaffin, the isoparaffins with numerous short side chains are undesirable because they have a lower viscosity index, which decreases as the number of these chains increases.
The wax hydroisomerization (see again Fig. 5 in Chapter 8.1) makes it possible to produce group III base oils with adequate yields, i.e. about 50-60 wt% of the feedstock (Helton et al., 1998). In order to obtain this yield from the catalysts the extent of the cracking reactions is reduced as much as possible by means of low acidity amorphous matrices, which include alumina or silica-alumina, or by the use of molecular sieves containing phosphorus, or of zeolites. This function is balanced by the amount of noble metals used; sometimes fluorine is used as a promoter.

Fig. 6 shows the simplified flow diagram for the wax hydroisomerization process. If the highest viscosity indexes are desired, it is necessary to distinguish untreated slack wax, scale wax and soft wax feedstocks from hydrogenated hard wax and scale wax feedstocks. These categories also include waxes from the solvent dewaxing of bright stock. The former (dotted line) must be hydrotreated, sometimes following solvent deoiling if the oil content is too high with respect to the desired viscosity index. The latter, on the other hand, like the oil-free treated Fischer-Tropsch waxes, can be fed directly to the hydroisomerization reactor.

Overall, the reaction is exothermic because hydrogenation reactions dominate over the endothermic cracking reactions; it is therefore unfavoured by an increase in temperature. Operating conditions such as about 315-400°C as reaction temperatures and above 5.0 MPa pressures are used (Everett and Suchanek, 1996). The hydroisomerization reactor is followed by the final hydrotreating reactor, where among other things, the aromatics can be completely saturated.

The 80-85% conversion of normal paraffins is optimal; unconverted normal paraffins would cause problems at low temperatures and are removed by solvent filtration and recycled to a minimal extent after the separation by distillation of light components. Instead of filtration, a catalytic isomerization reactor may be used for the selective removal of normal paraffins; this is not shown in Fig. 6 (Baker and...
In order to take account of the larger oil content in comparison to wax, and thus of the more significant adsorption of oil molecules onto the active sites, where, also in this case, they undergo similar reactions to hydrocracking, the waxy oil hydrosisomerization (see again Fig. 4 in Chapter 8.1) is carried out on more acidic catalysts (amorphous matrices are not used) based on zeolites or molecular sieves with phosphorus. To minimize cracking reactions, acidity must not be excessive, and must be balanced by the hydrogenating-dehydrogenating function of the noble metals.

Fig. 7 shows a simplified flow diagram of the waxy oil hydrosisomerization process. In the first reactor the waxy oil is hydrosisomerized and in the second is hydrotreated as the finishing step. Recycle gas separation and lights distillation follow.

The waxy oil hydrosisomerization is similar in the 315-370°C reaction temperatures to wax hydrosisomerization process, but the pressure above 7 MPa is higher (Everett and Suchanek, 1996), due to the greater complexity of the reactions. It is also possible to use the waxy oil from mild hydrocracking of the deasphalted oil as a feedstock in order to obtain group II bright stock (see again Fig. 4 in Chapter 8.1). The base oils yields depend on the desired viscosity index and pour point, on the quality of the feedstock, and may reach higher values (about 80-90%) than the yields from solvent dewaxing in the production of group I base oils (Baker and McGuiness, 1995).

Due to the dilution of the paraffins with the oil, the products cannot attain the group III⁺ quality level; this can be achieved by hydrosisomerizing waxes. Specifically, the production of the bright stock nowadays is of the group II quality level.

In comparison to the waxy oil hydrosisomerization of oils, the removal of normal paraffins by cracking requires a higher degree of acidity in the catalyst matrices in order to increase cracking. To avoid excessively low yields, these matrices must be more selective as molecular filters for normal paraffins than the catalysts used for the waxy oils hydrosisomerization. Noble metals are used on matrices, which contain molecular sieves or zeolites with pore diameters such as to allow the process to be applied to all waxy distillates and bright stocks (Hargrove et al., 1979; Baker and McGuiness, 1995; Helton et al., 1998).

The process is thus more flexible than urea dewaxing, which is restricted exclusively to light distillates, and is of interest for paraffinic crudes in order to manufacture specialty oils with high fluidity at low temperatures (e.g. transformers and refrigerator oils). These specialty oils are traditionally produced by refining naphthenic crudes.

In comparison to hydrosisomerization processes, milder operating conditions are used in order to moderate cracking reactions; the highest temperatures and pressures are lower than the highest temperatures and pressures in hydrosisomerization processes (Everett and Suchanek, 1996). The conversion of normal paraffins by cracking increases as the contact time and the temperature increase and, as a consequence, the pour point is lowered until it tends towards a minimum.

Due to the cracking of the normal paraffins, the base oil is inevitably obtained with a 7-8 points viscosity index lower than the feedstock, along with lighter olefinic products, in this case basically gas, LPG, virgin naphtha and small amounts of kerosene and gas oil. As compared with solvent dewaxing, given the same pour point, the more selective removal of normal paraffins leads to the production of base oils with better low temperature properties, but the yield is generally lower.

The olefin and aromatic content in the dewaxed base oil makes a final stabilization by means of a hydrofinishing reactor necessary; this is integrated into the process. After the reactors the recycle gas is recovered in two separators in a series, of which the first is at higher temperature and pressure. The liquid bottoms from the separators are then fractionated. This selective cracking is thus similar in flow diagram to the waxy oil hydrosisomerization (see again Fig. 7).

Other treatments

The preceding remarks are not an exhaustive description of the technologies for lubricant base oil refining: base oil finishing, white oil production and the wax treating are still to be described.

The production of specialty oils remaining in the liquid state at extremely low temperatures, e.g. transformers and refrigerator oils has already been discussed above, where the production of white oils was also mentioned.

Aromatic extracts are used without further treatment, unless it is necessary to reduce their sulphur content by means of hydrotreating in order to facilitate the production of fuel oils.

Base oil finishing

After solvent and hydrogen processing, the base oils are usually still not ready to be used in the production of lubricating oils. It is necessary to employ finishing processes in order to improve oxidation stability over time, reduce foaming tendency and remove residual acidity by removing some of the aromatics and sulphur and the remaining oxygenate compounds. It is important to remove nitrogen compounds because these promote the oxidation of the
oil and alter its colour, which is the first indicator of the constant quality of the base oil over time.

The conventional finishing treatments use natural or activated absorbent clays, which are generally based on aluminium silicates, by means of percolation cylindrical filters, or by mixing (clay contacting). In the latter the clay is constantly mixed with the oil and the mixture is then filtered, e.g. through a rotating drum filter (Nelson, 1958; Normand, 1973).

Clay treating has been replaced by more modern treatments with hydrogen over hydrogenating catalysts at moderate pressures (hydrofinishing) in order to eliminate spent clay disposal, and provide better results in terms of yields, selectivity and colour stability.

This technology also helps to raise the quality of the base oil, completing the processes of hydrocracking and hydroisomerization downstream when the base oil is turned into a highly refined extremely stable white oil by the removal of sulphur, nitrogen, aromatics and trace quantities of olefins in order to reduce engine emissions and improve the life and effectiveness of catalytic exhaust converters. In this case, noble metals on weakly acidic matrices are used as catalysts. These plants provide higher yields, because it is unnecessary to increase the severity in hydrocracking and hydroisomerization processes in order to obtain the same quality levels in the base oil (Werner et al., 2001; Refining [...], 2002).

White oil production

White oils are characterized by their water-like transparency (water-white) and by excellent stability, due to their extremely low aromatics and nitrogen content. They are subdivided into two grades: technical white oils and medicinal white oils (paraffinum liquidum). The former are used to lubricate precision mechanisms, such as clocks and sewing machines and in the latter the aromatics content is present in minimal or trace quantities.

The medicinal grade white oils, used e.g. to manufacture cosmetics, must meet legal purity standards set by the pharmacopoeias of different countries on aromatics content and level of refining, which is expressed as the quantity of unsulphonated compounds (i.e. the percentage of the oil which is not attacked by sulphuric acid).

White oils are produced from paraffinic base oils by means of a sulphuric acid treatment followed by the separation of the acid sludges, neutralization with slaked lime and the filtration over clay; today this is usually followed, or replaced, by catalytic hydrogenation in plants similar to hydrofinishing units.

The naphthenic distillate processing may consist of a single severe hydrogenation stage at high pressure to saturate aromatics. Two stages are required if the quality of the feedstock is low: after the vacuum steam stripping of the more volatile components, the first stage supplies the technical white oils, which can be sent to feed the second stage of hydrogenation in order to obtain the medicinal white oils. The hydrogen processing involves the use of catalysts based on nickel or noble metals on an acidic amorphous matrix for the final reactor and for the more severe treatments (Normand, 1973; Everett and Suchanek, 1996; Refining [...], 2002).

Wax treating

The fractionation of the deoiled waxes into soft and hard waxes and their finishing are not always carried out in the refinery due to the relatively small market demand. Often, independent manufacturers buy slack wax, scale wax and petrolatum in order to further process them. They operate by means of MEK/toluene or MIBK/MEK fractionation of scale wax to produce soft wax and hard wax. Waxes are decoloured until a white colour is obtained and deodorized by means of treatments which may involve the use of sulphuric acid (decolourizing).

In this way, waxes and petroleum jellies are produced for industrial uses. To prepare full refined petroleum jellies and waxes, i.e. meeting the purity standards required by law (e.g. for use in the cosmetics industry), it is necessary to use more severe acid treatments involving separation of acid sludge, neutralization and filtration over clays (Normand, 1973).

The modern wax finishing processes involve catalytic hydrogenation similar to catalysts to base oils hydrofinishing and white oils processing, but under more severe operating conditions than hydrofinishing and sometimes in plants equipped with two reactors in series (Refining [...], 2002).

8.2.2 Blending

The mixing of the base oils with additives must be carefully carried out in order to prevent additives from stratifying or precipitating over time. Lubricating oils are non-homogeneous liquids, in which over time the heavier molecules tend to form a layer at the bottom of a tank or container.

To facilitate blending and transportation, the more viscous additives are mixed with a base oil to form a ‘concentrate’ easier to blend due to its lower viscosity. This also occurs for those additives in solid at room temperature: dissolving them in a base oil, or forming a suspension as in the case of graphite, makes them easy to use for blending.
Blending is operated under controlled temperature conditions. The feeding of the various components to the mixer depends on their viscosity and quantity, and is determined on the basis of experience and laboratory tests. In general, the manufacture of a lubricating oil is checked in the laboratory on a blend of the same components to be used.

For the production of small quantities of lubricating oils, the various equipment is often run in batch mode. The mixer agitation is carried out by means of air blowing, or by a rotating impeller, or the intake of an oil jet, or the circulation of the product. For continuous production in bulk, large mixers with impellers are used, or the blending is carried out in line, successively adding the components in a flow mixer placed on a pipe and recycling the mixture until it is finished.

The production of greases requires tailored additives and the use of special equipment, usually kettles equipped with rotating impellers, where the temperature is first raised to a maximum and then lowered under controlled conditions.

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