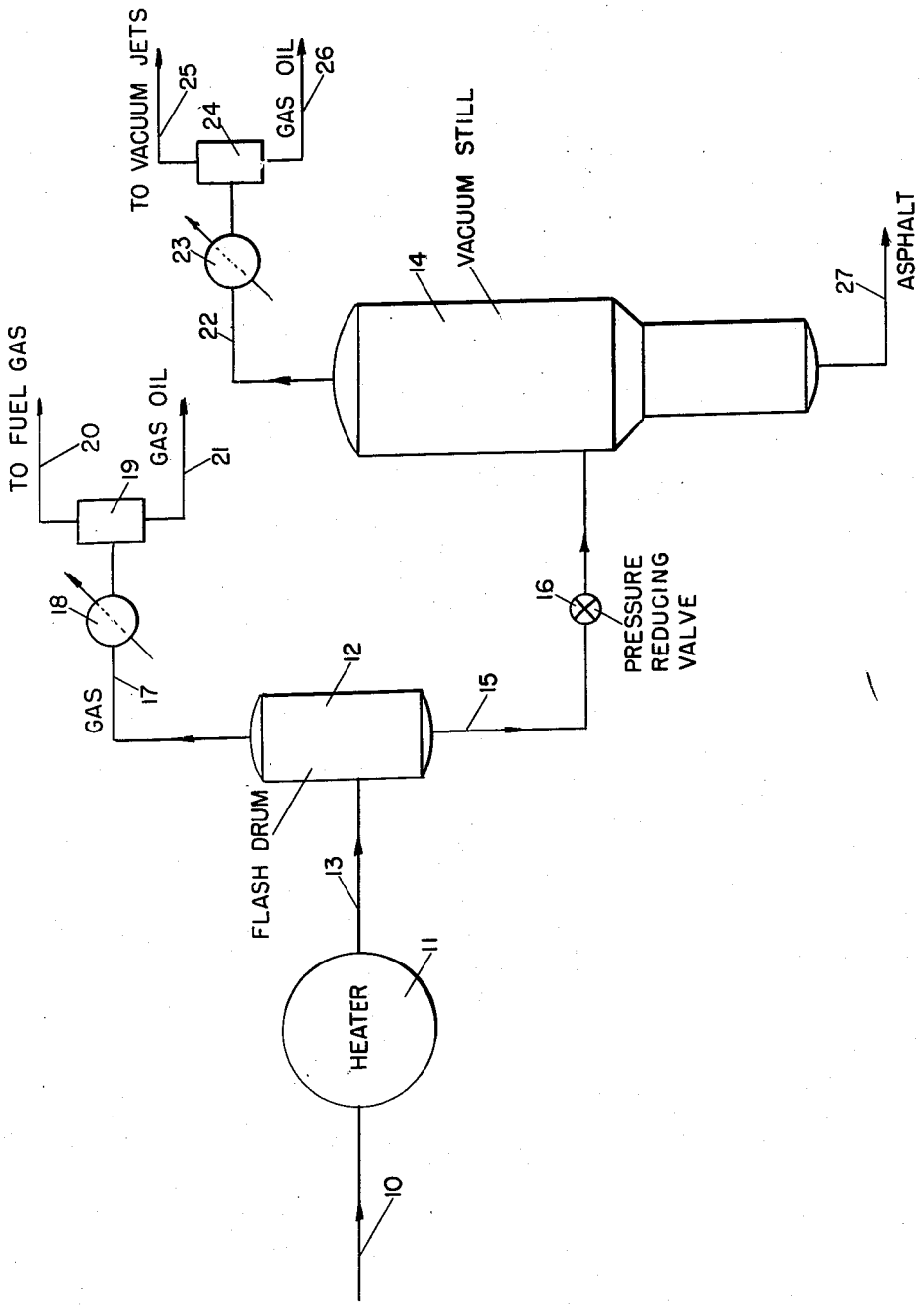
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VACUUM DISTILLATION OF REDUCED CRUDE

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INVENTOR
READING BARLOW SMITH
BY
Pennie, Edmonds, Morton & Barrows
ATTORNEYS

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VACUUM DISTILLATION OF REDUCED CRUDE

Reading Barlow Smith, Hammond, Ind., assignor to Sinclair Refining Company, New York, N. Y., a corporation of Maine

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1 Claim. (Cl. 196—77)

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My invention relates to improvements in vacuum distillation of heavy hydrocarbon stocks which are susceptible to cracking. More particularly, it is characterized by vacuum operation under higher conditions of temperature and pressure than have been before applied so as to increase vacuum still capacity, or, conversely, reduce size of vacuum equipment.

The immense demands placed upon the petroleum industry by military necessity and mechanization in all phases of modern life have magnified many problems of petroleum refining from problems of efficient and economic operation to problems of productive necessity. The wide spread introduction of so-called fluid catalytic cracking and other catalytic cracking processes has greatly increased production of high-grade gasoline and heating oils, but in turn has placed great pressure upon existing refining means for preparing cracking feed stocks. The preparation of cracking feed stocks is a ready example of the role vacuum distillation plays in refining practice in economically obtaining ultimate yields of useful products from each barrel of crude oil.

In running reduced crudes, for example, in the preparation of cracking feed stocks, the reduced crude charge is conventionally heated in a tube or coil heater and is flashed into a vacuum tower or still from which as much clean distillate or gas oil as possible is taken overhead. In conventional practice it is commonly accepted that the flash temperature and consequently the heater temperature is limited by the temperatures at which incipient cracking occurs; that is, about 700° to 800° F. As is well-known, cracking makes gas, and even small quantities of fixed gases in the vacuum zone seriously interfere with the efficiency of the vacuum operation. For the maintenance of subatmospheric pressures in the vacuum still requires in practical installations the use of vacuum jets which depend for efficiency upon substantially complete condensation of the vapor throughput. Thus the presence of small amounts of fixed gases sharply reduces jet efficiency and requires such large and costly jets that operation quickly becomes impractical. Hence, in conventional practice it is customary to set a flash temperature, which when carried back through the system to the heating zone, will avoid cracking in the system, and then reduce the pressure in the vacuum flash tower to the point necessary for the desired overhead yield. Obviously, the demand for high percentage overhead yields results in flash towers which are disproportionate in size to the throughput.

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Moreover, certain well-known phenomena of cracking further restrict flexibility of operation in conventional practice. In the past considerable difficulty with coking at the point of initial vaporization in the heater has been experienced. This difficulty is generally avoided by holding pressure on the heater coil, thereby preventing vaporization. But, since vaporization is prevented, the temperature runs considerably higher because all of the heat input goes to sensible heat. Furthermore, since the feed is maintained in the liquid phase, the residence time in the heater is increased. The combination of higher temperatures and increased residence time of course promotes cracking, and accordingly limits the benefits of liquid phase operation in conventional practice.

I have found, however, that if I depart from conventional practice and deliberately establish a higher pressure and temperature in the system, I can significantly increase vacuum still capacity or significantly reduce vacuum still size by removing fixed gases made in cracking prior to the vacuum step. I thus avoid interference with the efficiency of the vacuum jets while running under conditions of higher throughput, and provide for improved flexibility of operation with respect to pressure and phase, temperature and percentage overhead yields.

Where conventional practice is limited to flash temperatures in the neighborhood of 650° to 750° F. and requires a low vacuum still pressure within the range of 20 to 75 mm., I maintain a pressure within the approximate range of 150 to 350 mm. of mercury, preferably say about 200 mm., and a temperature within the approximate range of 750° to 850° F., preferably say about 830° F. I find that my improved system of operating results in a reduction in the diameter of the flash still of from one-quarter to one-half or more. To effect such results, I prefer to operate at a heater temperature of about 900° F. and to interpose a relatively small flash drum between the vacuum still inlet and the heater outlet. The pressure on the drum is usually about 90 p. s. i. a., accounting for pressure drop through the heater coil, and the temperature is a few degrees below that of the heater outlet.

I find that my improved system is particularly valuable in running reduced crudes to ultimate gas oil yields in cracking stock preparation for it greatly increases the capacity of existing equipment. Thus, it significantly relieves the pressure on refining capacity at a time of crude short-

age and gasoline and heating oil shortages. Conversely, it permits reduction in the size of equipment necessary for the same production, and permits the use of vacuum jets in the most efficient size range and at the most economical cost. Since vacuum jet operation is necessarily inflexible, my improved scheme of vacuum distillation introduces a certain degree of flexibility into a previously inflexible process. I have also found that my improved method eliminates operating difficulties encountered by pushing capacity under the conventional scheme of operation. For high pressure-high temperature vacuum operation not only brings the system into size, but reduces entrainment of asphaltic constituents and clears up the color of the gas oil overhead.

An embodiment of my invention is further illustrated in the drawing and the following illustrative example, neither of which is offered, however, in a limiting sense.

In the drawing, charge oil is introduced to the system through line 10 and is charged to heater 11 at a pressure that will maintain 90 p. s. i. a. in flash drum 12. The heated oil enters flash drum 12 through line 13, and freed from light gases passes to vacuum still 14 through line 15 and pressure reducing valve 16. Light gases are taken overhead by means of line 17, and are passed through condenser 18 to receiving drum 19 from whence fixed gases are released through line 20 to refinery fuel or other purposes. Light gas oil is collected and pumped to the cracking process through line 21. In vacuum still 14, gas oil is flashed overhead at 200 mm. of mercury and a temperature of 830° F. through line 22, condenser 23 and receiving drum 24. Line 25 leads to a conventional jet condenser system (not shown) and line 26 serves to remove gas oil to the cracking process. Asphalt bottoms are removed through line 27.

The operation of my improved system is perhaps best illustrated by comparison with conventional operation on a unit running 510 barrels per hour of 20° API feed stock. The unit is run to gas oil and asphalt of the following characteristics:

Gas oil		*Asphalt butts*	
Gravity, °API	23.1	Gravity, °API	9.9
Bbls./hr	410	Bbls./hr	100
Aniline pt	187	Penetration	211
Carbon res	0.21		

A comparison of the feed and overhead is seen in the following tests:

Feed, Vacuum Dist. at 5 mm. Hg		Overhead at 760 mm. Hg
	°F.	°F.
IBP	490	593
10	580	640
20	635	669
30	675	694
40	730	718
50	760	735
60	800	750
70	840	759

The above rates and product qualities are produced by operation at a flash temperature of 705° F. and a flash pressure of 36 mm. of mercury. The tower top temperature is 665° F. and the heater effluent is 748° F. The diameter of the tower necessary for this operation, which is pushing maximum capacity, is 16½ feet. When changed over to my improved method by utilizing a flash drum of 4 feet by 10 feet intermediate the heater outlet and the vacuum still, a vacuum tower diameter of only 11 feet is required. The rate and product qualities are the same, but the flash temperature is raised to 830° F., the tower top temperature is 800° F., and the flash pressure is 200 mm. of mercury. The heater effluent is 905° F. and the pressure 90 p. s. i. a. for liquid phase heating. At lower pressures which permit vaporization, the heater effluent is about 870° to 880° F. Cracking occurs under these conditions although in terms of cracking still operation, it is relatively small. However, in terms of the size of jets on the vacuum tower, the amount of gas produced would require such abnormally large jets that the operation would not be feasible. In the described example, however, removal of the gas facilitates the illustrative high temperature and pressure operation.

I claim:

In vacuum distillation of reduced crude oil stocks to prepare gas oil cracking stocks, the steps of heating a reduced crude stock in the liquid phase under superatmospheric pressure to a temperature of the order of about 900° F. to effect incipient cracking, passing the treated stock to a relatively small flash zone and removing substantially all fixed gases while maintaining the stock substantially in the liquid phase under superatmospheric pressure, flashing the preheated and degassed stock in an enlarged vacuum zone at about 750° to 850° F. under a pressure within the approximate range of 150 to 350 mm. of mercury, removing vaporized gas oil overhead from the vacuum zone and removing unvaporized bottoms from the vacuum zone.

READING BARLOW SMITH.

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