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NATURAL GAS LIQUEFACTION AND SEPARATION

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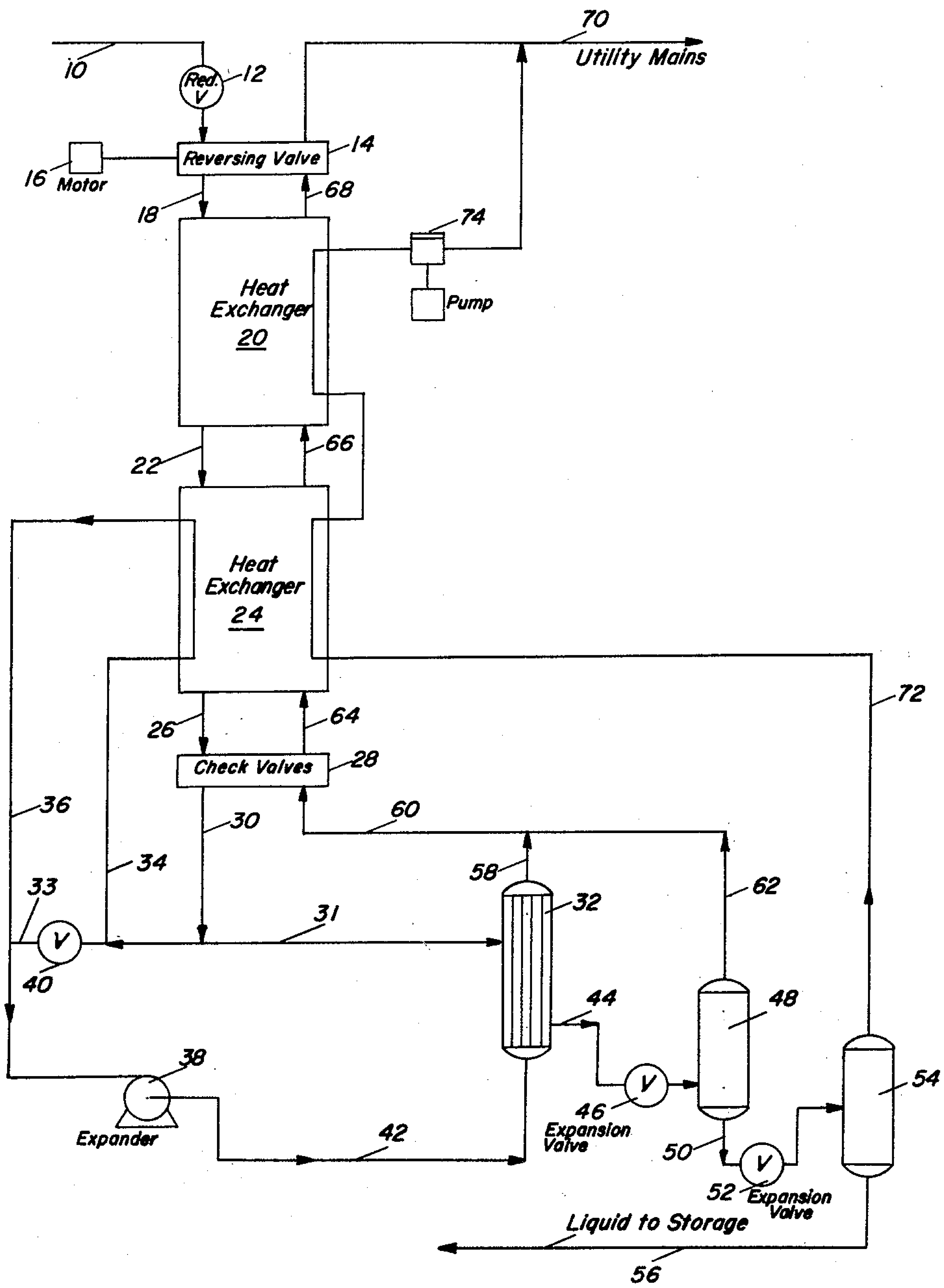

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NATURAL GAS LIQUEFACTION AND SEPARATION

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This invention relates to the distribution of natural gas and more particularly to the recovery of a portion of the gas as liquid by the appropriate conversion of the inherent energy of the gas in the pipeline.

It is well known that the transmission systems for natural gas customarily used in the transcontinental pipeline deliver the gas at the industrial terminals at pressures from as little as 200 p.s.i.g. to pressures in excess of 400 p.s.i.g. It is usually necessary, however, for domestic and industrial purposes to reduce the pressure to a distribution grid to a maximum pressure of about 50 p.s.i.g. Heretofore, this has been accomplished by expansion without any useful work or effect and thus with a definite loss of the energy potential. While this energy level is below that which can be economically recovered by the work of an expansion engine, nevertheless, there is a substantial value available if it can be utilized.

In accordance with my invention, I propose to recover a small percentage of the gas in liquid form by the use of this potential energy and to store such liquid for supplementary energy during periods of excessive gas requirements.

More particularly, by use of suitable heat exchangers, expansion engines and process flows, without the addition of more than nominal amounts of power, I find that it is possible to obtain a net condensate from typical pipeline natural gas which equals in heating value, molecular weight, and combustion characteristics the original pipeline gas, so that the liquefaction of a fraction of the gas can be carried out without degradation of the unit value of the original pipeline gas and the net condensate can be used without modification when there is a subsequent requirement for increased quantities of pipeline gas.

Although it is recognized that the percentage of gas that can be liquefied by this system is small, it is recognized that the energy is available all of the time and the recovery of amounts in the order of 4 to 8% over a year represent a very large added source of energy that can be used during the severe peak loads.

Further objects and advantages of my invention will appear from the following description of a preferred form of embodiment thereof taken in connection with the attached drawing which is a schematic flow diagram of a gas liquefaction cycle in accordance with my invention.

As pointed out, the natural gas from a transcontinental pipeline which enters the system in the line **10** is normally at a pressure of the order of 200 to 400 p.s.i.g. and generally at ambient temperature, as for example, 100° F. For my purposes this gas is passed through a reducing valve **12** to provide a control and to reduce the gas to about 200 p.s.i.g. as it enters the reversing valve **14**. This valve is operated through the motor **16** in a suitable cycle.

The gas in line **18** is thus at about 95° F. as it enters heat exchanger **20**. The gas then passes in contact with a cool surface hereinafter described so that the gas discharging from the heat exchanger through line **22** is at about −40° F. This gas then passes through the second heat exchanger **24** and is again in contact with a cold surface so that the gas is further reduced in temperature and discharges at **26** at about −178° F. The gas then passes through check valve **28** and the gas discharging in line **30** passes in part through line **31** to the methane liquefier **32**, and in part through valve **40** and line **33** to expander

38. A part of this gas passes through line **34** and heat exchanger **24** and thence through line **36** to the expander **38**. This by-pass line **34** serves to control the temperature in the heat exchanger **24** and serves as the unbalance stream characteristic of reversing exchangers for control of temperature differences for the reversible condensation and revaporization of less volatile constituents.

The position of the reversing valves **16** is alternated periodically, as for example every three to sixty minutes, so that the pipeline gas passes alternately through one side and then the other side of heat exchangers **20** and **24**. By this means water, butane, and propane are condensed from the feed gas on one side of heat exchanger **20** in the first part of the cycle, and revaporized into lower pressure gas from exchanger **24** in the second part of the cycle, while additional water, butane and propane are condensed from pipeline gas on the second side of exchanger. Similarly, carbon dioxide and ethane are condensed and revaporized in heat exchanger **24**, using low pressure gas from line **60** (which see below) to receive the vaporized material. The design and attitude of heat exchangers **20** and **24** are such that condensed liquid cannot migrate longitudinally and disturb the temperature difference pattern essential to the complete revaporization of material. By this means, a gas essentially containing methane is available at line **30** to be used in expander **38** or methane liquefier **32**.

The amount of gas which passes through the expander **38** is usually in excess of 90% of the total flow from heat exchanger **24** and as the pressure drops from about 200 p.s.i.g. to 50 p.s.i.g., the temperature of the gas will drop from −148° F. to about −220° F. The expanded gas in line **42** then passes through the methane liquefier **32** in countercurrent contact with the gas entering from the line **31** under such circumstances that the minor portion of the gas in line **31** is sub-cooled and partially liquefied. The cooled gas and liquid discharge from methane liquefier **32** through line **44** and the gas is further expanded in valve **46** into the flash separator **48**. The liquid removed from the bottom of flash separator **48** through line **50** is again expanded in valve **52** into the low pressure flash separator **54**. From 5 to 10% of the total gas may be removed through line **56** as liquid and passed to storage from which it is available for use in peak seasons. The low pressure gas in line **72** may then serve as an unbalance or heating stream for heat exchangers **24** and **20**.

The partially heated gas at about 50 p.s.i.g. from the methane liquefier **32** passes upwardly through the line **58** and line **60** to the check valve **28** with its temperature raised to about −190° F. In a similar manner, the gas from the high pressure separator **48** is removed through the line **62** and joins with the gas line **60**. This gas at 45–50 p.s.i.g. is used to revaporize condensed water, butane, propane, ethane and carbon dioxide in the reversing cycle of heat exchangers **20** and **24** as described above.

The combined high pressure gas then passes upwardly through the line **64** through heat exchanger **24** and through line **66** into heat exchanger **20** and thence through line **68** and reversing valve **14** to the utility gas main line **70**, which in the present case, is assumed to be at 40 p.s.i.g.

The low pressure gas from the low pressure separator **54** passes upward from the line **72** and through heat exchangers **24** and **20** and is then compressed at **74** before joining the utility main line **70**.

To some extent, depending upon the efficiency of the heat exchangers, it is found that, by the cycle disclosed, pipeline gas at 200 p.s.i.g., can be converted to utility main gas at 50 p.s.i.g. with approximately 6% of the gas flow liquefied. If the initial gas is at 400 p.s.i.g. approximately 8% of the gas can be liquefied without external energy.

Reference has been made to the condensation and revaporization of the higher hydrocarbons and if it is undesirable to retain them in the system they can be removed as usually the ethane, propane and butane command preminimum prices relative to natural gas.

With the foregoing cycle there is no degradation of the gas being processed and the amount of liquid which is removed is ultimately used without any utility or any other loss. As mentioned, the unit is self-sustaining and it is merely a question of recovering more or less of the liquid depending upon pressure and heat exchanger efficiency.

While I have shown a preferred form of embodiment of my invention, I am aware that modifications thereof can be made within the scope and spirit of the specification herein and of the claims appended hereinafter and I, therefore, desire a broad interpretation of such specification and claims.

I claim:

1. A method of operating a plant for converting a high pressure natural gas predominantly methane supplied by a transmission line to a substantially lower pressure heating gas and for distributing said heating gas to consumers with a minimum of interruption, which comprises: (1) withdrawing said natural gas under transmission line pressure from the transmission line by which said gas is supplied to said plant; (2) cooling said gas in a reversing heat exchanger while under substantially transmission line pressure to form a gaseous fraction; (3) removing a part of the cooled gaseous fraction and returning it as an unexpanded unbalance stream to the exchanger; (4) expanding said part of said gaseous fraction to a pressure substantially corresponding to the distribution line pressure maintained by said gas plant; (5) passing said cold expanded gaseous fraction in heat exchange relationship with the balance of said natural gas to liquefy a part thereof at the high pressure; (6) substantially continuously, at least during periods of ample supply of said natural gas, conducting said liquid fraction in the liquid state to at least one storage vessel; and (7) passing the expanded gaseous fraction after the heat exchange of step 5 in heat exchange relation to the first mentioned gas to accomplish the cooling of step (2).

2. The method of operating a plant as claimed in claim 1 wherein the liquid removed is of substantially the same thermal value as the distribution line gas and is adapted to be reintroduced as vapor into said distribution line gas.

3. The method of claim 1 wherein the pipeline natural gas contains impurities of the class of water, butane, propane, ethane and CO_2 and a portion of the high pressure low temperature liquid is flashed in part and the vapor is conducted through at least one of the heat exchange steps to unbalance the temperature therein and revaporize impurities formed on the heat exchange surfaces.

4. The method of claim 1 wherein the pipeline natural gas contains impurities of the class of water, butane, propane, ethane and CO_2 and the heat exchange step is carried out in reversing exchangers and a portion of the gas discharged therefrom is separately returned through one of the heat exchangers for unbalance and to condition the gas for entry into the expansion step and a portion of the high pressure low temperature liquid is flashed in part and the vapor is conducted through at least another part of the heat exchangers to unbalance the temperature therein and revaporize impurities formed on the heat exchange surfaces.

5. The method of operating a plant as claimed in claim 1 wherein the cooling of the gas is to a temperature in the order of below −200° F. at the high pressure, the expansion of step (4) is accomplished in doing work, and the part of the gas liquefied is in the order of about 4 to 8%.

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