# CHAPTER 10
# LUBE BASE STOCKS

Lubricants are required in machines to reduce friction and wear between moving parts. Lubricant base stocks make up a large portion of finished lubricants, from about 75 to 80 percent in automotive engine oils to 90 percent or more in some industrial oils. Thus, base stocks contribute significantly to the finished product properties. Base stock has a major impact on the viscosity, volatility, low temperature fluidity, solvency for additives and contaminants, demulsibility, air release, foam, oxidation, and thermal stability of the finished product. Petroleum lubricating base stocks are made of a higher boiling portion of crude oil that remains after the removal of lighter hydrocarbons. Starting material for their manufacture is usually atmospheric residue boiling above 650°F. However, atmospheric resid from every crude is not suitable for lubricating oil manufacture. Careful selection of a base stock is key to formulating a quality finished lubricant. Base stock properties are related to base stock composition. Base stocks contain three types of hydrocarbon; paraffins, naphthenes, and aromatics. In the paraffin group, isoparaffins are the preferred type because they exhibit excellent oxidation stability, low volatility, and good viscosity characteristics. Normal paraffins, however, are not a desirable component because of their poor cold flow properties such as pour point, cold filter plug point (CFPP). Aromatics are good for the solvency of additives and contaminants but generally have poor oxidation stability and high volatility. Naphthenes also have good low temperature fluidity and oxidation stability. Sulfur and nitrogen are often present in combination with hydrocarbons in a base stock, particularly the aromatics. Sulfur can improve oxidation stability but may also contribute to deposit formation and color instability. Nitrogen promotes oxidation and deposit formation. The preferred lubricant base stocks for a wide range of product application contain predominantly isoparaffins and naphthenes with a balance of aromatics and sulfur for proper solvency and oxidation stability.

The manufacture of lube base stocks from crude oil involves a series of steps aimed at the removal of certain undesirable components resulting in a base oil that meets the performance requirements of lubricating oils. There are two basic routes for making lube base stocks; the conventional process, consisting of solvent extraction, solvent dewaxing, and hydrofinishing, and the hydroprocessing route, consisting of lube hydrocracking, hydrodewaxing, and deep hydrotreating. The hydrotreating route produces higher viscosity index (VI) lubes with superior quality but cannot produce high-viscosity lube base stocks.

## *CONVENTIONAL PROCESS*

The conventional lube base stock manufacturing process consists of the following steps:

- Vacuum distillation of atmospheric resid to yield several distillate cuts and vacuum resid
- Propane deasphalting of vacuum residuum to yield bright stock and asphaltic pitch
- Solvent extraction of vacuum distillates and bright stock to remove aromatics and improve the viscosity index of lubricating oil base stock
- Solvent dewaxing of distillate cuts to yield slack wax and various lube cuts, which improves the cold flow properties such as pour point and, cloud point of the lube base stock
- Hydrofinishing or clay treatment to improve color, oxidation stability, and thermal stability of lubricating oils

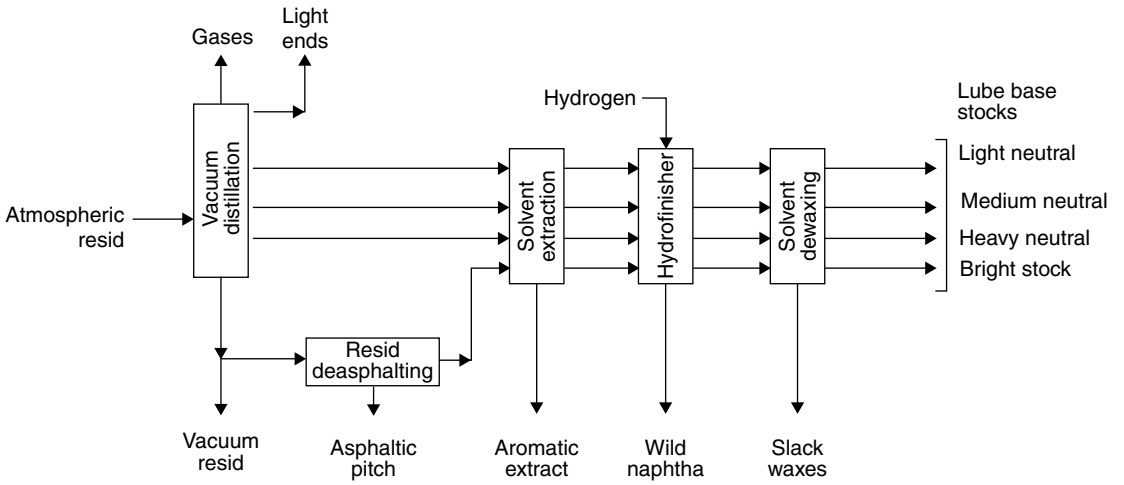


**FIGURE 10-1** Conventional lube base stocks manufacture.

Figure 10-1 shows the processing scheme for a conventional lube plant comprising vacuum distillation, furfural extraction, MEK dewaxing, and propane deasphalter. Tables 10-1 and 10-2 show the material balance and stream qualities for a plant processing Arabian Light crude.

**TABLE 10-1** Lube Oil Plant Overall Material Balance*

| Stream | Product | Wt fraction on ATM resid | Cut point, °F | Specific gravity | Viscosity, cSt, 122°F | Sulfur, Wt % |
|---|---|---|---|---|---|---|
| 1 | Atmospheric resid | 1.0000 | 633.3 | 0.9710 | 830 | 4.20 |
| 2 | Vacuum resid | 0.4700 | 919.4 | 1.0320 | 131,500 | 5.40 |
| 3 | Feed to PDA | 0.1900 | | | | |
| 4 | Vacuum overhead oil | 0.0160 | 365 | 0.8640 | 3 | 1.50 |
| 5 | Vacuum top excess | 0.0820 | 494.6 | 0.8800 | 4 | 1.50 |
| 6 | Asphalt | 0.1368 | | 1.0740 | * | 6.40 |
| 7 | Light extract | 0.0576 | | 1.0011 | 39 | 6.10 |
| | Medium extract | 0.0678 | | 1.0136 | 140 | 5.80 |
| | Heavy extract | 0.0539 | | 1.0088 | 1786 | 5.50 |
| | Bright stock extract | 0.0161 | | | | |
| 8 | Wild naphtha | 0.0093 | | | | |
| 9 | Light slack wax | 0.0205 | | 0.8100 | 8 | 0.25 |
| | Medium slack wax | 0.0204 | | 0.8330 | 15 | 0.07 |
| | Heavy slack wax | 0.0134 | | 0.8500 | 28 | 0.10 |
| | Bright stock slack wax | 0.0066 | | 0.8700 | 65 | 0.31 |
| 10 | Light neutral | 0.0721 | | 0.8613 | | |
| | Medium neutral | 0.0759 | | 0.8729 | | |
| | Heavy neutral | 0.0419 | | 0.8824 | | |
| 11 | Bright stock | 0.0297 | | 0.8984 | | |
| 12 | Net vacuum resid | 0.2800 | | | | |

*Comprising vacuum distillation, propane deasphalter, solvent extraction, and *MEK* Dewaxer.
Basis: Light Arabian crude processing

**TABLE 10-2** Lubricating Oil Cut Properties from Lube Plant

| | Specific gravity | Viscosity, cSt 104°F | Sulfur, Wt % | VI | Oil content, Wt % |
|---|---|---|---|---|---|
| Solvent extracts | | | | | |
| Light | 1.016 | 68.6 | 3.00 | | |
| Medium | 1.012 | 565 | 5.00 | | |
| Heavy | 1.015 | 2640 | 5.00 | | |
| Bright stock | 0.983 | — | — | | |
| Slack waxes | | | | | |
| Light | 0.820 | 0.02 | | | 6.4 |
| Medium | 0.869 | 0.03 | | | 9.9 |
| Heavy | 0.890 | 0.08 | | | 30.5 |
| Bright stock | 0.905 | 0.16 | | | 15.1 |
| Base oils | | | | | |
| Light | 0.860 | 19.10 | 0.10 | 95 | |
| Medium | 0.880 | 60.20 | 0.20 | 96 | |
| Heavy | 0.885 | 131.50 | 0.15 | 96 | |
| Bright stock | 0.897 | 473.00 | 0.35 | 97 | |

## Vacuum Distillation

Vacuum distillation separates the atmospheric residue into a series of fractions representing different viscosities/molecular weight ranges. The objective is to isolate hydrocarbons with the proper boiling range and viscosity characteristics suitable for lubricant manufacture. These fractions, by convention, are identified by "neutral numbers," which are, in fact, their Saybolt Universal second (SUS) (measure of kinematic viscosity) at 100°F. Typically the neutral number ranges from 90 to 600 for various straight run fractions (Table 10-3). For example, "90 neutral" indicates a straight run lube cut with a viscosity of 90 SUS (approximately 17.5 cSt) at 100°F. A process flow diagram of a refinery vacuum distillation column for lube manufacture is shown in Fig. 10-2. Operating conditions of a refinery vacuum distillation unit (VDU) column are shown in Table 10-4.

**TABLE 10-3** Viscosities of Raw Lube Cuts from Vacuum Distillation and Propane Deasphalter Unit

| Neutral number | Viscosity, 100°F SUS | Viscosity, 100°F cSt |
|---|---|---|
| 50 | 50 | 7.38 |
| 60 | 60 | 10.35 |
| 70 | 70 | 13.08 |
| 80 | 80 | 15.66 |
| 90 | 90 | 18.13 |
| 100 | 100 | 20.53 |
| 150 | 150 | 31.90 |
| 175 | 175 | 37.45 |
| 200 | 200 | 42.90 |
| 250 | 250 | 59.25 |
| 300 | 300 | 64.65 |
| 350 | 350 | 75.46 |
| 500 | 500 | 107.90 |
| 650 | 650 | 140.32 |

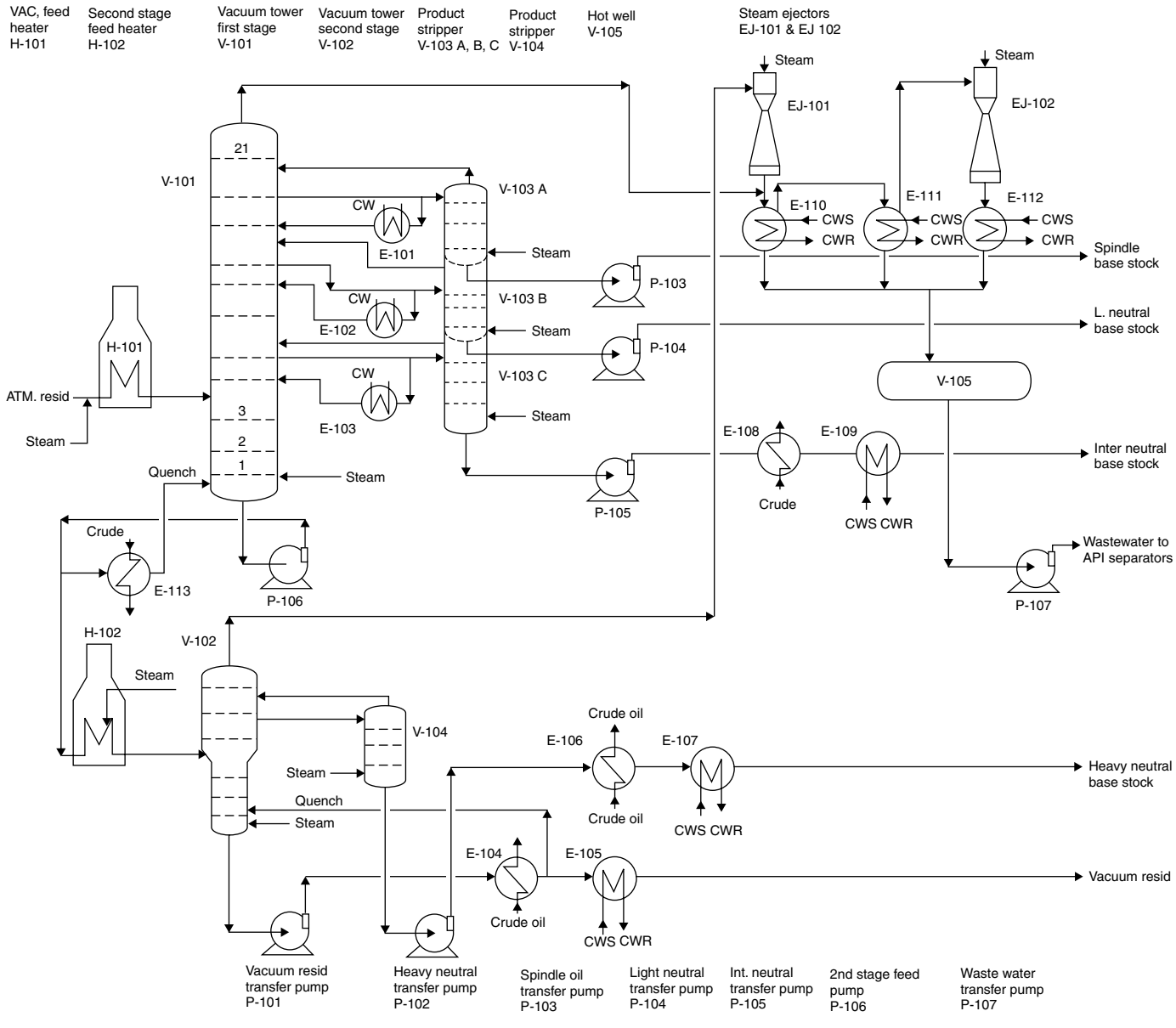


**FIGURE 10-2** Vacuum distillation for lube base stocks manufacture.

**TABLE 10-4** Operating Conditions for Vacuum Column Lube Operation

| | Operating conditions | |
|---|---|---|
| | 1 | 2 |
| Pressure | | |
| Top, mm Hg | 97 | 66 |
| Flash zone, mm Hg | 156 | 106 |
| Temperature | | |
| Top, °F | 293 | 185 |
| Flash zone, °F | 765 | 740 |
| Fractionation efficiency | | |
| Streams | Overlap 95/5, °F | |
| Vacuum gas oil–100N | 62 | |
| 100N–195N | 88 | |
| 195N–340N | 104 | |
| 340N–650N | 128 | |
| 650N–residuum | 139 | |

## Propane Deasphalting Unit

Vacuum residue contains heavier base oils that are separated from asphaltenes and resins by a solvent deasphalting process. Propane is the preferred solvent. The oil separated by solvent deasphalting of vacuum residue is known as Bright stock, which is much more viscous and heavier than straight run vacuum distillates. Solvent deasphalted base stocks are identified by their SUS viscosity at 210°F. Thus, "150 Bright Stock" implies a nominal viscosity of 150 SUS at 210°F. Table 10-5 shows the typical operating conditions for a PDA unit.

**TABLE 10-5** Propane Deasphalting Unit Operating Conditions
Feed: Vacuum residue from Aghajari crude

| | Viscosity 1350 cSt at 210°F |
|---|---|
| Propane-to-feed ratio, V/V | 5.30 |
| Extractor pressure, lb/in$^2$ | 455 |
| Extractor top temperature, °F | 154 |
| Extractor bottom temperature, °F | 126 |
| Furnace coil outlet temperature, °F | 410 |
| Deasphalted oil viscosity, cSt, 210°F | 38 |
| DAO conradson carbon, Wt % | 2.1 |
| PDA asphalt penetration, ASTM D5 | 5 |

## Solvent Extraction

The objective of solvent extraction of lube base stocks is to improve the viscosity index (VI) of lube base stocks. The feed to the unit are vacuum distillate cuts such as spindle, light neutral, intermediate neutral, heavy neutrals, and bright stock (produced from propane deasphalting of vacuum resid). The operation of the unit is same for all the grades except the operating conditions such as the solvent-to-feed ratios and feed temperatures, which are adjusted as required. All the feeds are processed in a solvent extraction unit in a blocked-out operation. Solvent extraction process removes undesirable components such as aromatics (low VI) and compounds containing heteroatoms such as oxygen, nitrogen, and sulfur, present in vacuum distillates and residual stocks.

The separation of low lubricating oil quality components is controlled by the quantity of the solvent employed and by the solvent temperature. The solvents that have been used in the past are benzene, phenol, nitrobenzene, liquid sulfur dioxide, furfural, and many others. Except for furfural, these solvents are no longer used because of their high toxicity and low biodegradability. Furfural is by far the most popular solvent because it is nontoxic and biodegradable. The molecular structure of furfural is shown in Fig. 10-3. The physical properties of furfural are shown in Table 10-6. In recent years, another solvent, normal methyl pyrrolidine (NMP), has claimed popularity. Figure 10-4 shows the molecular structure of NMP. Many new solvent extraction plants are being built using NMP solvent, and the physical properties of furfural and NMP are shown in Tables 10-6 and 10-7.

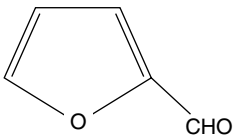


**FIGURE 10-3** Furfural structure.

The following advantages are claimed for NMP solvent over furfural in lube oil extraction plants:

- NMP requires a much smaller solvent-to-feed ratio, resulting in smaller plant size.
- The stability of NMP is better than that of furfural; therefore solvent consumption is much lower than that for furfural solvent.
- NMP is noncoking and has lower toxicity.

**TABLE 10-6** Properties of Furfural Solvent 2-Furfuraldehyde*

| | |
|---|---|
| Formula | $C_5H_4O_2$ |
| Formula weight | 96.082 |
| Specific gravity | 1.161 |
| Melting point, °F | –33.7 |
| Boiling point, °F, 760 mm | 323.1 |
| Viscosity, CP, 77°F | 1.494 |
| Refractive index | 1.5262 |
| Flash point, °F, CC | 138.2 |
| Ignition temperature, °F | 599 |

*At 1 atm, furfural forms an azeotrope with 65.0% water, boiling at 207.5°F.

The solvent extraction process produces high-quality lubricating oils characterized by a high VI, good thermal and oxidation stability, light color, and a good additive response. The by-product extract phase, rich in aromatics, is used as a carbon black feed stock, rubber extender oils, and many other nonlube uses. Figure 10-5 shows a process flow diagram of a refinery solvent extraction unit. The lube feed stock (distillate or deasphalted residual stock) is contacted with solvent in an extraction tower of a rotary disk contactor where both the solvent and feed are fed continuously and the raffinate phase, consisting mainly of paraffins and some solvent, and the extract phase, consisting of mainly aromatics and some solvent, are separated. The raffinate and extract phase are then taken to separate solvent recovery trains, consisting mainly of heating furnaces and distillation columns. Both raffinate and

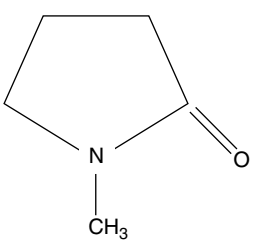


**FIGURE 10-4** n-Methyl-2-Pyrrolidinone structure.

**TABLE 10-7** Properties of n-Methyl-2-Pyrrolidinone Solvent

| | |
|---|---|
| Formula | $C_5H_9NO$ |
| Formula weight | 99.13 |
| Density | 1.0279 |
| Melting point, °F | –11.92 |
| Boiling point, °F | 395.6 |
| Viscosity, CP, 100°F | 1.65 |
| Refractive index | 1.468 |
| Flash point, °F | 196 |
| Surface tension, dynes/cm | 40 |
| Evaporation rate (NBAC = 1) | 0.03 |
| Kauri butanol value | 350 |

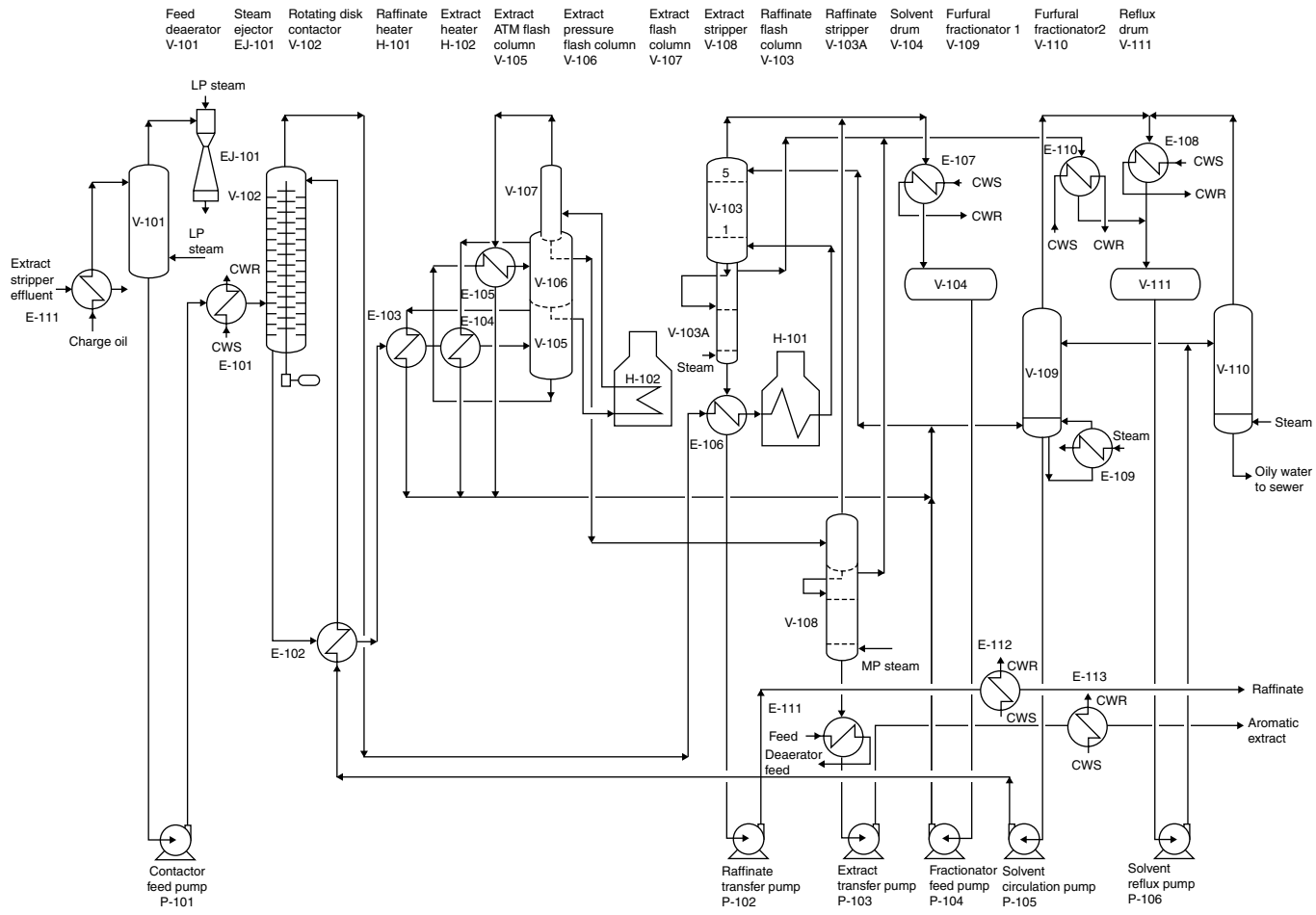


**FIGURE 10-5** Furfural extraction plans.

extract streams are steam stripped to remove traces of solvent from these streams. Feed and solvent flow rates are also controlled, and these are contacted in a countercurrent manner in the tower. Raffinate stream exits from the top of the tower and is routed to a solvent recovery section for separation of solvent from this stream. The extract stream containing the bulk of the solvent exits the bottom of the extraction tower and is routed to the recovery section for removal of solvent from this stream. Solvent is separated from the extract phase by multiple effect evaporation at various pressures followed by vacuum flashing and steam stripping under vacuum. Solvent is separated from the primary raffinate by vacuum flashing and steam stripping under vacuum. The overhead vapors from steam strippers are condensed and combined with solvent condensate from recovery sections and are distilled at low pressure to remove water from solvent. Furfural forms an azeotrope with water and requires two fractionators. One fractionator separates furfural from azeotrope, and the second separates water from azeotrope. Water is drained to a oily water sewer. Solvent is cooled and recycled to extraction tower. Typical operating conditions, yields, and stream properties for a solvent extraction unit processing Middle Eastern crude are shown in Tables 10-8, 10-9, and 10-10.

**TABLE 10-8** Furfural Extraction for Lube Manufacture Lube Base Stocks Yield
Feed: Vacuum distillates, propane deasphalted oil from Aghajari (Iranian) crude

| | Spindle oil | Intermediate oil | Heavy oil | Vacuum residue |
|---|---|---|---|---|
| Yield on crude, Wt % | 12.1 | 5.5 | 4.5 | 20 |
| Deasphalted oil yield, Wt % | | | | 40 |
| Products | Spindle oil | Intermediate oil | Heavy oil | Bright stock* |
| Raffinate yield, Wt % | 56 | 57 | 54 | 63 |
| Extract yield, Wt % | 44 | 43 | 46 | 37 |
| Properties (Raffinate) | | | | |
| Viscosity index | 95 | 95 | 90 | 90 |
| Viscosity,† 102°F, cSt | 10.4 | 62.8 | 143.5 | 507.5 |
| Viscosity,† 210°F, cSt | 2.7 | 10.2 | 13.7 | 31 |

*Bright stock yield on vacuum residue.
†Viscosity after dewaxing.

**TABLE 10-9** Solvent Extraction of Lube Feedstocks Operating Conditions
Basis: Feed vacuum distillate and bright stock feed ex 34.5° API light Arabian crude
Furfural extraction unit

| Operation*,† | 100 N | 300 N | 650 N | 150 BS |
|---|---|---|---|---|
| Feedstock nominal viscosity, cSt, 100°F | 20.53 | 64.65 | 140.32 | 530.3 |
| cSt, 210°F | | | | 31.9 |
| Solvent-to-feed ratio, V/V | 1.5 | 1.8 | 2.8 | 2.9 |
| Furfural miscibility temperature at ratio, °F | 244.4 | 267.8 | 276.8 | 311 |
| Extracted raffinate VI | 95 | 96 | 96 | 97 |

*Term "100 N," or 100 neutral, indicates that the nominal viscosity of feedstock at 100°F in SUS units is 100. "300 N" and "650 N" operations are similarly defined.
†Term "150 BS," or "150 bright stock," indicates that the nominal viscosity of feedstock at 210°F is 150 SUS.

**TABLE 10-10** Furfural Extraction of Bright Neutral Operating Conditions
Feed: Bright neutral from light Iranian (Aghajari crude)

| | Operating conditions | |
|---|---|---|
| | 1 | 2 |
| Solvent-to-feed ratio, V/V | 1.470 | 1.720 |
| Extractor temperature | | |
| Top, °F | 269.6 | 266 |
| Bottom, °F | 197.6 | 176 |
| Raffinate density, gm/cc | 0.9095 | 0.9068 |
| RI, 140°F | 1.48 | 1.483 |
| Furfural content | | |
| Raffinate, ppm | 50 | 20 |
| Extract, ppm | 40 | 50 |

## Solvent Dewaxing

The next step in conventional lube oil manufacture is removal of wax to improve the flow characteristics at low temperatures. Feed, the waxy oil, is mixed with a solvent (such as methyl ethyl ketone or methyl isobutyl ketone and the mixture is cooled from 10 to 20°F (6 to 12°C) below the desired pour point of lube oil. The wax crystals that are formed are removed by filtration. A process flow diagram of a refinery solvent dewaxing unit is shown in Figs. 10-6 and 10-7, and the operating conditions for a solvent dewaxing unit are shown in Tables 10-11 and 10-12. Properties of some dewaxing solvents such as methyl ethyl ketone (MEK), methyl isobutyl ketone (MIBK), benzene, and toluene are shown in Table 10-13.

## Lube Hydrofinishing

Some base stocks produced by conventional processing scheme of solvent extraction and dewaxing require a finishing step, such as hydrofinishing or clay treatment, to improve color, oxidation stability, and thermal stability lubricating oil base stock produced. Hydrofinishing consists of passing the heated oil and hydrogen through a bed of catalyst. The process removes some color bodies and some unstable compounds containing nitrogen and sulfur. It saturates residual olefins to form paraffins. This process stabilizes base stock color and improves demulsibility and air release characteristics. A slight improvement in oxidation resistance may also result. An alternative process is clay treatment, which removes dark-colored and unstable molecules. Hydrofinishing represents a relatively mild operation at relatively low temperatures and pressures. A process flow diagram for a lube hydrofinishing unit is shown in Fig.10-8. Table 10-14 presents the overall lube yield from a conventional lube plant. Operation conditions for lube hydrofinishing unit are shown in Tables 10-15 and Table 10-16.

## Hydroprocessing Route

***Lube Hydrocracking.*** An alternative method to solvent extraction is hydrocracking to reduce aromatics in the base stocks. In this process, the aromatics are converted to naphthenes, paraffins, and fuel components by breaking carbon to carbon bonds at a high temperature and high hydrogen pressure, in the presence of a catalyst, rather than their physical removal. Depending on the severity,

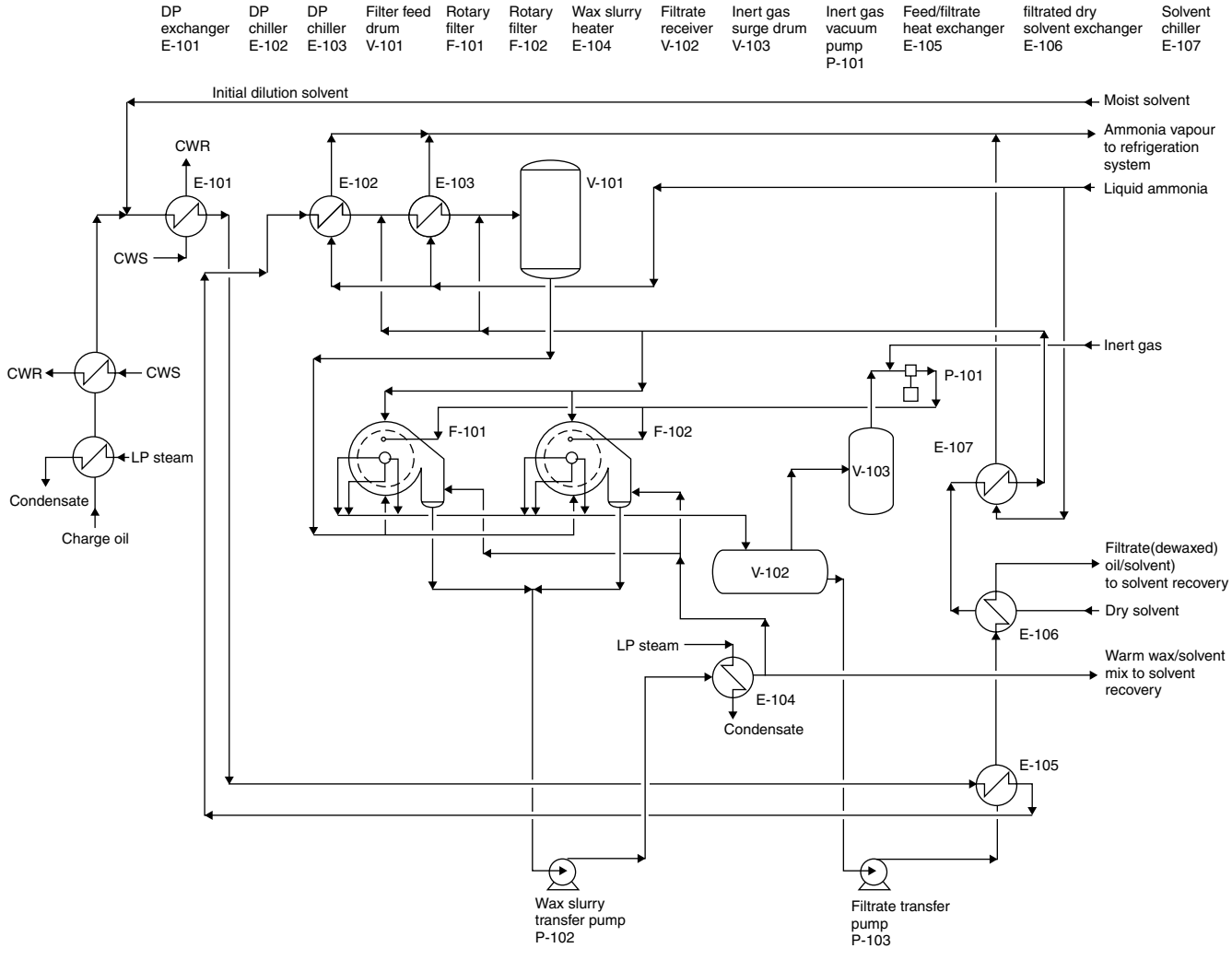


**FIGURE 10-6** Solvent dewaxing of lube oils.

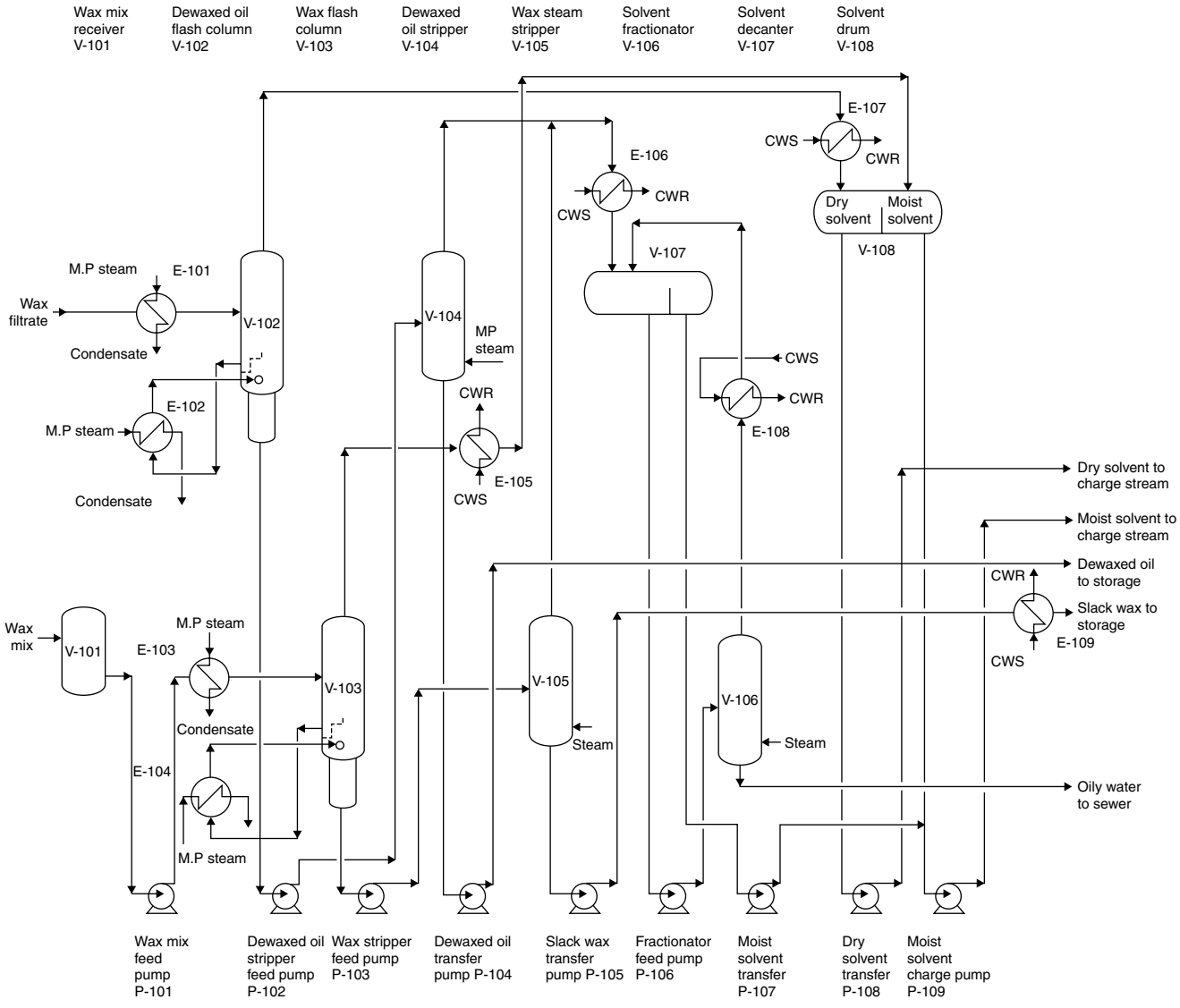


**FIGURE 10-7** Lube dewaxing (solvent recovery section).

**TABLE 10-11** Solvent Dewaxing of Lube Base Stocks Operating Conditions
Feed: Ex waxy asian crude

| | | | | |
|---|---|---|---|---|
| Cut, °F | 840–915 | | | |
| Density | 0.8585 | | | |
| Viscosity, cSt, 210°F | 5.81 | | | |
| Pour point, °F | 134.6 | | | |
| Wax content, Wt % | 69 | | | |
| Solvents | | MEK + BZ + TOL | MIBK + BZ + TOL | MIBK |
| Solvent composition: | Vol % | | | |
| Methyl ethyl ketone | | 40 | | |
| Benzene | | 30 | 30 | |
| Toluene | | 30 | 30 | |
| MIBK | | | 40 | 100 |
| Solvent-to-feed ratio | | 5/1 | 5/1 | 5/1 |
| Solvent addition temperatures | °F | | | |
| | 158 | 1.0 | 1.0 | 1.0 |
| | 96.8 | 1.0 | 1.0 | 1.0 |
| | 32 | | | |
| | –7.6 | 1.5 | 1.5 | 1.5 |
| | –13 | 1.5 | 1.5 | 1.5 |
| Total solvent-to-feed ratio | | 5.0 | 5.0 | 5.0 |
| Average chilling rate | | | | |
| Between 158 and 132°F | °F/min | 2.5 | 2.5 | 3 |
| Below 32°F | °F/min | 1 | 1 | <0.5 |
| Dewaxing temperature | °F | –13 | –13 | –13 |
| Wash solvent-to-feed ratio | V/V | 1.5/1 | 1.5/1 | 1.5/1 |
| Dewaxed oil | | | | |
| Yield | Wt % | 32.2 | 23.7 | 27.4 |
| Pour point | °F | 24.8 | 26.6 | 6.8 |
| Kinematic viscosity | cSt, 210°F | 11 | 11.4 | 12.1 |
| Viscosity index | | 61 | 67 | 58 |
| Slack wax | | | | |
| Yield | Wt % | 67.8 | 73.9 | 72.6 |
| Conjgealing temperature | °F | 144.5 | 138.56 | 135.5 |
| Delta T (dewaxing temp-pour point) | °F | 37.8 | 39.6 | 19.8 |

BZ = benzene; MEK = methyl ethyl ketone; MIBK = methyl isobutyl ketone; TOL = toluene.

**TABLE 10-12** Solvent Dewaxing Operating Conditions
Feed: Bright neutral Aghajari crude

| | |
|---|---|
| Solvent-to-feed ratio, V/V | 4.00 |
| SDA dosing, ppm | 200 |
| Chiller outlet temperature, °F | 6.8 |
| Filter feed drum temperature, °F | 8.6 |
| Cold wash-to-feed ratio | 1.00 |
| Cold wash temperature, °F | –0.4 |
| Solvent composition | |
| Methyl ethyl ketone, Vol % | 50 |
| Toluene, Vol % | 50 |
| Viscosity of dewaxed oil, cSt, 210°F | 35 |
| VI of dewaxed oil | 93 |
| Pour point of dewaxed oil, °F | 23 |
| Flash point of wax, °F | 424 |

**TABLE 10-13** Properties of Lube Dewaxing Solvents

| | | MIBK | MEK | Benzene | Toluene |
|---|---|---|---|---|---|
| Molecular weight | | 100.1 | 72.11 | 78.1 | 92.1 |
| Boiling point, °F | | 240.62 | 175.28 | 176.18 | 231.08 |
| Specific gravity | | 0.802 | 0.805 | 0.879 | 0.867 |
| Viscosity, CP, 68°F | | 0.585 | 0.425 | 0.649 | 0.587 |
| Water solubility, 68°F | | | | | |
| | Wt % water | 2.41 | 11.95 | | |
| | Wt % ketone | 2.04 | 27.33 | | |
| Azeotropic data | | | | | |
| | Wt % water | 24.3 | 11 | 8.33 | 13.5 |
| | Boiling point, °F | 189.86 | 156.02 | 156.65 | 183.38 |
| Flash point | °F, PMCC* | 60 | 19 | 12 | 40 |
| Freezing point | °F | –112.47 | –124.42 | 41.95 | –138.98 |
| Specific heat (liquid) | 68°F, cal/g | 0.46 | 0.498 | 0.411 | 0.392 |
| Heat of vaporization | 1 atm, cal/g | 86.5 | 106 | 94.1 | 86.8 |

*PMCC = Pensky–Martens Closed Cup.

essentially all sulfur and nitrogen is eliminated from the base stock. This process changes the structure of many molecules in the feedstock. Aromatics are converted into naphthenes, many naphthene rings are broken open, and many paraffinic molecules are rearranged. Lube hydrocracking generally results in base stocks with improved oxidation and color stability. However, with very low aromatic levels, additives and contaminant solubility are a concern for the finished product formulation. The lube hydrocracking process is similar to distillate hydrocracking, a process used for the production of distillate, that is, naphtha, kerosene, diesel, and so on, from vacuum gas oil feed. However, processing conditions are less severe, which minimizes cracking. The process produces molecules that have a high VI and greater oxidation stability. The hydrocracking process allows a great deal of flexibility relative to crude source for the production of high-quality lube base stocks from inferior feedstocks. Although hydrocracking is less dependent on feed stocks than solvent refining, feed stocks still can have a significant impact on the product properties. Hydrocracking vacuum distillates usually produces base stocks in the 95 to 105 VI range. Higher operating severity can increase this to 115+ VI but with loss of yield. Use of a high wax (paraffin) content feed will result in even higher VI stocks. However, the hydrocracking produces predominantly lower viscosity base stocks (approximately 15 to 110 cSt at 100°F) due to cracking of larger, heavier molecules into smaller, lighter molecules. Thus hydrocracked base stocks cannot be used in many heavy industrial and engine oil products for which these must be blended with solvent-refined base stocks or other thickening agents. As described earlier, hydroprocessing results in lower aromatics, sulfur, and nitrogen than most solvent-refined base stocks. In some lubricant applications, hydroprocessed base stocks do offer desirable benefits for formulating finished products, but, they have lower additive and contaminant solubility, which needs to be addressed when formulating many lubricants. Additives response of hydroprocessed lube base stocks is different from those of solvent-refined base stocks. However, they do offer significant benefits in selected lubricant applications. This is particularly true when oxidation stability is critical and where additive content is low, such as in turbine oils. They also provide lower volatility lubes, more economically. Hydrocracked base stocks can be viewed as an important component for lube formulation that is used to develop finished products with superior performance but not a complete replacement of solvent extracted lubes base stocks.

The process flow diagram for a hydrocracker unit for producing lubricating oils is shown in Fig. 10-9. Operating conditions and product qualities are presented in Tables 10-17 and 10-18.

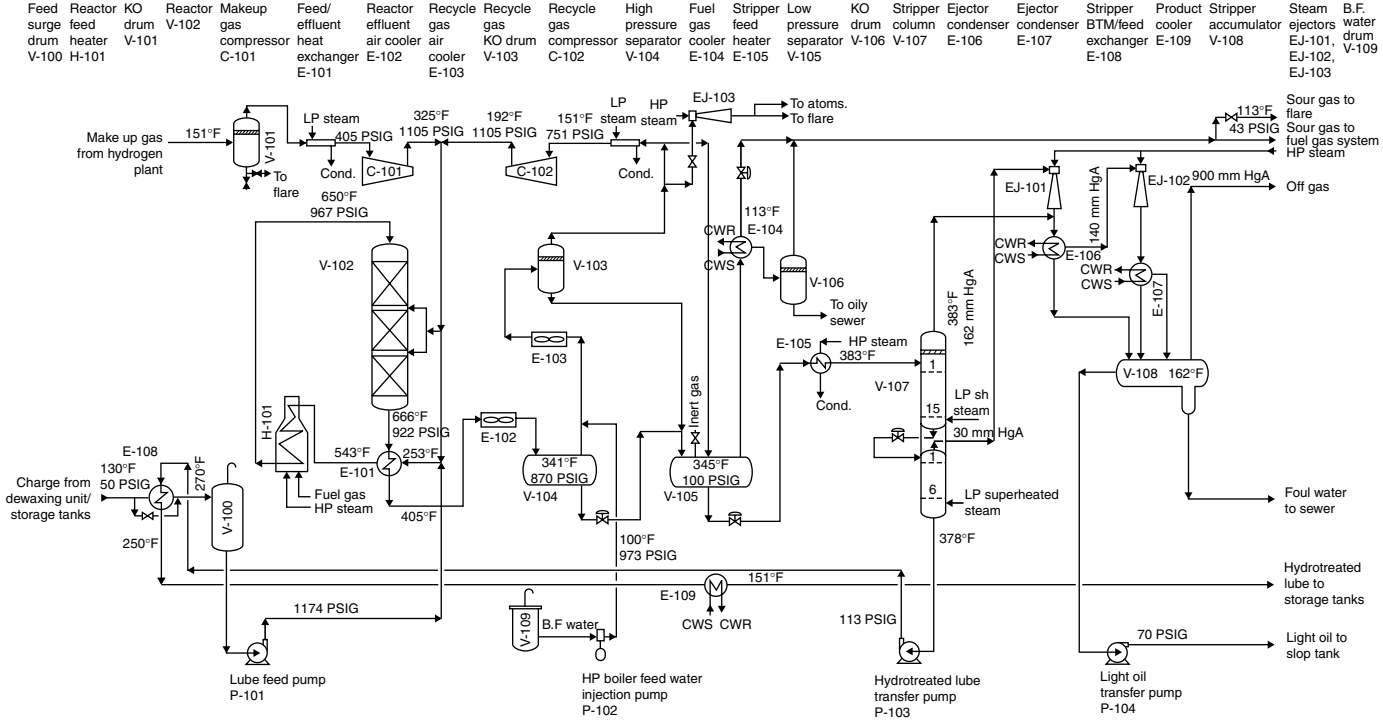


**FIGURE 10-8** Lube hydrofinishing unit.

**TABLE 10-14A** Material Balance, Lube Plant
Feed: Aghajari crude

| Distillation | Spindle | Light neutral | Intermediate neutral | Heavy neutral | Short resid | | | |
|---|---|---|---|---|---|---|---|---|
| Yield on crude, Wt % | 2.3 | 9.8 | 5.5 | 4.5 | 20 | | | |
| Deasphalted oil yield on short resid, Wt % | | | | | 40 | | | |
| Furfural extraction neutral | Spindle | Spindle | Intermediate neutral | Intermediate neutral | Heavy neutral | Heavy neutral | Bright stock | Bright stock |
| | 95 VI | 60 VI | 95 VI | 70 VI | 90 VI | 70 VI | 90 VI | 60 V1 |
| Raffinate yield, Wt % | 56 | — | 57 | 78 | 54 | 75 | 63 | |
| MEK dewaxing | | | | | | | | |
| Raffinate yield, Wt % | 70 | 82 | 79 | 80 | 77 | 80 | 74 | 77 |
| Hydrofinishing | | | | | | | | |
| Yield, Wt % | 99 | 98 | 99 | 98 | 98 | 98 | 98 | 97 |
| Overall yield | 39 | 80 | 44 | 61 | 41 | 59 | 46 | 75 |

**TABLE 10-14B** Product Properties

| | Spindle | Spindle | Intermediate neutral | Intermediate neutral | Heavy neutral | Heavy neutral | Bright stock | Bright stock |
|---|---|---|---|---|---|---|---|---|
| | LVI | HVI | LVI | HVI | LVI | HVI | LVI | HVI |
| Viscosity, 120°F, cSt | 2.7 | 2.7 | 10.2 | 8.1 | 17.3 | 13.7 | 39.5 | 31 |
| Viscosity index | | 95 | — | 95 | | 90 | | 90 |
| Pour point, °C | –12.2 | –12.2 | –6.7 | –6.7 | –6.7 | –6.7 | –6.7 | –6.7 |
| Color ASTM, max. | 1–2 | 0–1/2 | 3–4 | 1–1.5 | | | | |

**TABLE 10-15** Lube Hydrofinishing Unit Operating Conditions and Yields
Feed: 851–972°F Aghajari cut undewaxed and unextracted lube cut
Catalyst: Co-Mo on Alumina

| | | Operating conditions | |
|---|---|---|---|
| Properties | | 1 | 2 |
| Pressure, lb/in$^2$ | | 924 | 924 |
| Temperature, °F | | 707 | 662 |
| LHSV | | 1 | 0.5 |
| Gas-to-oil ratio, SCF/BBL | | 2807 | 2807 |
| Feed and product properties | Feed | Product 1* | Product 2† |
| Density | 0.9095 | 0.8878 | 0.89 |
| Viscosity, 100°F, cSt | 52 | 41.46 | 48.73 |
| Color, ASTM | 8+ | 2 | 2– |
| Total sulfur, Wt % | 1.4 | 0.14 | 0.17 |
| Conradson carbon, Wt % | 0.014 | 0.014 | 0.018 |

*Product 1 as per operating conditions 1.
†Product 2 as per operating conditions 2.

**TABLE 10-16** Lube Hydrofinishing Operating Conditions
Feed: SAE-30 Stock, 27.7 API

| | | |
|---|---|---|
| Reactor | | |
| Reactor inlet pressure, lb/in$^2$ | | 981 |
| Reactor outlet pressure, lb/in$^2$ | | 936 |
| Feed inlet temperature, °F | | 649 |
| Feed outlet temperature, °F | | 666 |
| Space velocity, LHSV | $Hr^{-1}$ | 1.0 |
| Flash drums | | |
| HP flash drum pressure, lb/in$^2$ | | 885 |
| HP flash drum temperature, °F | | 351 |
| LP flash drum pressure, lb/in$^2$ | | 114 |
| LP flash drum temperature, °F | | 345 |
| Recycle gas rate | NSCF/BBL | 2416 |
| Makeup gas (95% purity) | NSCF/BBL | 197 |
| Stripper column | | |
| Feed temperature, °F | | 383 |
| Column top temperature, °F | | 383 |
| Number of plates | | 15 |
| Flash column | | |
| Pressure, mm Hg A | | 30 |
| Temperature, °F | | 378 |
| Recycle compressor | | |
| Recycle gas inlet pressure, lb/in$^2$ | | 765 |
| Recycle gas inlet temperature, °F | | 151 |
| Recycle gas outlet pressure, lb/in$^2$ | | 1119 |
| Recycle gas outlet temperature, °F | | 192 |

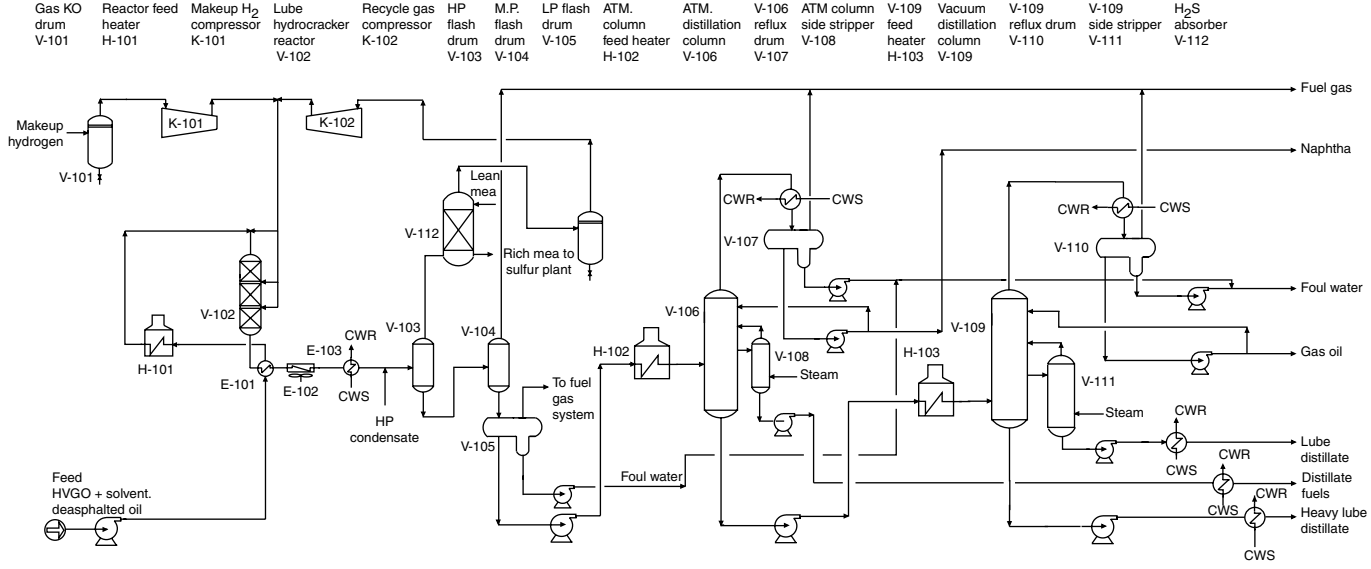


**FIGURE 10-9** Lube hydrocracking unit.

**TABLE 10-17** Lube Hydrocracking Operating Conditions

| Operating parameter | Units | |
|---|---|---|
| Catalyst | | Ni-Mo/Silica-alumina base |
| Reactor inlet temperature | °F | 750 |
| Reactor pressure | $lb/in^2$ | >2500 |
| Hydrogen partial pressure at reactor outlet | $lb/in^2$ | 2000 |
| Space velocity | LHSV | 0.5–1.0 |
| Hydrogen rate | SCFB | 9000 |
| Hydrogen consumed | SCF/BBL | 800–1000 |
| Catalyst life | YR | 1–3 |

**TABLE 10-18** Lube Hydrocracking Feed and Product Properties

| Property | Units | Feed | Product |
|---|---|---|---|
| Feed | | | |
| Deasphalted oil | | | |
| API gravity | | 21.5 | |
| Viscosity | SUS, 210°F | 60.1 | |
| Viscosity | cSt, 210°F | 10.4 | |
| Pour point | °F | 95 | |
| Sulfur | Wt % | 1.3 | |
| Carbon residue | Wt % | | |
| Product | | | |
| 650°F + Lube cut after dewaxing | | | |
| VI | | | 109 |
| Viscosity | SUS, 210°F | | 43.3 |
| | cSt, 210°F | | 5.3 |
| Sulfur | Wt % | | 0.03 |
| Carbon residue | Wt % | | 0.05 |
| Yield | Vol % feed | | 42 |

## *CATALYTIC DEWAXING*

In conventional lube making processes, wax (n-paraffin) is physically separated from the feedstock by solvent dewaxing to improve the cold flow properties of lube oil. In catalytic dewaxing processes, n-paraffins are isomerized to isoparaffins in the presence of a catalyst. Catalytic dewaxing is an alternative to solvent dewaxing, particularly in regions with a low demand for waxes. Feed to the process can be slack paraffin wax, microcrystalline wax, or Fischer-Tropsch wax. The pour point may range from –9 to –15°C. Feed stock is mixed with hydrogen, heated to the desired reactor temperature, and charged to a fixed bed hydrodewaxing reactor. Catalyst in the reactor selectively cracks or isomerizes the feed paraffin using a zeolite shape selective catalyst. Catalytic dewaxing can be applied to either solvent extracted or hydrocracked base stocks to reduce the pour point. Isoparaffins have high VI, low pour points, and a good resistance to oxidation. Because the process converts rather than remove paraffin, the process yield of lube base oil is higher compared to that from the solvent dewaxing process.

In the catalyst dewaxing process,[1] feed is first passed over an isomerizing catalyst. The catalysts employed have an acidic component and a hydrogenation component. The acidic component comprises an intermediate pore size silico-alumino-phosphate[2] (SAPO) molecular sieve. The hydrogenation function is provided by platinum and/or palladium metals. By intermediate pore size is meant an effective pore structure in the range of about 5.3 to 6.5 Å when the molecular sieve is in the calcined form. Intermediate pore size molecular sieves typically admit molecules having kinetic diameters of 5.3 to 6 Å with little resistance. The most preferred silico-alumino-phosphate is SAPO-11, although SAPO-31 and SAPO-41 are also used. The most preferred metal component is platinum. Some isomerization of n-paraffins takes place, thus reducing the pour point of the feed. The effluent from the first reactor passes over a second catalyst. This catalyst comprises an aluminosilicate zeolite having a pore size that admits straight chain normal paraffins either alone or with slightly branched chain paraffins but which excludes more highly branched hydrocarbons such as naphthenes and aromatics. Zeolites such as ZSM-5, ZSM-11, ZSM-12, ZSM-23, with platinum and/or palladium hydrogenation function are typically used. In this reactor bed, the waxy paraffins undergo mild cracking reactions to yield nonwaxy products.

The combination of two catalysts produce a greater yield of lube oil than achieved with either catalyst alone. Pour point is reduced both by isomerization reactions and cracking of normal paraffins. Viscosity is also reduced due to cracking reactions. Due to selectivity of the catalyst employed, gas and light ends yield is minimal. Feed nitrogen must be less than 10 ppm for a reasonable catalyst cycle length.

The catalytic operating conditions employed depends on the feed properties and pour point desired. Operating conditions are presented in Table 10-19. The intermediate pore size aluminosilicate zeolite (dewaxing catalyst) may be used in the same reactor as the silico-alumino-phosphate molecular sieve (isomerization catalyst) or it may be used in a separate reactor. When both catalysts are placed in the same reactor, the isomerization catalyst is layered on the top of the dewaxing catalyst. The effluent from the dewaxing reactor is sent to a hydrofinishing unit to provide a more stable lubricating oil. Catalysts used for hydrofinishing are conventional Co-Mo or noble metals such as Pt or palladium on an alumina base.

**TABLE 10-19** Catalytic Isomerization Dewaxing Operating Conditions
Feed: Lube hydrocracker unconverted bottoms

| Property | Units | Value |
|---|---|---|
| Reactor temperature | °F | 480–550 |
| Reactor pressure | lb/in$^2$ | 1100 |
| Space velocity | LHSV | 1.00 |
| Hydrogen-to-oil ratio | SCF/B | 2000 |
| Hydrogen consumption | SCF/B | 300–500 |

## Deep Hydrotreating

Changes in finished lubricating oil specifications toward higher quality lubes is causing many lube plants to incorporate hydroprocessing technology into their existing conventional lube plant. These plants maintain the solvent extraction section. Effluent from the solvent extraction plant is fed to a hydrotreating unit that operates under moderate operating conditions to saturate the remaining aromatics and produce high VI and highly saturated lube. If wax production is important, the effluent from the hydrotreating unit is sent next to the solvent dewaxing unit. If wax is less important, the hydrotreated stream can be catalytically dewaxed[1] to further increase VI and yield. Hydrotreating in this case represents a more severe set of operating conditions than hydrofinishing but a less severe condition than that for lube hydrocracking. At higher pressures and with selected catalysts, aromatics

rings become saturated to become naphthenes. In addition to converting naphthenes, hydrotreating can remove most of the sulfur and nitrogen. The operating conditions are so chosen that allow for retention of selected aromatic compounds, which in turn has a positive effect on oxidation stability and deposit control. This process is particularly suitable to manufacture very high-quality stocks for turbine applications.

Various processing steps (lube hydrocracking, hydrodewaxing, or deep hydrotreating) can be combined in a number of configurations to make group I, group II, or group III base stocks. A conventional lube-making scheme comprising solvent extraction, hydrofinishing, and solvent dewaxing can make only group I lube base stocks (Fig. 10-1). Figure 10-10 shows lube hydrocracking followed by hydrodewaxing to give Groups II and III lube base stocks. Figure 10-11 shows solvent extraction

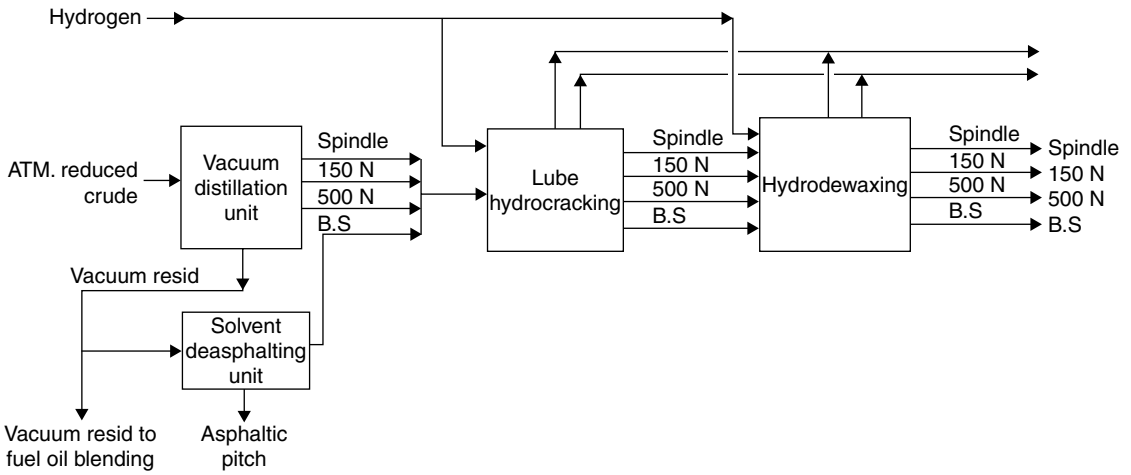


**FIGURE 10-10** Lube processing scheme.

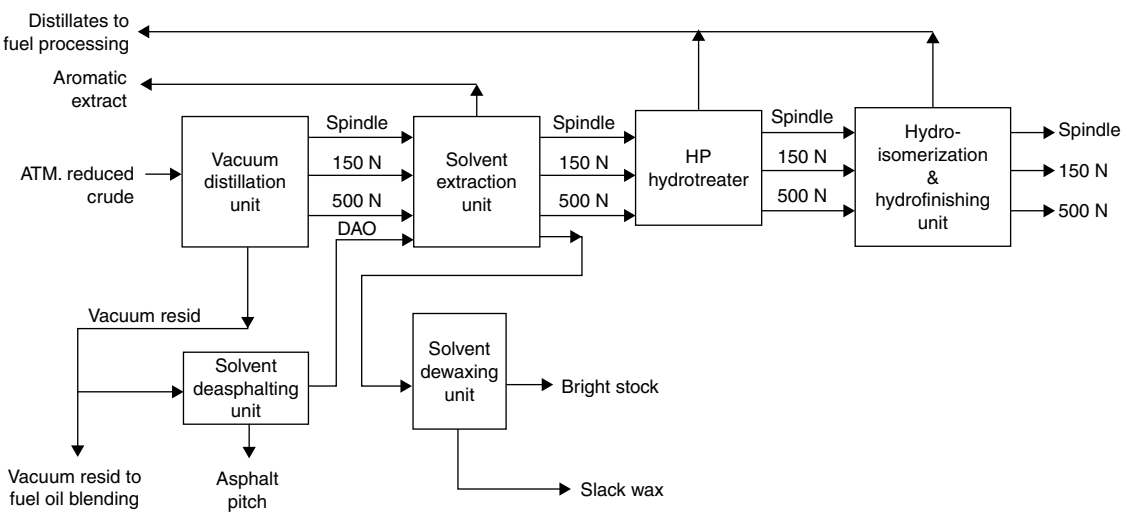


**FIGURE 10-11** Lube base stock processing scheme.

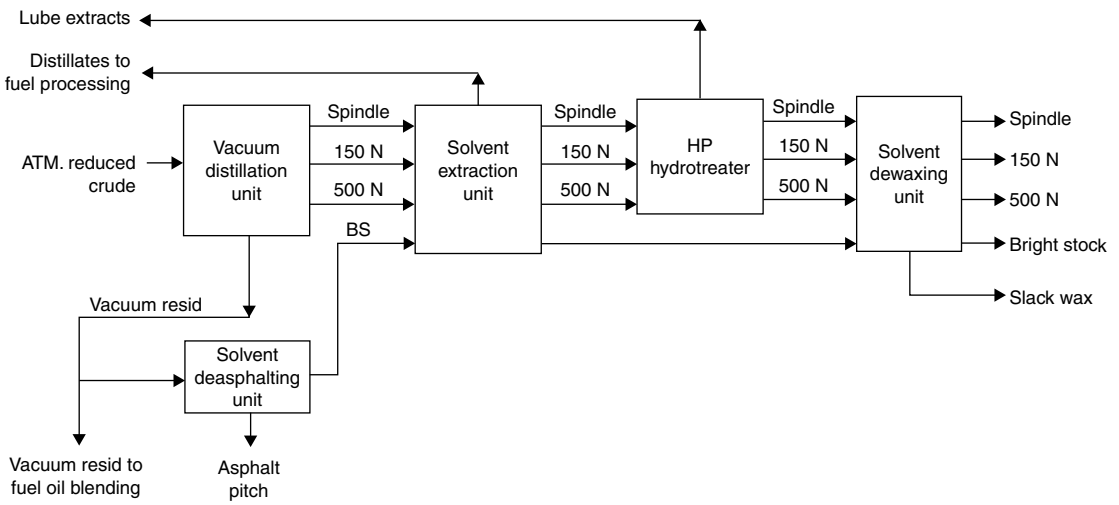


**FIGURE 10-12** Lube base stock processing scheme.

followed by deep hydrotreating and hydroisomerization unit to yield group II and group III base stocks. Figure 10-12 shows solvent extraction, deep hydrotreating, and conventional solvent dewaxing to produce high VI lube base stocks.

## *AMERICAN PETROLEUM INSTITUTE CLASSIFICATION OF BASE OILS*

Until recently, conventional lube-making schemes (solvent extraction, dewaxing, and hydrofinishing) dominated the industry. However, original engine manufacturers (OEMs) are now requiring lube base stocks for their high-performance new engines with the following properties:

- Lower viscosity for increased fuel economy
- Lower volatility for reduced oil consumption
- Improved oxidation and thermal stability for longer drain intervals
- Higher VI for improved lubricant performance at low and high temperatures

Changes in the specifications of finished motor oils are requiring an increased use of higher quality lubricants that cannot be produced by conventional lube-making schemes comprising solvent extraction, solvent dewaxing, and hydrofinishing. Lube base stocks for automobile lubricants meeting the requirements just listed can be produced by alternative hydrogen-treating processing options such as lube hydrocracking, deep hydrotreating, hydroisomerization of wax, and so on. In response to these changes, the American Petroleum Institute (API) established a base oil classification system in 1990 to help lube oil formulators to minimize retesting cost when blending licensed engine oils with base oils from different manufacturing sources. The system uses physical and chemical parameters to divide all base stocks into five groups as listed in Table 10-20.

As one moves from group I lubes to group II and group III lubes, the paraffin content of the base stock and the VI of the base stock increases. Only group I lubes can be economically produced by conventional lube technology consisting of solvent dewaxing and solvent dewaxing processes.

**TABLE 10-20** API Classification of Lube Base Stocks

| Group | Saturates, Wt % | Sulfur, Wt % | Viscosity index |
|---|---|---|---|
| I | $<90$ | $<0.03$ | $>80–<120$ |
| II | $\geqslant 90$ | $<0.03$ | $\geqslant 80–\leqslant 120$ |
| III | $\geqslant 90$ | $\leqslant 0.03$ | $\geqslant 120$ |
| IV | All poly alpha olefins | | |
| V | All base stocks not included in groups I to IV | | |

Group II and group III lubes base stocks with the very low sulfur, low volatility, and high VI required by modern automobile engines can only be produced by alternative hydroprocessing routes. It is not implied that conventional lube-making scheme is likely to be replaced by alternative processes. High-viscosity lubes can only be produced by the conventional methods. Also lubes for industrial usages, where high VI is not critical, could still be produced by a conventional lube-making method.

## REFERENCES

1. U.S. Patent 5149421, "Catalytic Dewaxing Process for Lubes Using a Combination of Silico-Alumino-Phosphate Molecular Sieve Catalyst and an Aluminosilicate Zeolite Catalyst," issued September 22, 1992.
2. U.S. Patent 4440871 "Crystalline silico alumino phosphates" issued April 3, 1984 "Crystalline silico alumino phosphates" issued April 3, 1984.